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DESIGN OF CONTROL CONFIGURATION FOR HIGHLY HEAT-INTEGRATED HDA PROCESS

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ชัยวัฒน์ ชูลิวรรณพี: การออกแบบโครงสร้างควบคุมสำหรับกระบวนการไฮโดรดีอัลคิลเลชันที่พลังงานเบ็ดเสร็จอย่างสูง. (DESIGN OF CONTROL CONFIGURATION FOR HIGHLY HEAT-INTEGRATED HDA PROCESS) อ. ที่ปรึกษา: ผศ. ดร. มนตรี วงศ์ศรี, อ. ที่ปรึกษาร่วม: ผศ. ดร. กุลชนาฐ ประเสริฐสิทธ์, 218 หน้า.

การออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์สำหรับกระบวนการที่มีการเบ็ดเสร็จของพลังงานสูงเป็นงานที่ค่อนข้างยาก เนื่องจากกระบวนการที่มีการเบ็ดเสร็จของพลังงานสูงนั้น มีหน่วยยูทิลิตี(Utility Unit) เช่น เครื่องทำความเย็น และเครื่องทำความร้อน ในจำนวนที่น้อยเพื่อที่จะใช้ในการดูสับภาระความแปรปรวนที่เกี่ยวกับพลังงาน ปัญหานี้สามารถแก้ได้โดยการเพิ่มหน่วยยูทิลิตีเสริม(Auxiliary Utility Unit) แต่อย่างไรก็ตามการเพิ่มหน่วยยูทิลิตีเสริมจะเป็นการเพิ่มค่าใช้จ่ายในส่วนของการก่อสร้าง การดำเนินการและการซ่อมบำรุง ในงานวิจัยนี้จึงได้เสนอวิธีการในการออกแบบเพื่อรับประกันการทำงานของกระบวนการที่มีการเบ็ดเสร็จของพลังงานสูงโดยใช้จำนวนหน่วยยูทิลิตีเสริมให้น้อยที่สุด ซึ่งจะนำไปประยุกต์ใช้กับกระบวนการไฮโดรดีอัลคิลเลชันที่มีการเบ็ดเสร็จของพลังงานแตกต่างกัน โดยเริ่มต้นด้วยการกำหนดรายละเอียดและขนาดของตัวรบกวน ตามด้วยออกแบบสภาวะการดำเนินงานที่แย่ที่สุด หลังจากนั้นทำการออกแบบเส้นทางเดินของความร้อนและสุดท้ายจำนวนหน่วยยูทิลิตีเสริมที่น้อยที่สุดจะถูกประเมิน จากวิธีการข้างต้นเราสามารถแก้ปัญหาเกี่ยวกับความยากในการควบคุมของกระบวนการไฮโดรดีอัลคิลเลชันได้โดยการเพิ่มจำนวนหน่วยยูทิลิตีเสริมในกระบวนการเพียงหนึ่งตัว ในขณะที่ลูเบนได้ทำการเพิ่มจำนวนหน่วยยูทิลิตีเสริมถึง 3 และ 4 ตัวตามความซับซ้อนของการเบ็ดเสร็จพลังงานเพื่อที่จะแก้ปัญหาเดียวกันนี้ นอกจากนี้ในงานวิจัยนี้ยังได้ทำการเสนอโครงสร้างการควบคุมแบบใหม่ขึ้นมาอีก 3 โครงสร้างอีกทั้งยังประเมินสมรรถนะของโครงสร้างดังกล่าวไว้ด้วย ผลการจำลองทางพลวัตพบว่า สมรรถนะในการควบคุมสำหรับกระบวนการที่มีการเบ็ดเสร็จของพลังงานสูงโดยใช้จำนวนหน่วยยูทิลิตีเสริมน้อยที่สุดให้สมรรถนะเหมือนกันกับกระบวนการที่มีการเบ็ดเสร็จของพลังงานสูงแต่ใช้หน่วยยูทิลิตีเสริมเต็มที่ โดยโครงสร้างการควบคุมแบบที่ 1 เป็นโครงสร้างที่สามารถรับมือกับภาระของตัวรบกวนได้ดีที่สุดเนื่องจากสมรรถนะการควบคุมของโครงสร้างดังกล่าวดีกว่าโครงสร้างอื่นๆ ซึ่งในโครงสร้างแบบที่ 1 ได้ทำการติดตัวควบคุมแบบซีเล็คทีฟซึ่งมีตัวสลับแบบโลว์ซีเล็คเตอร์ เพื่อที่จะนำพลังงานกลับคืนสูงสุดทางพลวัต ซึ่งพบว่าสามารถประหยัดพลังงานไป 0.45 % สำหรับการทดสอบการเปลี่ยนแปลงภาระความแปรปรวนของพลังงานในกระแสเย็น และนอกจากนี้อุณหภูมิทางเข้าในกระแสร้อนของหม้อต้มซ้ำจะถูกควบคุมให้คงที่เพื่อที่จะป้องกันการกระจายของภาระความแปรปรวนของพลังงานไปยังหน่วยแยก

ภาควิชา........วิศวกรรมเคมี........	ลายมือชื่อนิสิต..
สาขาวิชา......วิศวกรรมเคมี........	ลายมือชื่ออาจารย์ที่ปรึกษา..................................
ปีการศึกษา.........2550............	ลายมือชื่ออาจารย์ที่ปรึกษาร่วม..................................

4770572921 : MAJOR CHEMICAL ENGINEERING

KEY WORD : HEAT HDA PROCESS/PLANTWIDE PROCESS CONTROL/HEAT EXCHANGER NETWORK

CHAIWAT CHULIWANLEE: DESIGN OF CONTROL CONFIGURATION FOR HIGHLY HEAT-INTEGRATED HDA PROCESS. THESIS ADVISOR: ASST. PROF. MONTREE WONGSRI, D.Sc. THESIS CO-ADVISOR: ASST. PROF.KULCHANAT PRASERTSIT, Ph.D., 218 pp.

The design plantwide control structures for a highly heat integrated plant is quite difficult task since the highly heat integrated plant has a few utility unit (i.e. heater and cooler) to absorb thermal disturbance load. This problem can be solved by adding auxiliary utility unit. However, more auxiliary utility unit will be increase capital operating and maintenance costs. In this research, we propose the strategy to design the workable highly heat integrated plant like alternatives 5 and 6 of hydrodealkylation of toluene (HDA) process with minimum auxiliary utility unit. It starts with specifying the disturbances and their magnitudes, next designing the worst case condition, and then designing the heat pathway. Finally, the minimum auxiliary utility units are evaluated. We can solve the control difficulties associated with alternatives 5 and 6 by adding an auxiliary utility unit to the process instead of three and four as suggested by Luyben (1999), respectively. The three new control structures are proposed and their performances are evaluated. As shown in dynamic simulation study, the control performance for the highly heat integrated plant with minimum auxiliary utility unit is same with the highly heat integrated plant with full auxiliary utility units. CS1 is the best control structure for handle disturbances due to it gives better control performances. In this control structure, the selective controller with low selector switch (LSS) is employed to achieve dynamic maximum energy recovery that can save energy about 0.45 % for change in the heat load disturbance of cold stream. Besides, the inlet hot temperature at entrance of reboiler is maintained to prevent the propagation of thermal disturbance to separation section.

Department……Chemical Engineering…	Student's signature……………
Field of study…Chemical Engineering…	Advisor's signature………
Academic year…….2007……………...	Co-advisor's signature………………

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Finally, I would like to dedicate this dissertation to my family for their edification, support and endless love. Without them, I would not have been the person I am and certainly this work would never have reached completion.

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CHAPTER I

INTRODUCTION

This chapter introduces the importance and reasons for research, research objectives, scope of research, contribution of Research**,** procedure and method and the research framework.

1.1 Background

Today's chemical industry has become even more competitive as companies try to improve profit and reduce production times to shorten the supply chain to the customer. Part of this competitiveness has required the complexity of the processes to increase through attempts to recover unreacted material, reduce pollution, and capture energy traditionally lost to the surroundings. The increased complexity has also served to increase the interactions among the process units by allowing back-propagation of disturbances that traditionally would have exited the plant. Because of this integration, the potential exists that disturbances might stay within the process and that their effects might amplify, causing severe stability and controllability concerns. Individual control systems are designed for each unit operation or piece of equipment in a plant, after which any conflicts between control loops are reconciled. Plantwide control, an approach that guides control system design for an entire plant is considered to design control structure for complex plant.

In hydrodealkylation (HDA) process of toluene to benzene that consists of a reactor, furnace, vapor-liquid separator, recycle compressor, heat exchangers and distillations. This plant is a realistically complex chemical process so this process will be used to study and design process with plantwide process control theory.

Terrill and Douglas (1987) developed six heat exchanger network (HEN) alternatives for a base case design for HDA process which energy saving 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 designs is alternative 6, recovers 43 percent of the base case net energy consumption and use number of utility less than the other alternatives. Consequently, they evaluated total annual cost of each alternative. Their study showed that the total annual cost of alternatives 5 and 6 are less than the other alternatives. However complex heat-integration makes the plants more difficult to control. Luyben (1999) suggested that we can solve certain control difficulties associated complex heat-integration (i.e. alternative 5 and 6) by adding extra equipment as auxiliary utility coolers and reboilers to the process. However adding extra equipment will be increase capital, operating and maintenance cost. Therefore, the strategy to design the workable highly heat integrated plant like alternatives 5 and 6 of hydrodealkylation of toluene (HDA) process with minimum auxiliary utility unit is proposed in this research. Additional, we present three new plantwide control structures for highly heat integrated HDA process alternative 5 and 6.

1.2 Research objectives

Objectives of this study are listed below

1. Design to guarantee the workable highly heat integrated HDA process alternatives 5 and 6 with minimum auxiliary utility unit.
2. Design the new plantwide control structures for highly heat integrated HDA process alternatives 5 and 6 with minimum auxiliary utility unit.

1.3 Scopes of research

Scopes of this research are listed below

1. Simulation of the hydrodealkylation (HDA) of toluene process alternatives 5 and 6 are performed by using a commercial process simulator – HYSYS.
2. Description and data of hydrodealkylation (HDA) process is obtained from Douglas, J. M. (1988), William L. Luyben, Bjorn D. Tyreus, and Michael L. Luyben (1998), and William L. Luyben (2002). And the energy integrated hydrodealkylation (HDA) process is obtained from Terrill and Douglas 1987.
3. Heuristic design is used for the design of plantwide control structure

1.4 Contribution of Research

The contributions of this work are as follows:

1. The strategy to design the workable highly heat integrated plant with minimum auxiliary utility unit is proposed.
2. The equipment cost of process for highly heat integrated hydrodealkylation (HDA) of toluene process alternatives 5 and 6 are decreased.
3. The new control structures for highly heat integrated hydrodealkylation (HDA) of toluene process alternatives 5 and 6 with minimum auxiliary utility units are designed.

1.5 Procedures Plan

The procedures of this research are as follows:

1. Study plantwide process control theory
2. Study HDA process and concerned information
3. Steady state simulation of HDA process highly complex energy integration
   - 3.1. Determine the operating condition for HEN based on the input disturbance loads
   - 3.2. Determine design conditions for HEN based on the input disturbance loads (worst case condition is designed)
   - 3.3. Determine appropriate heat pathway at worst case condition
   - 3.4. Check the workable HDA process in HEN at worst case conditions based on the input disturbance loads
4. Design new control structure of the HDA process with highly complex energy integration.
5. Simulate the dynamic of HDA process with control structures design.
6. Evaluate the dynamic performance of the designed control structures based on the input disturbance loads.
7. Analyze the design and simulation results.
8. Conclude the thesis.

1.6 Research Framework

This thesis has been divided into seven chapters.

In Chapter I, the background, objectives, scopes, contributions and research planning of this research is introduced in this chapter.

In Chapter II, a review of the previous work on heat exchanger networks (HENs) design and plantwide process control design are given.

In Chapter III, background information of plantwide control, plantwide control design procedure, plantwide energy management and heat exchanger network are presented.

In Chapter IV, the description of the HDA process and the steady state simulation of highly heat integrated HDA process alternatives 5 and 6 using HYSYS simulator are presented.

In Chapter V, the strategy to design the workable highly heat integrated plant with minimum auxiliary utility units is proposed. Additional, this strategy is also applied to the highly heat integrated HDA process alternative 5 and 6.

In Chapter VI, The three new plantwide control structures and dynamic simulation for the highly heat integrated HDA process alternatives 5 and 6 are present. Beside the effect of selective control with low selector switch (LSS) to energy management are proposed.

The overall conclusions and recommendations of this thesis are discussed in Chapter VII.

CHAPTER II

LITERATURE REVIEW

This chapter is to present a review of the previous work on plantwide control and heat exchanger network (HEN)

2.1 A Hierarchical Approach to 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 proceeded to the later decision level. Each decision level terminates in an economic analysis. Experience indicates that less than one percent of the ideas 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 (i.e. levels 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 them to define the controlled variables.

2.2 Plantwide Control

Plantwide control involved the systems and strategies required to control an entire chemical plant.

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 plant-wide control and multivariable control problems. Luyben et al. (1997) constructed nine steps of the proposed procedure center around the fundamental principles of planwide control: energy management; production rate; product quality; operational, environmental and safety constrain; liquid-level and gas-pressure inventories; makeup of reactants; component balances; and economic or process optimization.

After that Skogestad and Larsson (1998) considered “which variables should be controlled, which variables should be measured, which inputs should be manipulated, and which links should be made between them?” These are the question that plantwide control tries to answer. There are two main approaches to the problem, a mathematically oriented approach (control structure design) and a process oriented approach. It was shown that the idea of "self-optimizing control" provides a link between steady-state optimization and control.

Vasbinder and Hoo (2003) described a modular decomposition of plant flowsheets that is assessed according to a decision-based approach, the analytical hierarchical process, that consistently evaluates the merits of the decomposition among alternatives and between neighboring modules to arrive at a prioritization of objectives for each module and for the integrated plant.

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

For control structure designed Luyben et al. (1997) analyzed the effect of the process designed on control structure for a system with a reactor, two distillation columns, and two recycles streams. The reaction A +B $\rightarrow$ C occurred in a reactor, the two distillation columns recycled components A and B back to the reactor. This work modified control structure for this process by no reactor composition measurement was used, and throughput was directly fixed by flow controlling the fresh feed. For analyze in the dynamic and steady state limitations, it had no means to recover from disturbances that cause the two reactant compositions to cross. Change in reactor holdup and recycle flow rates away from their values in the economically optimal design improves the ability of this control strategy to handle large disturbances. After that Luyben (2000) studied process which had exothermic, irreversible, gas phase reaction A + B $\rightarrow$ C occurring in an adiabatic tubular reactor. A gas recycle returned unconverted reactants from the separation section. Four alternative plantwide control structures for achieving reactor exit temperature control were explored. 1 the set point of the reactor inlet temperature controller was changed (CS1), 2 the recycle flow rate was changed, 3 the flow rate of one of the reactant fresh feeds was changed (CS3) and 4 used an "on-demand" structure. Manipulation of reactor inlet temperature appeared to be the last attractive scheme. Manipulation of recycle flow rate given the best control but may be undesirable in some system because of compressor limitations. The on demand structure provided effective control in the face of feed composition disturbances. And then Luyben (2000) considered the design and control aspects of a ternary system with the gas phase reversible, exothermic reaction A + B ↔ C occurring in an adiabatic tubular reactor packed with solid catalyst. He designed different control structure by fresh feed control pressure. The reactor inlet temperature is fixed. The recycle flowrate is used to indirectly set the production rate (CS1). By pressure was controlled by recycle flowrate and the production rate was directly set by the fresh feed flowrate (CS2). Given a control structure where the recycle flowrate was fixed (CS3). If process had inert, the additional control loop added is the control of composition of the inert component in recycle and purge gas. Effective control was obtained in the face of quite large disturbances. In 2003 Larsson et al. considered control structure selection for a simple plant with a liquid phase reactor, a distillation column, and recycle of unreacted reactants. To optimize economics, they needed to control active constraints. For the case of both minimizing operating costs (case1) and maximizing production rate (case2), it is optimal to kept the reactor holdup at its maximum. For the unconstrained variable, they looked for self-optimizing variables where constant set points gave acceptable economic loss. To avoid the snowball effect, it had been proposed to fix a flow in a liquid recycle loop. The limitation is that it can handle small feed rate change or large variation in the reactor holdup.

Designed a process control structure for complex process was a complicate task. The designed control loop would effect the operation significantly. Poothanakul (2002) used plantwide control strategies for designed control structures of butane isomerization process to achieved impurity of normal butane in product and desired production rate. First control structure was controlled quality of product by product flow, second control structure concerned about reduction of effected of recycle by controlled temperature inside the distillation which could be controlled by adjusted outlet flow of distillate of column. And the last wanted to reduce effect of recycle indirectly by controlled temperature inside the distillation with outlet flow of bottom. For other process Hydrodealkylation process of toluene Sayfon (2002) designed 3 control structures for reduced effect from disturbances that caused production rate change. The first control scheme measured toluene flow rate in the process and adjusted the fresh toluene feed rate accordingly. The second was modified from the first scheme by added a cooling unit to controlled the outlet temperature from the reactor. In the third scheme, a ratio was introduced to the second control scheme for controlling the ratio of hydrogen and toluene within the process. These six control structures was compared with reference on plantwide process control book, Luyben 1998, the result was performance of these structure higher than reference. After that Tangsombutvisit (2004) developed rigorous model for the hydrodealkylation of toluene (HDA) process by using the commercial software, HYSYS.PLANT. The case of HDA process, The two control structures designed by Kietawarin (2002) are considered. The steady-state analysis is confirmed that the second control structure should be controlled. For using the controllability analysis it appeared that the problems mainly come from the interaction between the different units in the flowsheet. Controllability analysis described the control structure2 can give the result into satisfied bound. That means the effect of changing setpoint is less than the first one. However, the control structure1 can reject the disturbance better than the second one. In the same year Kasemchainun (2004) applied plantwide control strategy for designing control structures of a Vinyl Acetate monomer plant. Three alternative plantwide control structures was designed, tested and compared the performance with Luyben’s structure. For the result, the first control structure was designed control structure by used of the fresh acetic fed to manipulated the total acetic feed in vaporizer and controlled the water composition in overhead column. In the azeotrope column was high boilup ratio so the designed control structure II was designed that modifying from the first in column temperature loop. This scheme measured the tray temperature and adjusted the bottom flowrate to control the vinyl acetate composition and the level was controlled by changing the reboiler heat input. The last structure used when the reactant comes from upstream unit. The production rate was set by changing the fresh ethylene feed. All of controlled structure achieved a good controllability. In the same year Thaicharoen (2004) presents the new control structures for the hydrodealkylation of toluene (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 shows that hydrodealkylation of toluene 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 dynamic responses of the designed control structure and the reference structure are similar. The CS2 has been limited in bypass, so it is able to handle in small disturbance. 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 reboilers duty of column can control the column temperature more effective than bottom flow. CS5 on-demand structure has an advantage when downstream customer desires immediate responses in the availability of the product stream from this process.

Wongsri and Hermawan (2004), proposes the new heuristic of selection and manipulation heat pathway called heat pathway heuristics (HPH) for plantwide control. The HPH has been implemented in plantwide control for the hydrodealkylation of toluene(HDA) process with different energy integration schemes(i.e. alternatives 1, 4 and 6).A selective controller i.e. a low override switch (LOS) is employed in order to select an appropriate heat pathway through the process to carry the associated load to utility unit. The new designed plantwide control structure for HDA process is also compared with the earlier work given by Luyben et al. (1999). In general, better responses of the furnace and cooler utility consumptions are achieved compare to the Luyben's control structure. Both furnace and cooler duties could be decreased according to the input disturbance load, since the HPH is applied in the current work.

2.3 Heat Exchanger Network (HEN)

Energy conservation is important in process design. The fundamental result for the use of energy integration is to improve the thermodynamic efficiency of the process. This translates into a reduction in utility cost.

Linhoff and Hindmarsh (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% energy savings, coupled with capital savings, 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. There are:

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

2. The design for these separate problems 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 (increased utility usage) 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., 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 pinch of the process and either the reboilers or the condenser being integrated with the process. If these criteria can be met, energy cost for distillation can effectively be zero.

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 (the DESIGN problem), and sequence the control action of the loops, to accommodate set-point changes and reject load disturbances (the OPERATIONAL problem). The approach proposed exploits the structure characteristics of a HEN by identifying routes through the HEN structure that can allocate loads (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 implementational 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. Although this study provided the comprehensive summary of work on the design of control loop configuration in HENs, it did not report the control strategy, particularly in selecting and manipulating proper heat pathway. In this current study, we present the control strategy; how to select proper heat pathway to carry the associated load to a utility unit, so its duty will be decreased.

In series of papers, Terrill and Douglas (1987a,b,c) studied the sensitivity of the total processing cost to heat exchanger network alternatives and steady state operability evaluation. They considered the temperature-enthalpy (T-H) diagram and developed six HEN alternatives for a base case design for HDA process, in which their energy saving ranges 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 complex of the designs is alternative 6, recovers 43 percent of the base case net energy consumption. However, those alternatives have not been developed under dynamic simulation to study their dynamic aspects. In this dissertation, we present both the steady state and dynamic simulations for energy-integrated HDA plant (i.e. alternatives 5 and 6), as presented in Chapter 4, 5and 6

In 1990 Wongsri (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 changes 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 changers is call 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 condition where all process stream are at their minimum heat load, e.g. the input temperature of hot stream are at the lowest and those of cold streams are at the highest. Thus, only the positive disturbance load of process streams was considered. In this current work, we will consider both positive and negative disturbance loads that are originating from the process streams.

Ploypaisansang (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.

Wongsri and Sae-Leaw (2006) propose the guide line to design workable of highly heat integrated process with minimum auxiliary reboiler. It starts with specifying the disturbances and their magnitudes, and then designing the resilient heat exchanger network is designed at the worst case condition as the minimum heat supply and maximum heat demand condition. There considered only one worst case to find the number of minimum auxiliary heating unit and the heat path way for disturbance load at worst case condition is no considering dynamic maximum energy recovery(DMER).

CHAPTER III

PLANTWIDE CONTROL FUNDAMENTALS

Now a day many chemical plants are integrated process as material recycle and energy integration which increase interaction between unit operations. Therefore the control system that just combines the control schemes of each individual unit can't achieve its control objective. This can be solved by the plantwide process control strategy which designs a control system from the viewpoint of the entire plant. Hence, our purpose of this chapter is to present plantwide control fundamentals.

3.1 Incentives for Chemical Process Control

A chemical plant is an arrangement of processing units (reactors, heat exchangers, pumps, distillation columns, absorbers, evaporators, tanks, etc.), integrated with one another in a systematic and rational manner. The plant's overall objective is to convert certain raw materials into desired products using available sources of energy in the most economical way. There are three general classes of needs that a control system is called on to satisfy: suppressing the influence of external disturbances, ensuring the stability of a chemical process, and optimizing the performance of a chemical process (Stephanopoulos, 1984).

3.1.1 Suppressing the Influence of External Disturbances

Suppressing the influence of external disturbances on a process is the most common objective of a controller in a chemical plant. Such disturbances, which denote the effect that the surroundings have on a reactor, separator, heat exchanger, compressor, etc., are usually out of the reach of human operator. Consequently, we need to introduce a control mechanism that will make the proper change on the process to cancel the negative impact that such disturbances may have on the desired operation of a chemical plant. In other words, the strategies for control are very important to face all disturbances entering the process.

3.1.2 Ensuring the Stability of a Chemical Process

The process is stable or self-regulating if the process variable such as temperature, pressure, concentration, or flow rate stays at a certain point or at a desired steady state value as time progresses. Otherwise, the process is unstable and requires external control for the stabilization of their behavior.

3.1.3 Optimizing the Performance of a Chemical Process

Safety and the satisfaction of product specifications are the two principal operational objectives for a chemical plant. Once these are achieved, the next goal is how to make the operation of the plant more profitable. It is clear that we would like to be able to change the operation of the plant (flow rates, pressures, concentrations, temperatures) in such a way that an economic objective (profit) is always maximized.

3.2 Integrated Processes

Three basic features of integrated chemical processes lie at the root of the need to consider the entire plant's control system, as follows: the effect of material recycle, the effect of energy integration, and the need to account for chemical component inventories. However, there are fundamental reasons why each of these exists in virtually all-real processes.

3.2.1 Material Recycle

Material is recycled for six basic and important reasons

a. 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.

b. 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.

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

d. 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 so that reactor temperature increase will not be too large. High temperature can potentially create several unpleasant events, such as thermal runaway, deactivation of catalysts, cause undesirable side reaction, etc. So the heat of reaction is absorbed by the sensible heat required to raise the temperature of the excess material in the stream flowing through the reactor.

e. Prevent side reactions: A large excess of one of the reactants is often used 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 products components in the reactor effluent stream and recycled back to the reactor.

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

3.2.2 Energy Integration

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

3.2.3 Chemical Component Inventories

A plant's chemical species can be characterized into three types: reactants, products, and inerts. A material balance for each of these components must be satisfied. This is typically not a problem for products and inerts. However, the real problem usually arises when reactants (because of recycle) are considered and accounted for their inventories within the entire process. Because of their value, it is necessary to minimize the loss of reactants exiting the process since this represents a yield penalty. So we prevent reactants from leaving. This means we must ensure that every mole of reactant fed to the process is consumed by reactions.

3.3 Basic Concepts of Plantwide Control

3.3.1 Buckley Basics:

Buckley (1964) was the first to suggest the idea of separating the plantwide control problem into two parts: material balance control and product quality control. He suggested looking first at the flow of material through the system. A logical arrangement of level and pressure control loops is establishes, using the flowrates of the liquid and gas process streams. Note that most level controllers should be proportional-only (P) to achieve flow smoothing. He then proposed establishing the product-quality control loops by choosing appropriate manipulated variables. The time constants of closed-loop product quality loops are estimated. We try to make these as small as possible so that good, tight control is achieved, but stability constraints impose limitations on the achievable performance.

3.3.2 Douglas Doctrines:

Because of the cost of raw materials and the valves of products are usually much greater than the costs of capital and energy, Jim Douglas (1988) leads to the two Douglas doctrines:

a. Minimize losses of reactants and products
b. Maximize flowrates through gas recycle systems.

The first idea implies that we need tight control of stream compositions exiting the process to avoid losses of reactants and products. The second rests on the principle that yield is worth more than energy. Recycles are used to improve yields in many processes. The economics of improving yields (obtaining more desired products from the same raw materials) usually outweigh the additional energy cost of driving the recycle gas compressor.

3.3.3 Downs Drill

Chemical component balances around the entire plant are important things, and checking to see that the control structure handles these component balances effectively. The concepts of overall component balances go back to basic principle in chemical engineering, which is how to apply mass and energy balances to any system, microscopic or macroscopic. We check these balances for individual unit operations, for sections of a plant, and for entire processes. We must ensure that all components (reactants, products, and inert) have a way to leave or be consumed within the process. The consideration of inert is seldom overlooked. Heavy inert can leave the system in bottoms product from distillation column. Light inert can be purged from a gas recycle stream or from a partial condenser on a column. Intermediate inert must also be removed in some way, such as in side stream purges or separate distillation columns. Most of the problems occur in the consideration of reactants, particularly when several chemical species are involved. All of reactants fed into the system must either be consumed via reaction or leave the plant as impurities in exiting streams. Since we usually want to minimize raw material costs and maintain high-purity products, most of the reactant fed into the process must be chewed up in the reactions.

3.3.4 Luyben Laws

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

a. A stream somewhere in all recycle loops should be flow controlled. This is to prevent the snowball effect.

b. A fresh reactant feed stream cannot be flow controlled unless there is essentially complete one pass conversion of one of reactants. This law applies to systems with reaction types such as $A + B \rightarrow$ products. In system with consecutive reactions such as $A + B \rightarrow M + C$ and $M + B \rightarrow D + C$, the fresh feed can be flow controlled into the system, because any imbalance in the ratios of reactants is accommodated by a shift in the amounts of the two products (M and D) that are generated. An excess of A will result in the production of more M and less D. And vice versa, an excess of B results in the production of more D and less M.

c. If the final product from process comes out the top distillation column, the column feed should be liquid. If the final product comes out from the bottom of the column, the column feed should be vapor. Changes in feed flowrate or feed composition have less of a dynamic effect on distillate composition than they do on bottoms composition if the feed is saturated liquid. The reverse is true if the feed is saturated vapor: bottom is less affected than distillate.

3.3.5 Richardson Rule

Richardson (1995) suggested the heuristic that the largest stream should be selected to control the liquid level in a vessel. This makes good sense because it provides more muscle to achieve the desired control objective. The largest stream has the biggest effect to volume of vessel. An analogy is that it is much easier to maneuver a large barge with a tugboat that a life raft. The point is that the bigger the handle you have to affect a process, the better you can control it.

3.3.6 Shinskey Schemes

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

3.3.7 Tyreus Tuning

One of the vital steps in developing a plantwide control system is how to determine the algorithm to be used for each controller (P, PI or PID) and to tune each controller. The use of P-only controllers is recommended for liquid levels. Tuning of P controller is usually trivial, that is set the controller gain equal to 1.67. This will have the valve wide open when the level at 80% and the valve shut when the level is at 20%. For other control loops, the use of PI controllers is suggested. The relay-feedback test is simple and fast way to obtain the ultimate gain (Ku) and ultimate period (Pu). The Ziegler-Nichols settings or the Tyreus-Luyben settings can be used for tuning the parameters of controller:

$$K_{ZN} = K_u/2.2 \qquad \tau_{ZN} = P_u/1.2$$
$$K_{TL} = K_u/3.2 \qquad \tau_{TL} = 2.2\ P_u$$

The use of PID controllers should be restricted to those loops were two criteria are both satisfied: the controlled variable should have a very large signal-to-noise ratio and tight dynamic control is really essential.

3.4 Plantwide Control Design Procedure

The plantwide control procedure has been established based upon heuristics (Luyben et al., 1997). The nine steps of the design 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; make up of reactants; component balances; and economic or process optimization. This heuristic design procedure is described below.

3.4.1 Establish Control Objectives

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

3.4.2 Determine Control Degrees of Freedom

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

3.4.3 Establish Energy Management System

Term energy management is used to describe two functions. First, we must provide a control system that remove 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. If heat integration does occur between process streams, then the second function of energy management is to provide a control system that prevents propagation of the thermal disturbances and ensures that the exothermic reactor heat is dissipated and not recycled. Process-to-process heat exchangers and heat-integrated unit operations must be analyzed to determine that there are sufficient degrees of freedom for control. 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 (e.g., by setting the ratio of the flow rate of the limiting fresh reactant to the flow rate of a recycle stream acting as a thermal sink). Increased use of heat integration can lead to complex dynamic behavior and poor performance due to recycling of disturbances. If not already in the design, trim heaters/coolers or heat exchanger bypass lines 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.

3.4.4 Set Production Rate

Establish the variables that dominate the productivity of the reactor and determine the most appropriate manipulator to control production rate. Often design constraints require that production be set at a certain point. An upstream process may establish the feed flow sent to the plant. A downstream process may require on-demand production, with fixes the product flow rate from the plant. If no constraint applies, then we select the valve that provides smooth and stable production-rate transitions and rejects disturbances. We often want to select the 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 heating an operational constraint. This may be the feed flow to the separation section, the flow rate of recycle stream, the flow rate of initiator or catalyst to the reactor, the reactor heat removal rate, the reactor temperature, and so forth.

3.4.5 Control Product Quality and Handle Safety, Operational and Environmental Constraints

Select the best valves to control each of the product-quality, safety, and environmental variables. We want tight control of these quantities for economic and operational reasons. Hence we should select manipulated variables such that the dynamic relationships between controlled and manipulated variables feature small time constants and dead times and large steady-state gains. The former gives small closed-loop time constants, and the latter prevents problems with the range-ability of the manipulated variable (control-valve saturation)

3.4.6 Fix a Flow in Every Recycle Loop and Control Inventories (Pressure and Liquid Level)

Determine the valve to control each inventory variable. These variables include all liquid levels (except for surge volume in certain liquid recycle streams) and gas pressures. An inventory variable should typically be controlled with the manipulated variable that has the largest effect on it within that unit.

Proportional-only control should be used in non-reactive control loops for cascade unit in series. Even in reactor-level control, proportional control should be considered to help filter flow-rate disturbances to the down stream separation system. There is nothing necessarily sacred about holding reactor level constant. In most processes a flow controller should be present in all liquid recycle loops. This is a simple and effective way to prevent potentially large changes in recycle flows that can occur if all flows in recycle loops are controlled by levels. Two benefits result from this flow-control strategy. First, the plant's separation section is not subjected to large load disturbances. Second, consideration must be given to alternative fresh reactant makeup control strategies rather than flow control. In dynamic sense, level controlling all flows in recycle loop is a case of recycling of disturbances and should be avoided.

3.4.7 Check Component Balances

Identify how chemical components enter, leave, and are generated or consumed in the process. Ensure that the overall component balance for each species can be satisfied either through reaction or exit streams by accounting for the component's composition or inventory at some point in the process. Light, intermediate, and heavy inert components must have an exit path from the system. Reactant must be consumed in the reaction section or leaves as impurities in the product streams. Fresh reactant makeup feed stream can be manipulated to control reactor feed composition or a recycle stream composition (or to hold pressure or level as noted in previous step). Purge stream can also be used to control the amount of high- or low-boiling impurities in a recycle stream.

3.4.8 Control Individual Unit Operations

Establish the control loops necessary to operate each of the individual unit operations. For examples, a tubular reactor usually requires control of inlet temperature. High-temperature endothermic reactions typically have a control system to adjust fuel flow rate to a furnace supplying energy to the reactor. Crystallizers require manipulation of refrigeration load to control temperature. Oxygen concentration in stack gas from a furnace is controlled to prevent excess fuel usage. Liquid solvent feed flow to an absorber is controlled as some ratio to the gas feed.

3.4.9 Optimize Economic and 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 performance (e.g., minimize energy, maximize selectivity) or to improve dynamic response.

For example, an exothermic chemical reactor can be cooled with both jacket cooling water and brine to a reflux condenser. A valve-position control strategy would allow fast, effective reactor temperature control while minimizing brine use.

3.5 Plantwide Energy Management

Energy conservation has always been important in process design. Thus, it is common practice to install feed-effluent heat exchangers (FEHEs) around rectors and distillation columns. In any process flowsheet, a number of streams must be heated, and other streams must be cooled. For example, in HDA process, the toluene fresh feed, the makeup hydrogen, the recycle toluene, and the recycle gas stream must be heated up to the reaction temperature 621.1oC. And, the reactor effluent stream must also be cooled to the cooling water temperature to accomplish a phase split. Therefore, the energy integration is required to reduce the utility cost and also to improve thermodynamic efficiency of the process.

3.5.1 Heat Exchanger Dynamics

Heat exchangers have fast dynamics compared to other unit operations in a process. Normally the time constant is measured in second but could be up to a few minutes for large exchangers. Process-to-process exchangers should be modeled rigorously by partial differential equations since they are distributed systems. This introduces the correct amount of dead time and time constant in the exit stream temperatures, but the models are inconvenient to solve.

For the purpose of plantwide control studies it is not necessary to have such detailed descriptions of the exchanger dynamics, since these units rarely dominate the process response. Instead, it is often possible to construct useful models by letting two sets of well-stirred tanks in series exchange heat. This simplifies the solution procedure.

3.5.2 Heat Pathways

A path is a connection between a heater and a cooler in a network. In plantwide energy management, various pathways for heat need to be identified. Furthermore, a control strategy that allows effective delivery and removal of energy is needed to minimize propagation of thermal disturbances. It is important to realize that there are no thermodynamic restrictions on the energy requirement to transition streams between unit operations. In other words, the heating and cooling of streams are done for practical reasons and not to satisfy the laws of thermodynamics. This energy would not be an issue if all the processing steps operated at the same constant temperature. Furthermore, since raw materials and products are stored at roughly the same temperature, the net energy requirement for heating and cooling equal the heat losses from the process.

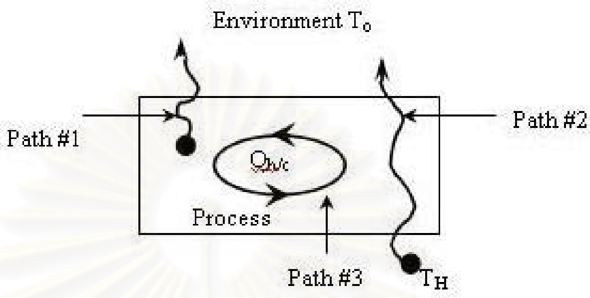


Figure 3.1 Heat pathways

From plantwide perspective we can now discern three different "heat pathways" in the process as illustrated in Figure 3.1. The first pathway, heat from the process is dissipated to the environment, e.g. heat generated by exothermic reactions and by degradation of mechanical work (e.g., compression, pressure drop and friction). This pathway is from inside the process and flows out. It is of course possible to convert some of the heat to work as it is removed from high temperature in the process.

A second pathway carries heat from utilities into the process. Mechanical work is extracted from the heat as it flows from a high supply temperature to the lower temperature of the environment. This pathway goes through the process and is needed to satisfy the thermodynamic work requirements of separation. Work is also extracted from the heat stream to overcome process inefficiencies associated with stream mixing and heat transfer.

The third pathway is internal to the process. Here heat flows back and forth between different unit operations. The magnitude of this energy path depends upon the heating and cooling needs and the amount of heat integration implemented. Whenever the internal path is missing, and there is a heating requirement, the heat has to be supplied from utilities. The same amount of heat must eventually be rejected to the environment elsewhere in the process.

3.5.3 Heat Recovery

Usually, chemical processes are thermally inefficient. First, the chemical work available in the reactants is dissipated as heat. Second, the work required for separation is usually supplied as heat to distillation column, which has internal inefficiencies. Finally, energy is needed for heating and cooling functions that are independent of thermodynamic constraints. This all adds up to a low thermal efficiency.

Fortunately, we can make great improvements in plant's thermal efficiency by recycling much of the energy needed for heating and cooling process streams. It is also possible to introduce heat integration schemes for distillation columns to reduce the separation heat. And finally the reaction heat can be recovered in waste heat boilers and use the steam for power generation. There is of course a capital expense associated with improved efficiency but it can usually be justified when the energy savings are accounted for during the lifetime of the project. Of more interest to us in the current context is how heat integration affects the dynamics and control of a plant and how we can manage energy in plants with a high degree of heat recovery.

3.6 Control of Process-to-Process Exchangers

Process-to-process (P/P) exchangers are used for heat recovery within a process. Most heat exchanger network are not operable at the optimum steady state design conditions; i.e., normally they can tolerate disturbances that decrease the loads but not those that increase loads and there are not an adequate number of manipulative variables to be able to satisfy the process constraints and to optimize all of the significant operating variables. These types of operability limitations can be identified by using steady state considerations, and normally these operability limitations can be overcome by installing an appropriate utility exchanger and by installing bypass around the exchangers.

3.6.1 Use of Auxiliary Exchangers

When the P/P exchanger is combined with a utility exchanger, we also have a few design decisions to make. The utility exchanger can be installed to P/P exchanger either in series or parallel. Figure 3.2 shows the combination of P/P exchanger with a utility exchanger. Generally, the utility system of a complex energy-integrated plant is designed to absorb large disturbances in the process, and making process-to-utility exchangers relatively easy to control.

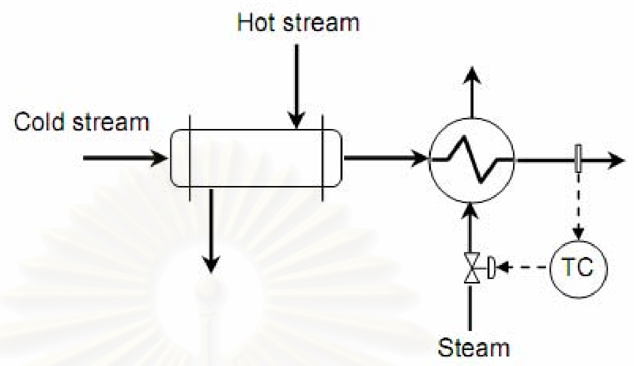


Figure 3.2 Control of process-to-process heat exchanger using the auxiliary utility.

The relative sizes between the recovery and the utility exchangers must be established. From a design standpoint we would like to make the recovery exchanger large and utility exchanger small. This gives the most heat recovery, and it is also the least expensive alternative from an investment standpoint.

3.6.2 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 Fig 3.3.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.

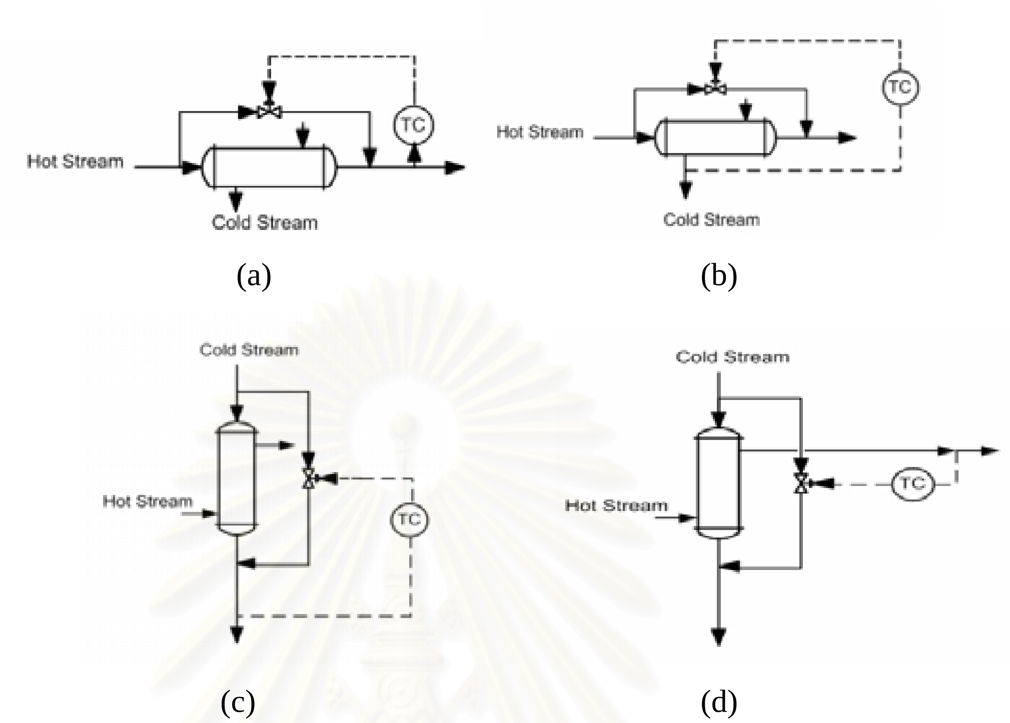


Figure 3.3 Bypass control of process-to-process heat exchangers. (a) Controlling and bypassing hot stream; (b) controlling cold stream and bypassing hot stream; (c) controlling and bypassing cold stream; (d) controlling hot stream and bypassing hot stream.

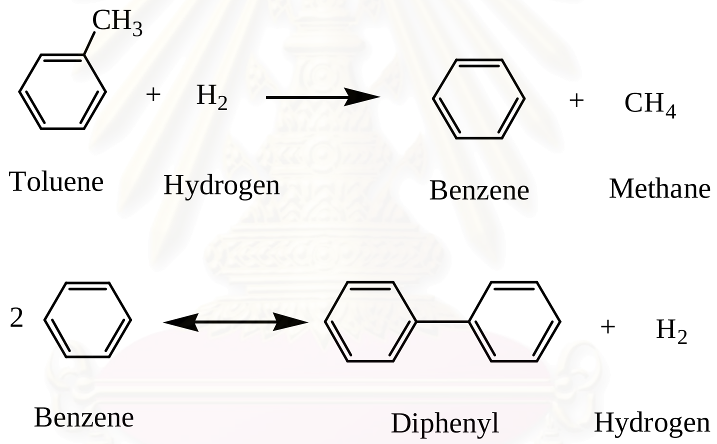
Finally, we must carefully consider the fluid mechanics of the bypass design for the pressure drops through the control valves and heat exchanger.

CHAPTER IV

HDA PROCESS

4.1 Process Description

The hydrodealkylation HDA of toluene process (alternative 1) by Douglas (1988) on conceptual design as in Figure 4.1 contain nine basic unit operations: reactor, furnace, vapor-liquid separator, recycle compressor, two heat exchangers, and three distillation columns. Two raw materials, hydrogen, and toluene, are converted into the benzene product, with methane and diphenyl produced as by-products. The two vapor-phase reactions are



The kinetic rate expressions are functions of the partial pressure (in psia) of toluene p_T, hydrogen p_H, benzene p_B, and diphenyl p_D, with an Arrhenius temperature dependence. Zimmerman and York (1964) provide the following rate expression:

$$r_1 = 3.6858 * 10^6 \exp(-25616/T)\, p_T\, p_H^{1/2}$$
$$r_2 = 5.987 * 10^4 \exp(-25616/T)\, p_B^2 - 2.553 * 10^5 \exp(-25616/T)\, p_D\, p_H$$

Where r_1 and r_2 have units of lb*mol/(min*ft^3) and T is the absolute temperature in Kelvin. The heats of reaction given by Douglas (1988) are –21500 Btu/lb*mol of toluene for r_1 and 0 Btu/lb*mol for r_2.

The effluent from the adiabatic reactor is quenched with liquid from the separator. This quenched stream is the hot-side feed to the process-to-process heat exchanger, where the cold stream is the reactor feed stream prior to the furnace. The reactor effluent is then cooled with cooling water and the vapor (hydrogen, methane) and liquid (benzene, toluene, and diphenyl) are separated. The vapor stream from the separator is split and the remainder is sent to the compressor for recycle back to the reactor.

The liquid stream from the separator (after part is taken for the quench) is fed to the stabilizer column, which has a partial condenser component. The bottoms stream from the stabilizer is fed to the product column, where the distillate is the benzene product from the process and the bottoms is toluene and diphenyl fed to the recycle column. The distillate from the recycle column is toluene that is recycled back to the reactor and the bottom is the diphenyl byproduct.

Makeup toluene liquid and hydrogen gas are added to both the gas and toluene recycle streams. This combined stream is the cold-side feed to the process-to-process heat exchanger. The cold-side exit stream is then heated further up to the required reactor inlet temperature in the furnace, where heat is supplied via combustion of fuel.

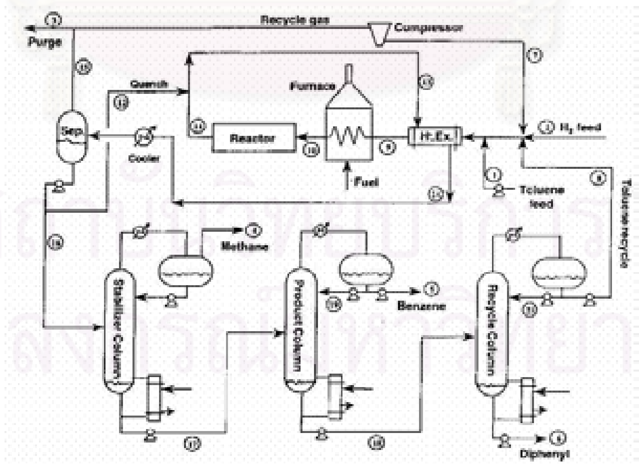


Figure 4.1 Hydrodealkylation(HDA) of Toluene Process (alternative 1)

Component physical property data for the HDA process were obtain from Luyben et al. (1999)

4.2 Heat Integration of Hydrodealkylation(HDA) of Toluene Process

Terrill and Douglas (1987) design six different energy-saving alternatives to the base case. The simplest of these designs (alternative 1) recovers an additional 29% of the base case heat consumption by making the reactor preheater larger and the furnace smaller.

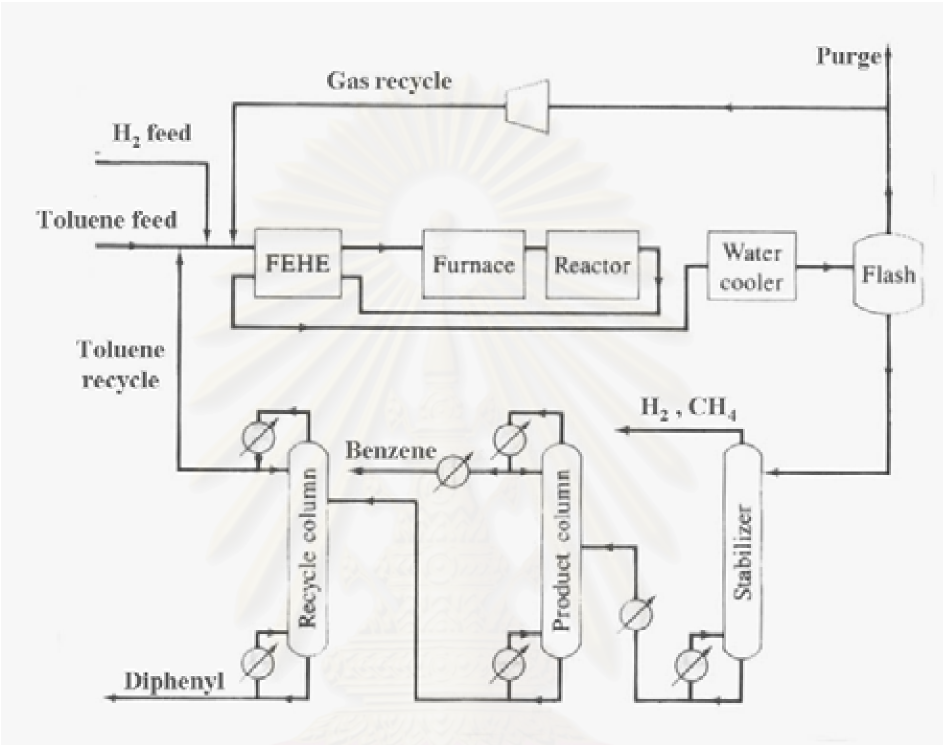


Figure 4.2 HDA process –alternative 1

In alternative 2 (Figure 4.3) is the same as alternative 1, except that recycle column was pressure shifted to be above the pinch temperature, and the condenser for the recycle column is used to drive the product column reboiler.

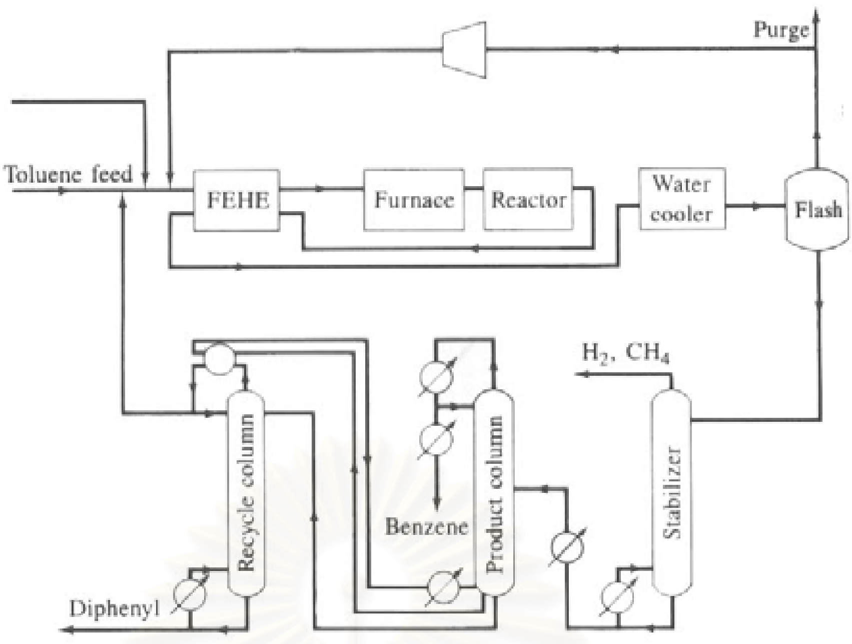


Figure 4.3 HDA process –alternative 2

In alternative 3 part of the heat in the reactor effluent stream is used to drive the stabilizer reboiler, recycle column was pressure shifted to be above the pinch temperature, and the condenser for the recycle column is used to drive the product column reboiler as in Figure 4.4

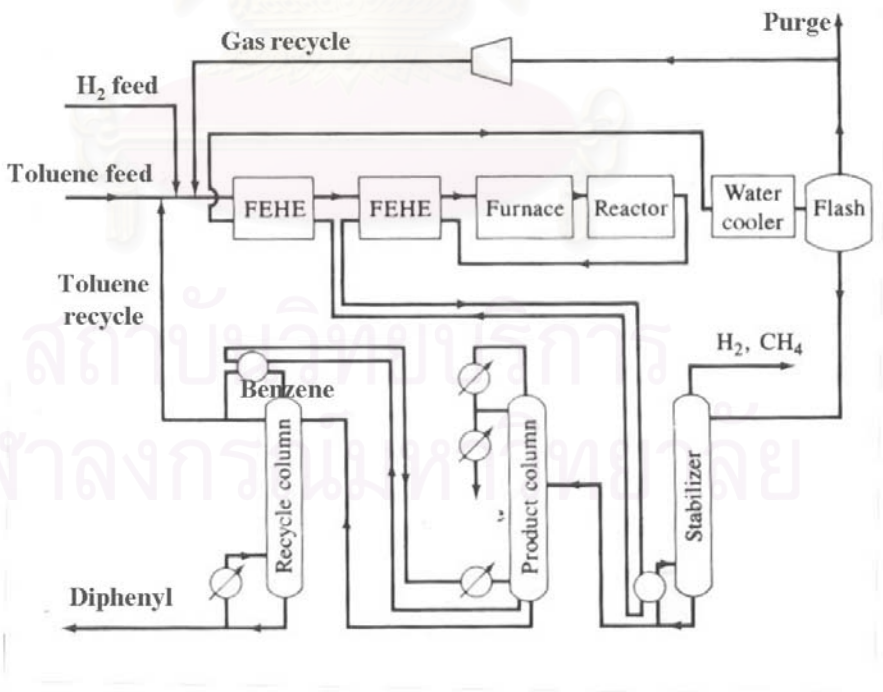


Figure 4.4 HDA process –alternative 3

In alternative 4 the reactor effluent is used to drive the product column reboiler. recycle column was pressure shifted to be above the pinch temperature

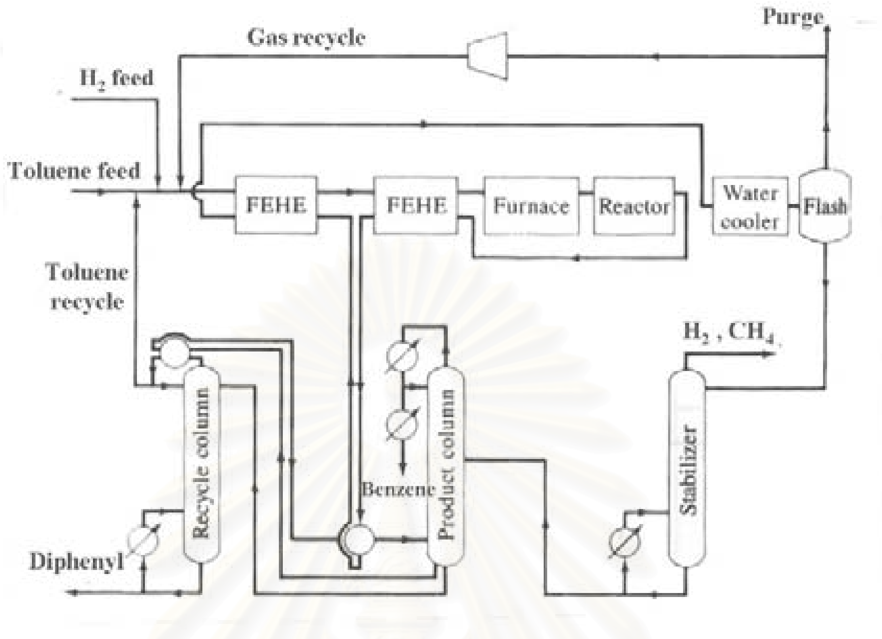


Figure 4.5 HDA process –alternative 4

For alternative 5, both the stabilizer reboiler and the product column reboiler are driven consecutively by the reactor effluent stream, recycle column was pressure shifted to be above the pinch temperature as in Figure 4.6

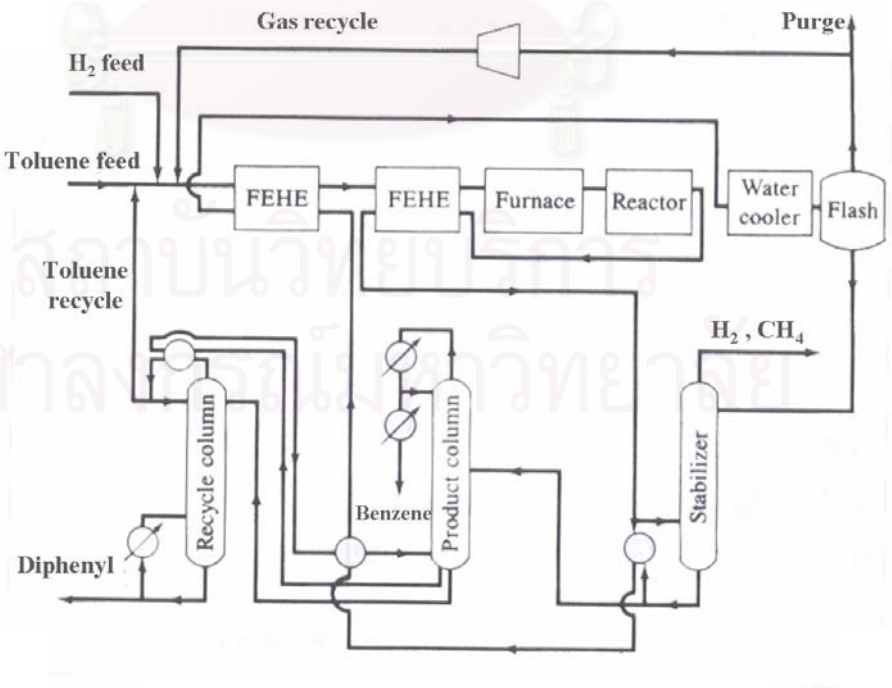


Figure 4.6 HDA process –alternative 5

In alternative 6 all three column reboilers are driven by the reactor effluent stream, recycle column was pressure shifted to be above the pinch temperature, and the condenser for the recycle column is used to drive the product column reboiler as in Figure 4.7

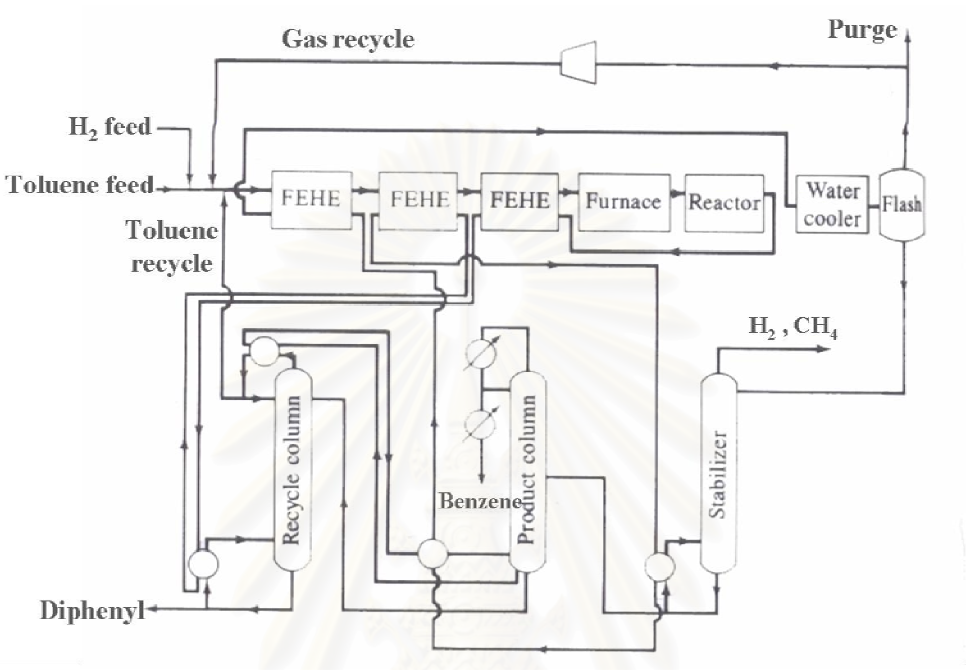


Figure 4.7 HDA process –alternative 6

The energy saving from the energy integration fall between 29 and 43 %, but the cost saving are in the range from -1 to 5 %. This work will consider the HDA 5 and 6 because there are highly heat integrated process.

4.3 Steady State Modeling

First, a steady-state model is built in HYSYS.PLANT, using the flowsheet and equipment design information, mainly taken from Douglas (1988); Luyben et al. (1998) to develop for HDA alterative 5 and 6. Figures 4.8 and 4.9 show the HYSYS flowsheets of the HDA process with energy integration schemes for alternatives 5 and 6, respectively. The data for the selected streams for these alternatives are not included in this chapter but listed in Appendix A. For our simulation, 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).

Since there are many material recycles, as RECYCLE operations in HYSYS are inserted in the streams. 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.

When columns are modeled in steady-state, besides the specification of inlet streams, pressure profiles, numbers of trays and feed tray, two specifications need to be given for columns with both reboiler and condenser. These could be the duties, reflux rate, draw stream rates, composition fractions, etc. We chose reflux ratio and overhead benzene mole fraction for the stabilizer column. For the remaining two columns, bottom and overhead composition mole fractions are specified to meet the required purity of products given in Douglas (1998). The tray sections of the columns are calculated using the tray sizing utility in HYSYS, which calculates tray diameters based on Glitsch design parameters for valve trays. Though the tray diameter and spacing, and weir length and height are not required in steady-state modeling, they are required for dynamic simulation.

The bypass flow of the process to process heat exchanger is designed at 5% of total flowrates. In order to increase the upper limit of bypass flow, the percent opening of bypass valve at 5% of total flowrates is design 3% opening valve for steady state design.

For the process to process heat exchanger, we would like to over design area heat exchanger in dynamic mode to increase the control range. This can do in HYSYS simulator by increasing UA of heat exchange.

Before the simulation change mode from steady state mode to dynamic mode, the sizing equipment has to specify. The data and specifications for the equipments are summarized in Appendix B.

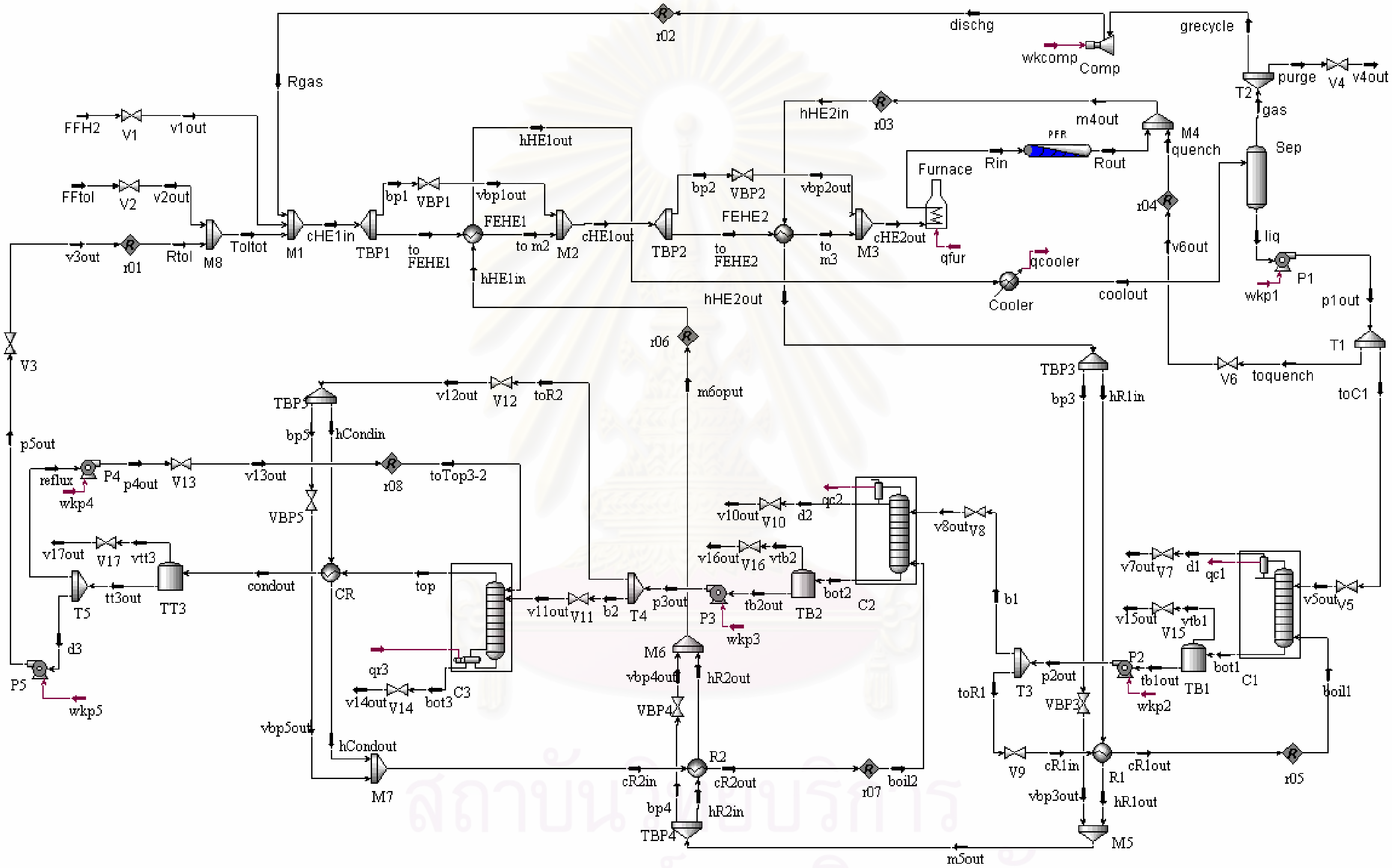


Figure 4.8 Hysy Flowsheet of the Steady State Modeling of HDA Process Alternative 5

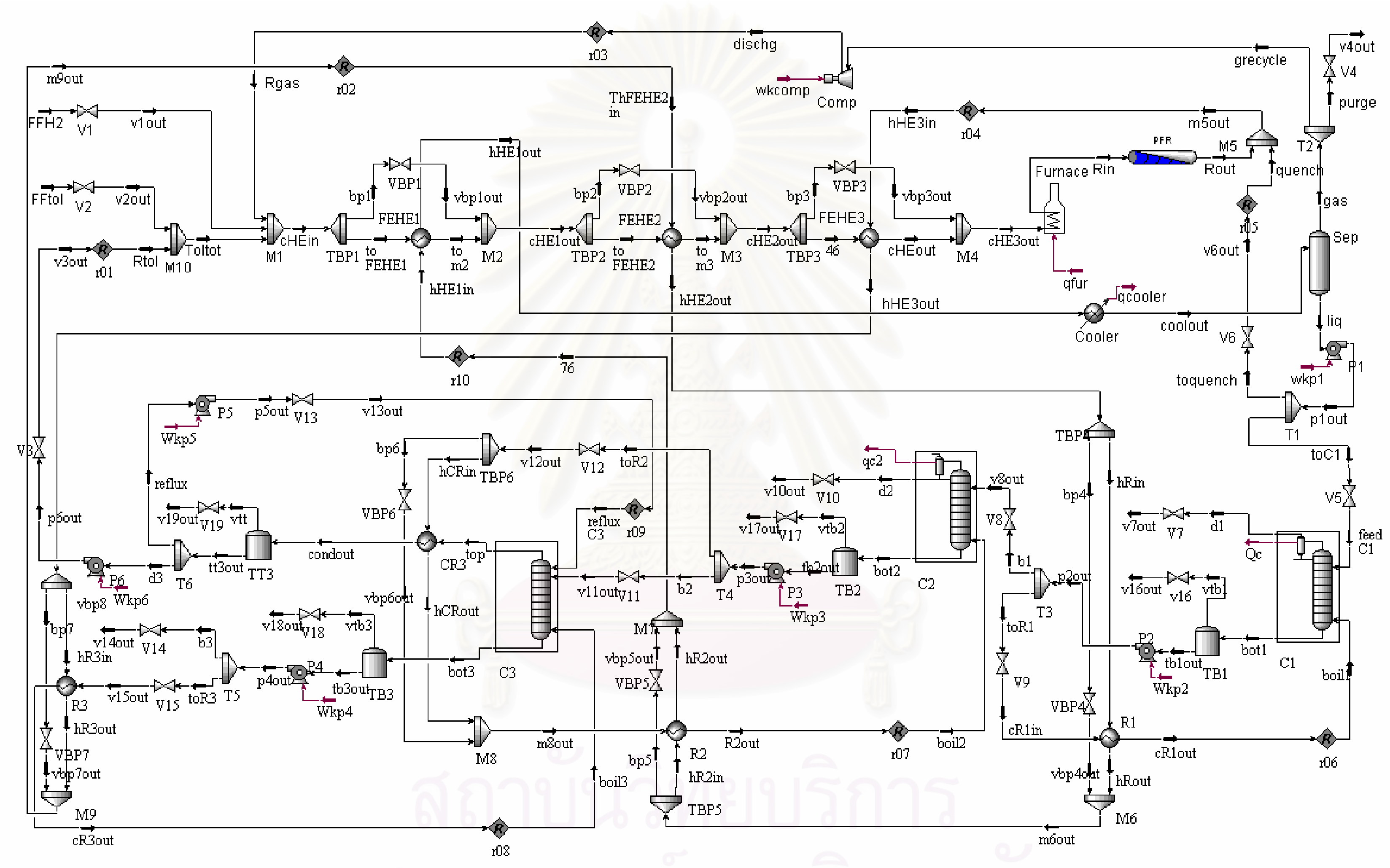


Figure 4.9 Hysy Flowsheet of the Steady State Modeling of HDA Process Alternative 6

CHAPTER V

DESIGN OF WORKABLE HEAT-INTEGRATED HDA PROCESS WITH MINIMUM AUXILIARY UNTILITY UNITS

For complex heat-integrated process, the heater and cooler are replaced by process to process heat exchanger. To control a highly heat integration scheme, design modifications to process is needed to ensure controllability and operability. Luyben (1999) solves some of the control difficulties by adding auxiliary utility to the end of hot and cold streams that have no utility unit in order to assure target temperature each stream when disturbance come through the process. The many auxiliary units introduce the rising economic cost therefore the purpose in this chapter is to illustrate the strategy for design of workable process for highly heat integrated process with minimum auxiliary utility units. Addition, this strategy is applied to HDA process with complex energy integration alternatives 5 and 6.

5.1. The Strategy for Estimating the Minimum Auxiliary Utility Units

Our strategy for design workable heat-integrated HDA process with minimum auxiliary utility units comprising of 4 steps, is described as follows:

1) Determined the Expected Disturbances and Their Magnitudes

The disturbance load requirement is determined in this step. Only temperature variation is considered here. The magnitude of the variation in this work is $\pm 10\ ^{o}C$ as sensible heat load variation.

2. Design the Worst Case Conditions

Wongsri (1990) defined four extreme conditions in order to design resilient heat exchanger network. Two of them are used to be worst case conditions as follows:

2.1. Minimum Heating Condition.

This is a condition where hot process streams are at their minimum heat loads (minimum cooling requirement) whereas cold process streams are at their maximum heat loads (maximum heating requirement). For example, inlet temperatures of hot and cold streams are the lowest. This worst case condition will be used to estimate the minimum number of auxiliary heating units at the end of cold streams.

2.2. Maximum Cooling Condition.

This is a condition hot process streams are at their maximum heat loads (maximum cooling requirement) whereas cold process streams are at their minimum heat loads (minimum heating requirement). For example, inlet temperatures of hot and cold streams are the highest. Therefore this worst case condition will be used to estimate the minimum number of auxiliary cooling units at the end of hot streams.

In previous step, the magnitude disturbance is ± 10 ^{o}C. Hence minimum heating condition is a condition where hot and cold process streams are the temperature at normal operating condition (T_{normal}) minus 10^{o}C (see Figure 5.1.a). For maximum cooling condition, hot and cold process streams are the temperature at normal operating condition (T_{normal}) plus 10^{o}C (see Figure 5.1.b).

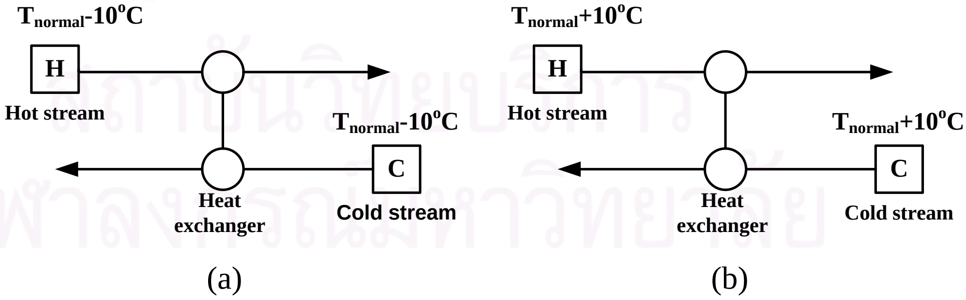


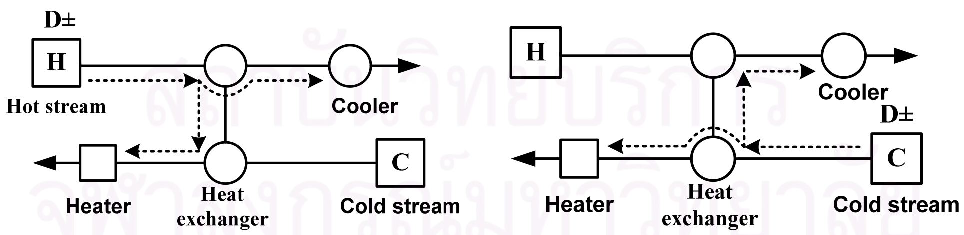
Figure 5.1 Worst Case Condition (a) Minimum heating (b) Maximum cooling condition

3. Design Heat Pathways for the Worst Case Conditions

When disturbance load come to hot and cold process streams, the important thing is how to remove disturbance load. It has a few heat pathways to remove disturbance. However the heat pathway that can achieve dynamic maximum energy recovery (DMER) is considered in this work.

3.1. Disturbance Loads and Propagations

In the process heat integration, there are two kinds of disturbance loads (Wongsri, 1990). The first disturbance load is Positive disturbance load D+ i.e. a disturbance that will increase the heat load of stream. For example, when the inlet temperature of a disturbed hot stream increases or when the inlet temperature of a disturbed cold stream decreases. The second disturbance load is Negative disturbance load D- i.e. a disturbance that will decrease the heat load of stream. For example, when the inlet temperature of a disturbed hot stream decreases or when the inlet temperature of a disturbed cold stream increases. The shift approach is presented by Wongsri, 1990 that the disturbance load can be viewed as a heat packet that must be propagated as much as possible by transferring or shifting it to heat sinks (coolers) or heat sources (heater) of a network such as a simplified heat exchanger network as shown in Figure 5.2. Another concept that will be used together with shift approach is disturbance propagation design (Wongsri, 1990) as the disturbance load of a smaller heat load stream will be shifted to a larger heat load stream.

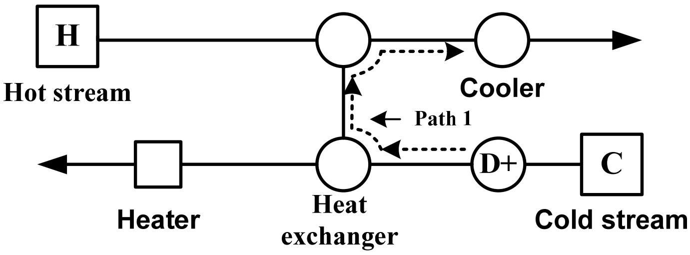


(a) (b)

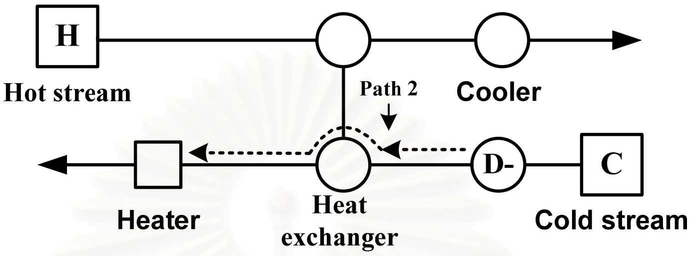
Figure 5.2 the Disturbance Load is Propagated to Heater and Cooler (a) positive and negative disturbance load in hot stream cases (b) positive and negative disturbance load in cold stream cases

3.2. Design of Heat Pathways to achieve dynamic maximum energy recovery (DMER)

Wongsri and Hermawan (2005) proposed the heat pathway heuristics (HPH) that is used for selecting an appropriate heat pathway to carry associated load to a utility in order to achieve dynamic maximum energy recovery (DMER) such as the positive effects of the disturbance loads on the utility requirements, we attempt to shift the heat load disturbances to either heater or cooler utility unit in a way that the DMER is realized. Therefore, the utility duties of these units will be decreased as the disturbance is entering the process. A simplified heat exchanger network as shown in Figure 4.5 is used to explain how an appropriate heat pathway should be activated to carry associated load to the utility unit. For instance, when the inlet temperature of a disturbed cold stream decreases, path 1 (Figure 5.3.a) should be activated by controlling the cold outlet temperature of heat exchanger. This will have the effect of shifting the positive disturbance load to the cooler. Thus, the positive disturbance load of a cold stream will result in decrease of the cooler duty. Consider the case when the inlet temperature of a disturbed cold stream increases, path 2 (Figure 5.3.b) should be activated by controlling the hot outlet temperature of heat exchanger to shift its negative disturbance load to heater. Thus, the negative disturbance load of a cold stream will result in decrease of the heater duty. On the other hand, when the inlet temperature of a disturbed hot stream increases, path 3 (Figure 5.3.c) should be activated by controlling the hot outlet temperature of heat exchanger to shift its positive disturbance load to heater. As a result, the heater duty will be decreased. Consider the case when the inlet temperature of a disturbed hot stream decreases, path 4 (Figure 5.3.d) should be activated by controlling the cold outlet temperature of heat exchanger to shift its negative disturbance load to cooler. As a result, the cooler duty will be decreased. The appropriate heat pathways for each disturbance are summarized in table 5.1.



(a)



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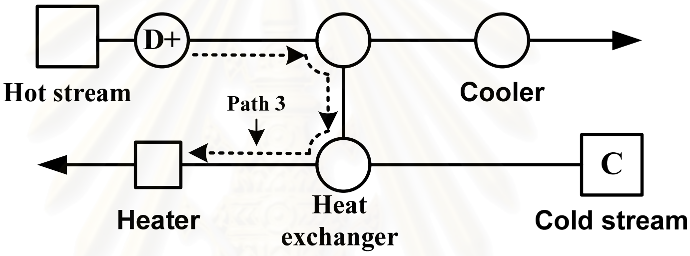
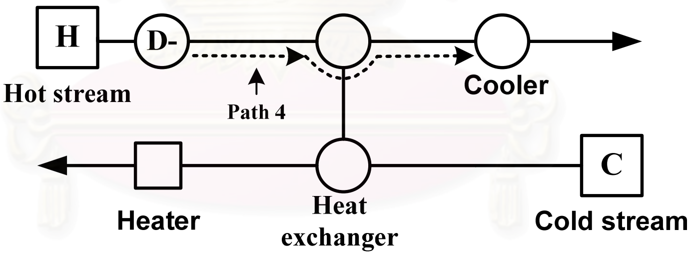


(c)



(d)

Figure 5.3 Heat Pathways through HEN with single heat exchanger to achieve DMER, where: (a) path 1 is used to shift the positive disturbance load of the cold stream to the cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream to heater, (c) path 3 is used to shift the positive disturbance load of the hot stream to heater, and (d) path 4 is used to shift the negative disturbance load of hot stream to the cooler.

Table 5.1 the Appropriate Utility for Removing Disturbance Loads to Achieve DMER

disturbance load	source	appropriate utility for achieved DMER
Positive disturbance Load(D+) of cold stream	the inlet temperature of cold stream decreases	cooler
Positive disturbance load(D+) of hot stream	the inlet temperature of hot stream increases	heater
Negative disturbance load(D-) of cold stream	the inlet temperature of cold stream increases	heater
Negative disturbance load(D-) of hot stream	the inlet temperature of hot stream decreases	cooler

Therefore in our worst case, the appropriate heat path way for minimum heating condition is shifting of the disturbance load to cooler (see figure 5.4.a).Another worst case, the appropriate heat path way for maximum cooling condition is shifting of the disturbance load to heater (see figure 5.4.b).

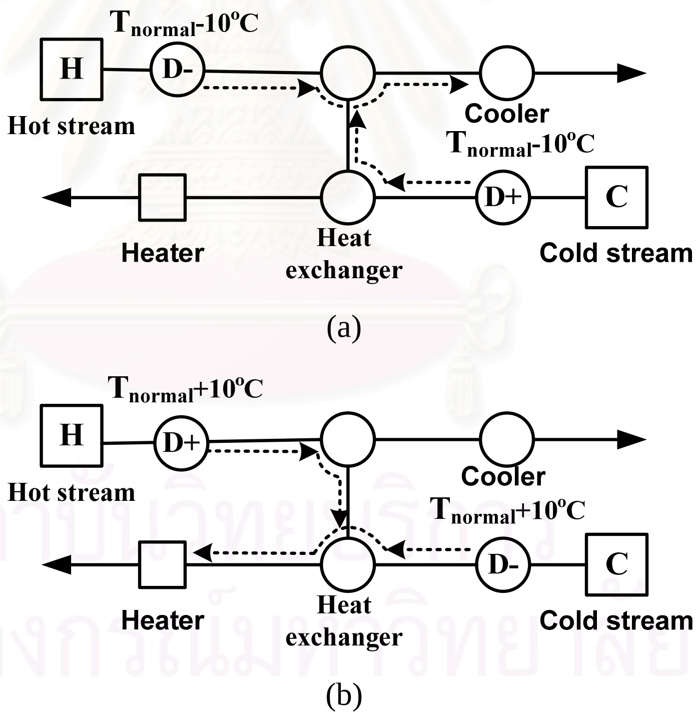


Figure 5.4 the Appropriate Heat Pathway for the Worst Case Condition a) minimum heating condition b) maximum cooling condition

4. Estimated the Minimum Auxiliary Utility Units

The minimum auxiliary utility unit is estimated in this step consists of 2 parts as estimated the minimum auxiliary heating and cooling units

4.1. Estimated the Minimum Auxiliary Heating Units

In this part, the minimum heating condition (see figure 5.1.a) is employed to heat exchanger network and then the disturbance load is dissipated to cooler that is the appropriate heat path way. If the end of cold stream can't achieve target temperature at worst case condition, the auxiliary heating unit has to be employed.

4.2. Estimated the Minimum Auxiliary Cooling Units

For this part, the maximum cooling condition (see figure 5.1.b) is employed to heat exchanger network and then the disturbance load is dissipated to heater that is the appropriate heat pathway. If the end of hot stream can't achieve target temperature in worst case condition, the auxiliary cooling unit has to be employed.

5.2. Evaluated the Minimum Auxiliary Utility Units for HDA Process with Complex Energy Integration Alternatives 5 and 6

The minimum auxiliary utility units for HDA process with complex energy integration alternatives 5 and 6 are proposed in this part and our strategy will be applied.

5.2.1 Evaluated the Minimum Auxiliary Utility Units for HDA Process with Complex Energy Integration Alternative 5

Heat exchanger network of HDA process alternative 5 is presented in figure 5.5. H1 is the quenched reactor product stream, and H2 is the hot-side stream from the top tray of recycle column. C1 is the reactor feed steam prior to the furnace. C2 and C3 are the cold-side steams coming from the bottom of product and stabilizer, respectively.

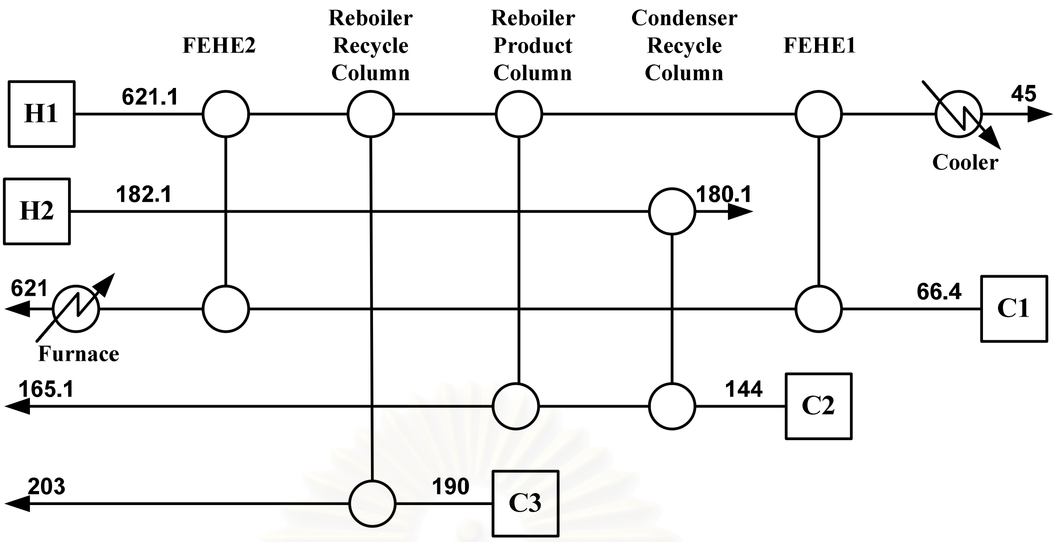


Figure 5.5 Heat Exchanger Network of HDA Process Alternative 5

1) Determined the Expected Disturbances and Their Magnitudes

Disturbance load for HDA process alternative 5 is defined that the magnitude of the disturbance variation is ± 10 ^{o}C from normal operating temperature (see table 5.2).

Table 5.2 the Magnitude of Disturbance Load for Heat Exchanger Network HDA Process Alternative 5

Stream	Disturbance load	
	Maximum inlet temperature (T_{normal}+10^{o}C)	Minimum inlet temperature (T_{normal}-10^{o}C)
H1	631.09	611.09
H2	192.14	182.13*
C1	76.42	56.42
C2	154.00	134.00
C3	200.00	180.00

Note * the phase change occurs in this stream from gas to liquid when the inlet temperature decreases by 10 ^{o}C. This will make a huge disturbances load. In order to avoid this problem, the heat flow in this stream is decreased equal to the amount of sensible heat for 10 ^{o}C. Therefore this stream will decrease to 182.13 ^{o}C.

2. Design the Worst Case Conditions

2.1. Minimum Heating Condition.

This is a condition where the inlet temperatures of hot and cold streams are the lowest (see table 5.3).

Table 5.3 the Inlet Temperature of Heat Exchanger Network HDA Process Alternative 5 at Minimum Heating Condition

Stream	inlet temperature (T_{normal}-10^{o}C)
H1	611.09
H2	182.13
C1	56.42
C2	134.00
C3	180.00

2.2. Maximum Cooling Condition.

This is a condition where the inlet temperatures of hot and cold streams are the highest (see table 5.4).

Table 5.4 the Inlet Temperature of Heat Exchanger Network HDA Process Alternative 5 at Maximum Cooling Condition

Stream	inlet temperature (T_{normal}+10^{o}C)
H1	631.09
H2	192.14
C1	76.42
C2	154.00
C3	200.00

3. Design Heat Pathways for the Worst Case Conditions

3.1. Minimum Heating Condition.

In order to achieve DMER for minimum heating condition, the disturbance load is dissipated to cooler. Therefore the appropriate heat pathway at minimum heating condition is designed as show in figure 5.6.

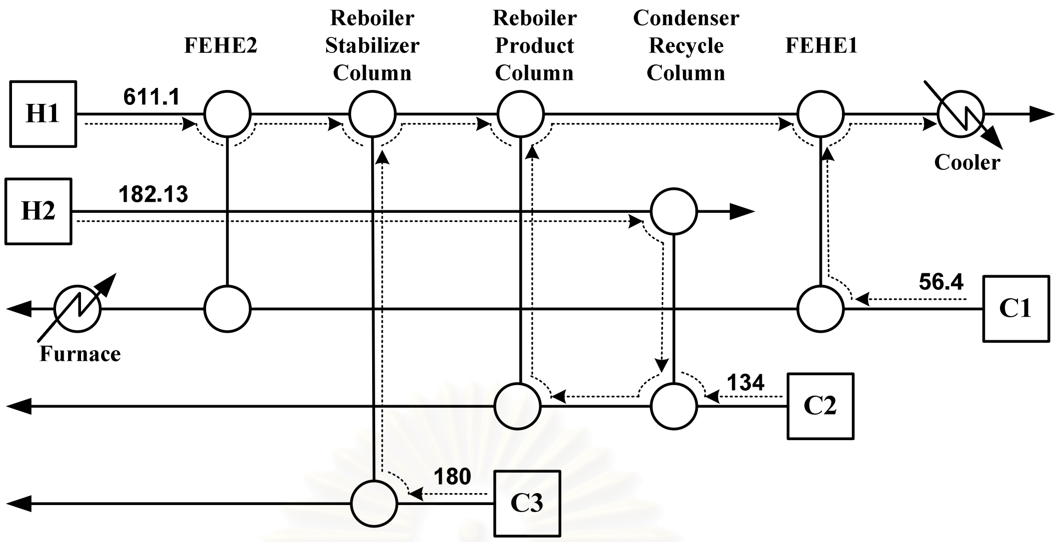


Figure 5.6 Design Heat Pathway at Minimum Heating Condition for Heat Exchanger Network HDA Alternative 5

3.2. Maximum Cooling Condition.

In order to achieve DMER for maximum cooling condition, the disturbance load is dissipated to heater as furnace. Therefore the appropriate heat pathway at minimum heating condition is designed as show in figure 5.7.

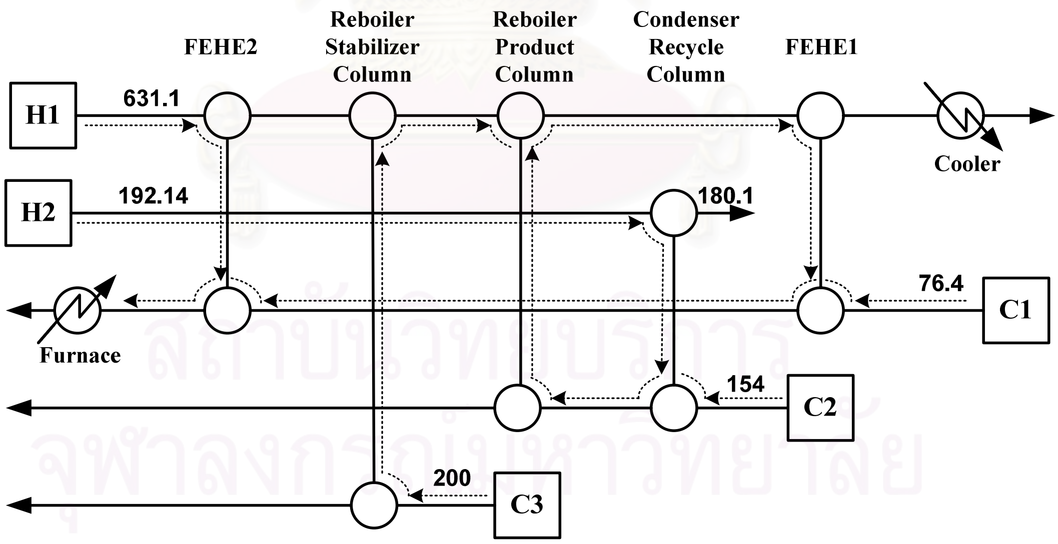


Figure 5.7 Design Heat Pathway at Maximum Cooling Condition for Heat Exchanger Network HDA Alternative 5

4. Estimated the Minimum Auxiliary Utility Units

The heat exchanger network of HDA process alternative 5 at normal operating condition is showed in figure 5.8.

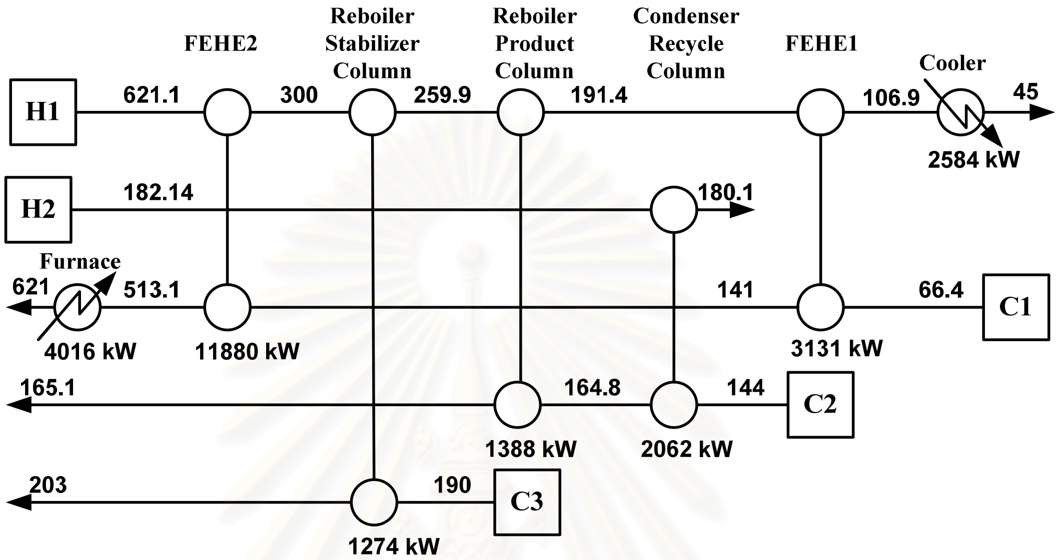


Figure 5.8 the Heat Exchanger Network of HDA Process Alternative 5 at Normal Operating Condition

4.1. Estimated the Minimum Auxiliary Heating Units

The minimum heating condition (see table 5.3) is introduced to heat exchanger network and then the disturbance load is dissipated to cooler as show in figure 5.6.

From the simulation results in figure 5.9, all the end of cold streams can achieve the target temperature except C2 stream. Although the outlet hot stream of reboiler product column decrease to 174.3 ^{o}C as maximum energy recovery (ΔT_{min} = 10 ^{o}C) however the end of C2 stream still can’t achieve the target temperature (165.1 ^{o}C).Therefore one auxiliary heating unit has to be employed at the end of C2 stream.

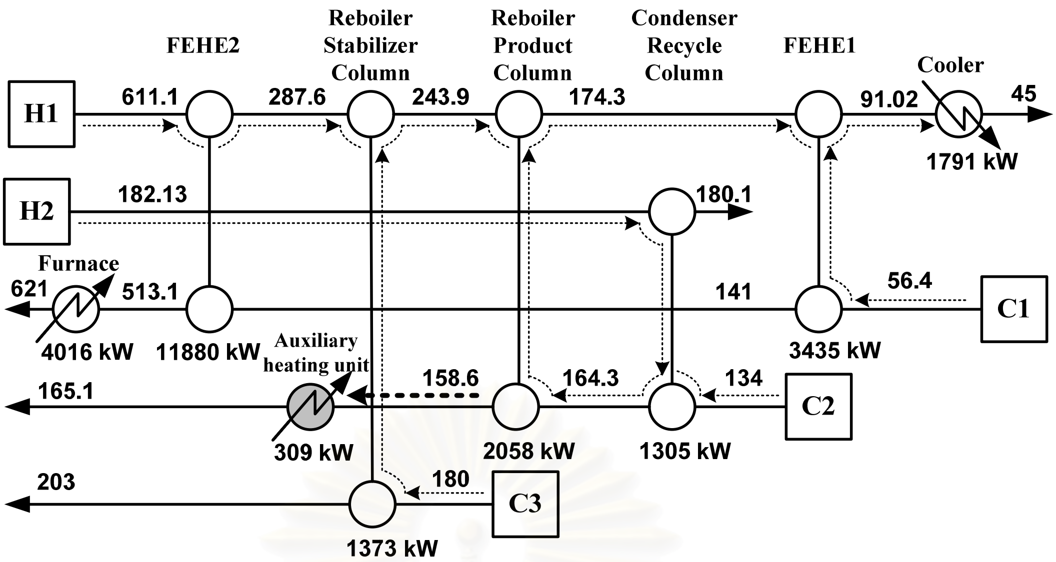


Figure 5.9 the Heat Exchanger Network of HDA Process Alternative 5 at Minimum Heating Condition

4.2. Estimated the Minimum Auxiliary Cooling Units

The maximum cooling condition (see table 5.4) is introduced to heat exchanger network and then the disturbance load is dissipated to heater as show in figure 5.7.

From the simulation results in figure 5.10, all the end of hot streams can achieve the target temperature. Therefore auxiliary cooling unit has not to be employed.

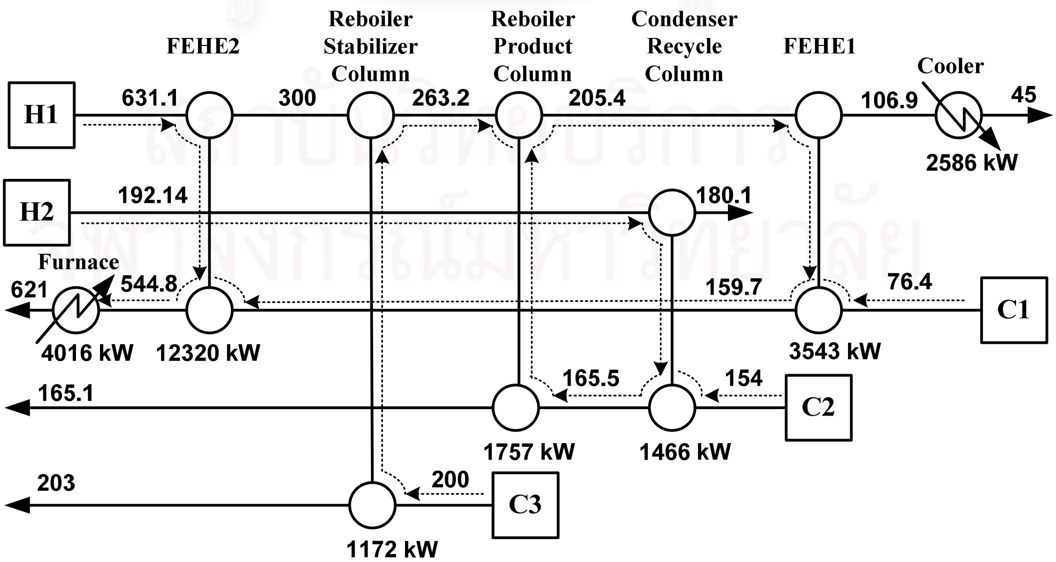


Figure 5.10 the Heat Exchanger Network of HDA Process Alternative 5 at Maximum Cooling Condition

The simulation results show that the HDA process alternative 5 needs only one auxiliary heating unit at the end of the cold-side steams coming from the bottom of product as auxiliary reboiler for product column see in figure 5.11.

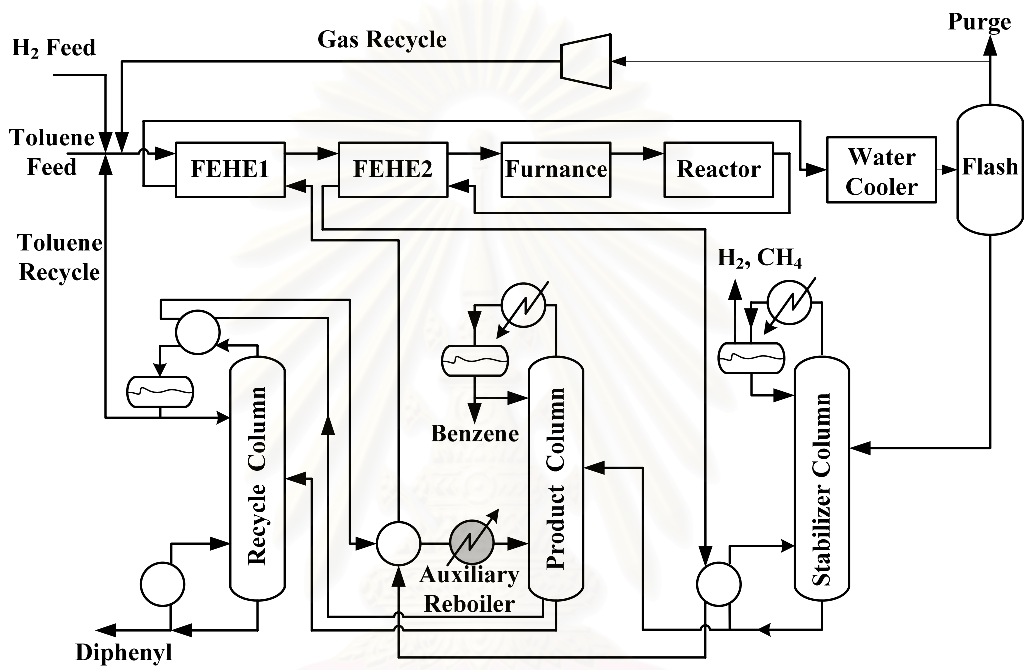


Figure 5.11 the HDA Process Alternative 5 with Minimum Auxiliary Utility Unit

5.2.1 Evaluated the Minimum Auxiliary Utility Units for HDA Process with Complex Energy Integration Alternative 6

Heat exchanger network of HDA process alternative 6 is presented in figure 5.12. H1 is the quenched reactor product stream, and H2 is the hot-side stream from the top tray of recycle column. C1 is the reactor feed steam prior to the furnace. C2, C3 and C4 are the cold-side steams coming from the bottom of product, stabilizer, and recycle columns, respectively.

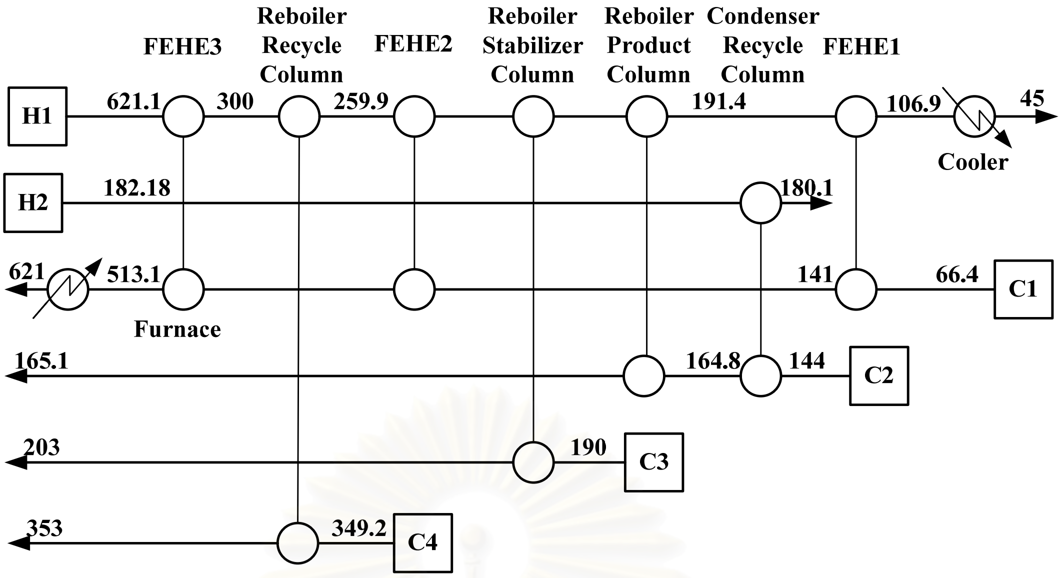


Figure 5.12 Heat Exchanger Network of HDA Process Alternative 6

1) Determined the expected disturbances and their magnitudes

Disturbance load for HDA process alternative 6 is defined that the magnitude of the disturbance variation is ± 10 ^{o}C from normal operating temperature (see table 5.5).

Table 5.5 the Magnitude of Disturbance Load for Heat Exchanger Network HDA Process Alternative 6

Stream	Disturbance load	
	Maximum inlet temperature (T_n+10^{o}C)	Minimum inlet temperature (T_n-10^{o}C)
H1	631.09	611.09
H2	192.18	182.15*
C1	76.42	56.42
C2	154.00	134.00
C3	200.00	180.00
C4	359.22	339.22

Note * the phase change occurs in this stream from gas to liquid when the inlet temperature decreases by 10 ^{o}C. This will make a huge disturbances load. In order to avoid this problem, the heat flow in this stream is decreased equal to the amount of sensible heat for 10 ^{o}C. Therefore this stream will decrease to 182.15 ^{o}C.

2. Design the Worst Case Conditions

2.1. Minimum Heating Condition.

This is a condition where the inlet temperatures of hot and cold streams are the lowest (see table 5.6).

Table 5.6 the Inlet Temperature of Heat Exchanger Network HDA Process Alternative 6 at Minimum Heating Condition

Stream	inlet temperature (T_{normal}-10^{o}C)
H1	611.09
H2	182.15
C1	56.42
C2	134.00
C3	180.00
C4	339.22

2.2. Maximum Cooling Condition.

This is a condition where the inlet temperatures of hot and cold streams are the highest (see table 5.7).

Table 5.7 the Inlet Temperature of Heat Exchanger Network HDA Process Alternative 6 at Maximum Cooling Condition

Stream	inlet temperature (T_{normal}+10^{o}C)
H1	631.09
H2	192.18
C1	76.42
C2	154.00
C3	200.00
C4	359.22

3. Design Heat Pathways for the Worst Case Conditions

3.1. Minimum Heating Condition.

In order to achieve DMER for minimum heating condition, the disturbance load is dissipated to cooler. Therefore the appropriate heat pathway at minimum heating condition is designed as show in figure 5.13.

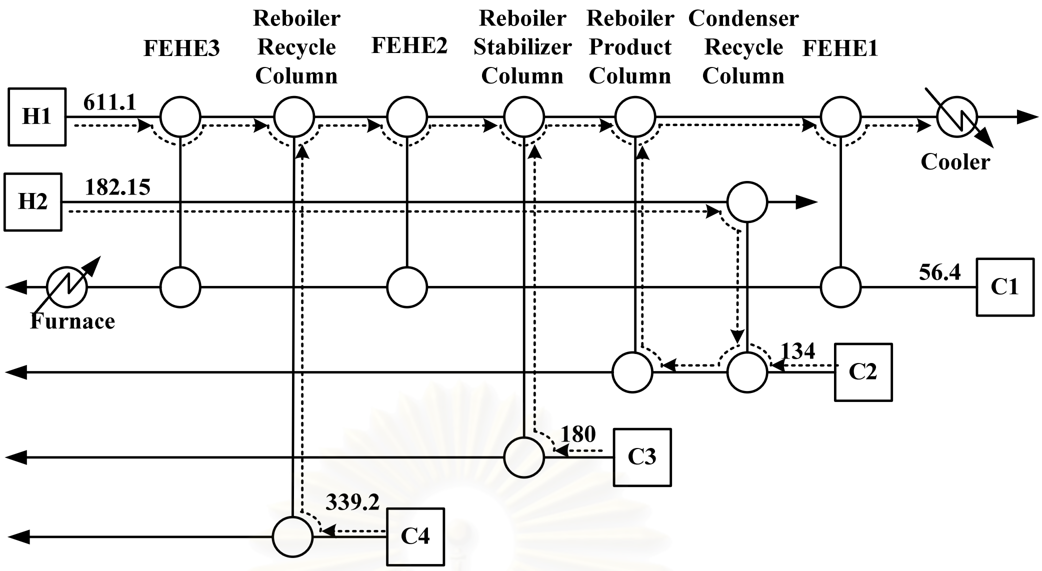


Figure 5.13 Design Heat Pathway at Minimum Heating Condition for Heat Exchanger Network HDA Alternative 6

3.2. Maximum Cooling Condition.

In order to achieve DMER for maximum cooling condition, the disturbance load is dissipated to heater. Therefore the appropriate heat pathway at maximum cooling condition is designed as show in figure 5.14.

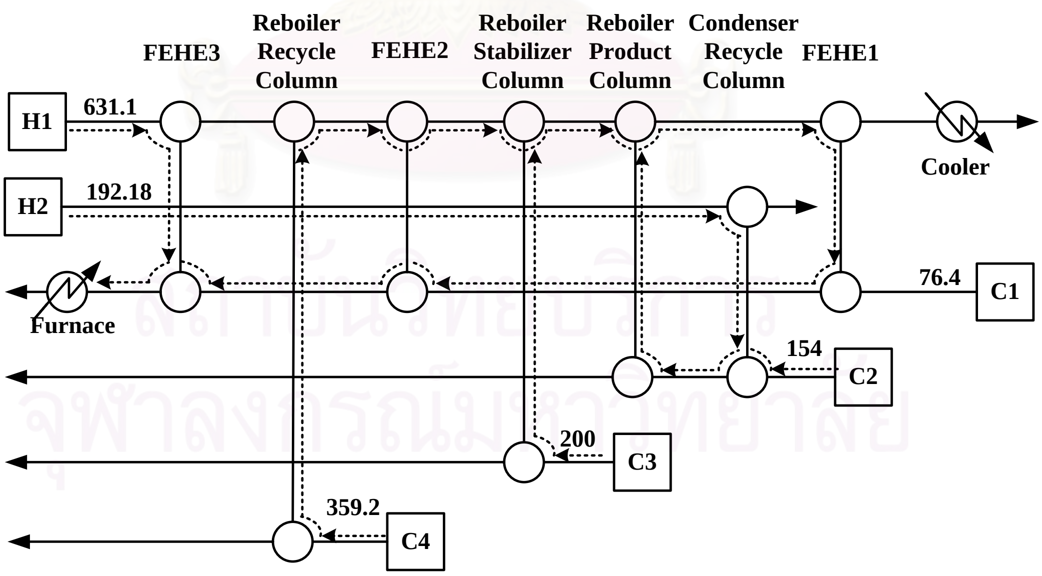


Figure 5.14 Design Heat Pathway at Maximum Cooling Condition for Heat Exchanger Network HDA Alternative 6

4. Estimated the minimum auxiliary utility units

The heat exchanger network of HDA process alternative 6 at normal operating condition is showed in figure 5.15.

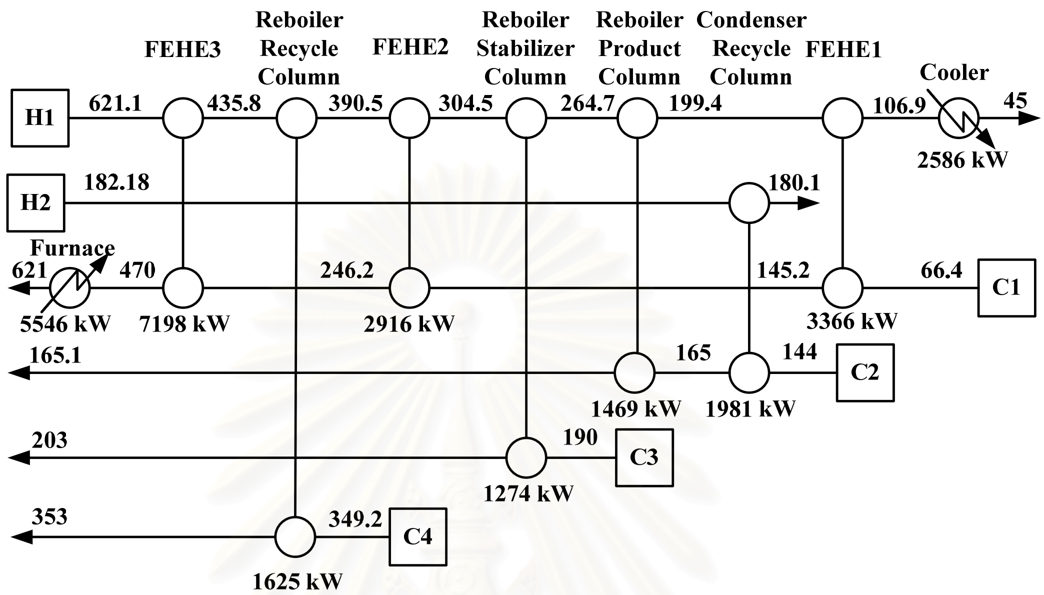


Figure 5.15 the Heat Exchanger Network of HDA Process Alternative 6 at Normal Operating Condition

4.1. Estimated the Minimum Auxiliary Heating Units

The minimum heating condition (see table 5.6) is introduced to heat exchanger network and then the disturbance load is dissipated to cooler as show in figure 5.13.

From the simulation results in figure 5.16, all the end of cold streams can achieve the target temperature except C2 stream. Although the outlet hot stream of reboiler product column decrease to 174.4 ^{o}C as maximum energy recovery (ΔT_{min} = 10 ^{o}C) however the end of C2 stream still can’t achieve the target temperature (165.1 ^{o}C).Therefore one auxiliary heating unit has to be employed at the end of C2 stream.

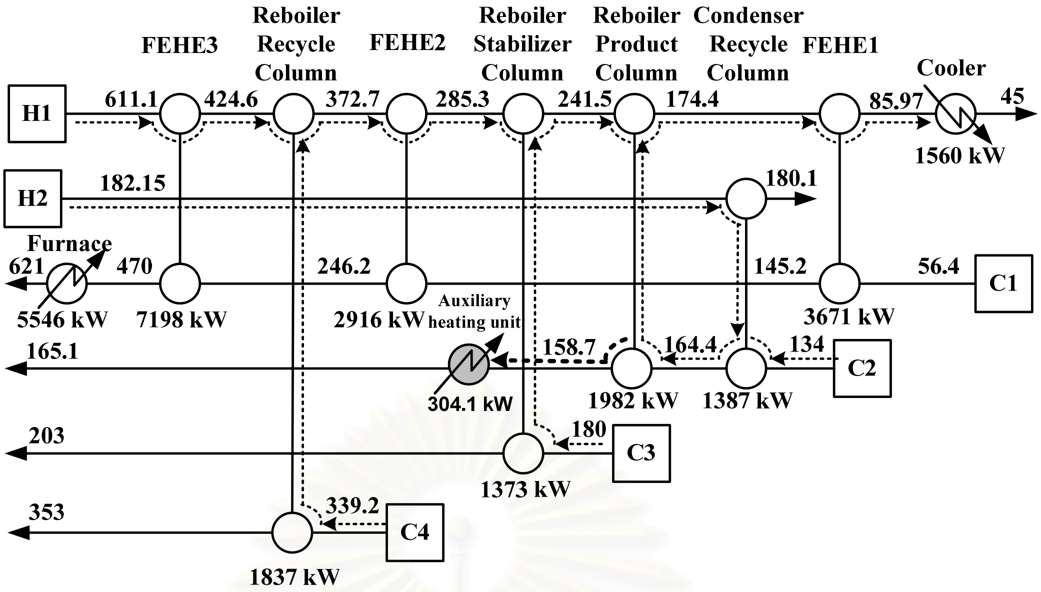


Figure 5.16 the Heat Exchanger Network of HDA Process Alternative 6 at Minimum Heating Condition

4.2. Estimated the Minimum Auxiliary Cooling Units

The maximum cooling condition (see table 5.7) is introduced to heat exchanger network and then the disturbance load is dissipated to heater as show in figure 5.14.

From the simulation results in figure 5.17, all the end of hot and cold stream can achieve the target temperature therefore auxiliary cooling unit have not to be employed.

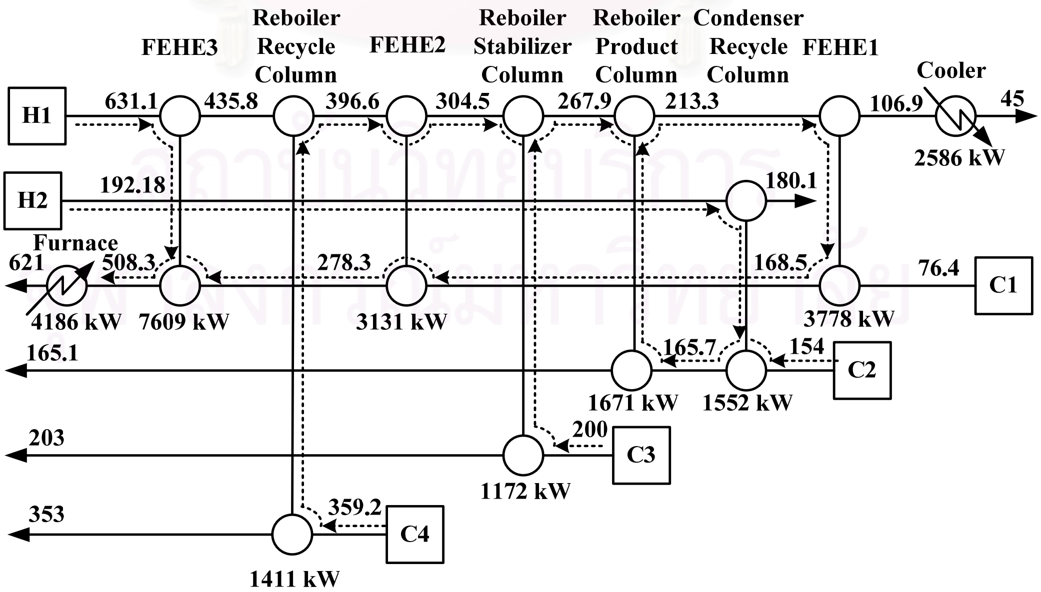


Figure 5.17 the Heat Exchanger Network of HDA Process Alternative 6 at Maximum Cooling Condition

The simulation results show that the HDA process alternative 6 needs only one auxiliary heating unit at the end of the cold-side steams coming from the bottom of product as auxiliary reboiler for product column see in figure 5.18.

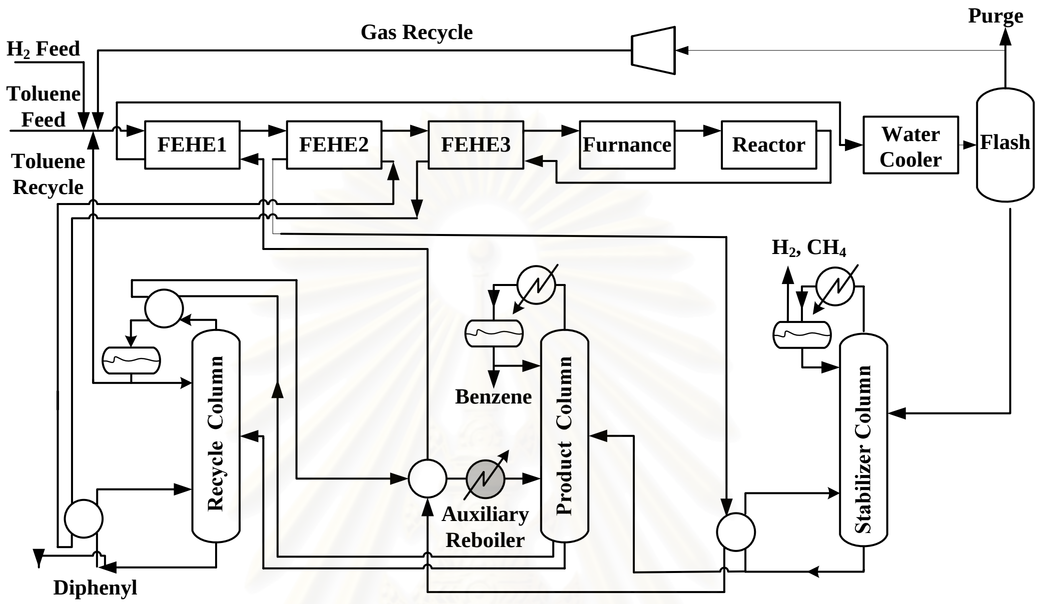


Figure 5.18 the HDA Process Alternative 6 with Minimum Auxiliary Utility Unit

5.3 Comparison the Number of Auxiliary Utility Unit Between our Strategy and Lyben (1998)'s Suggestion for HDA Process with Highly Heat-integrated Process Alternatives 5 and 6

HDA process alternative 6 is highly heat-integrated process. In order to solve some of the control difficulties for highly heat-integrated process, Luyben (1998) suggested that the auxiliary utility unit should be employed to improve controllability. In figure 5.20 we show HDA process alternative 6 that uses three auxiliary reboilers and one auxiliary condenser. Using the same suggestion by Luyben for HDA process alternative 5, we found that two auxiliary reboilers and one auxiliary condenser are employed as show in figure 5.19.

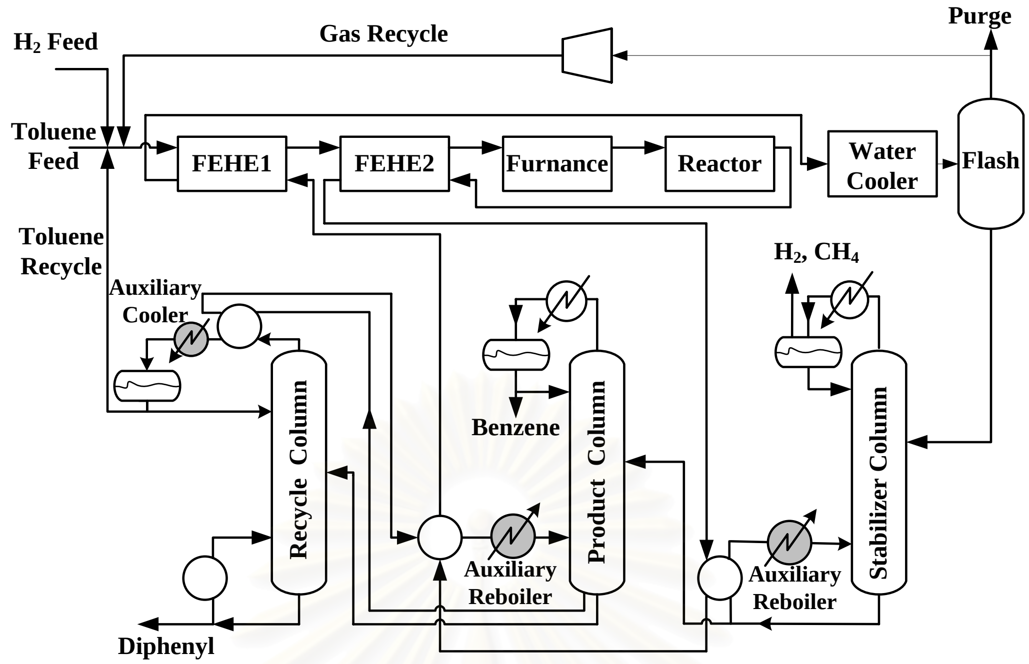


Figure 5.19 HDA Process Alternative 5 with two Auxiliary Reboilers and one Auxiliary Cooler

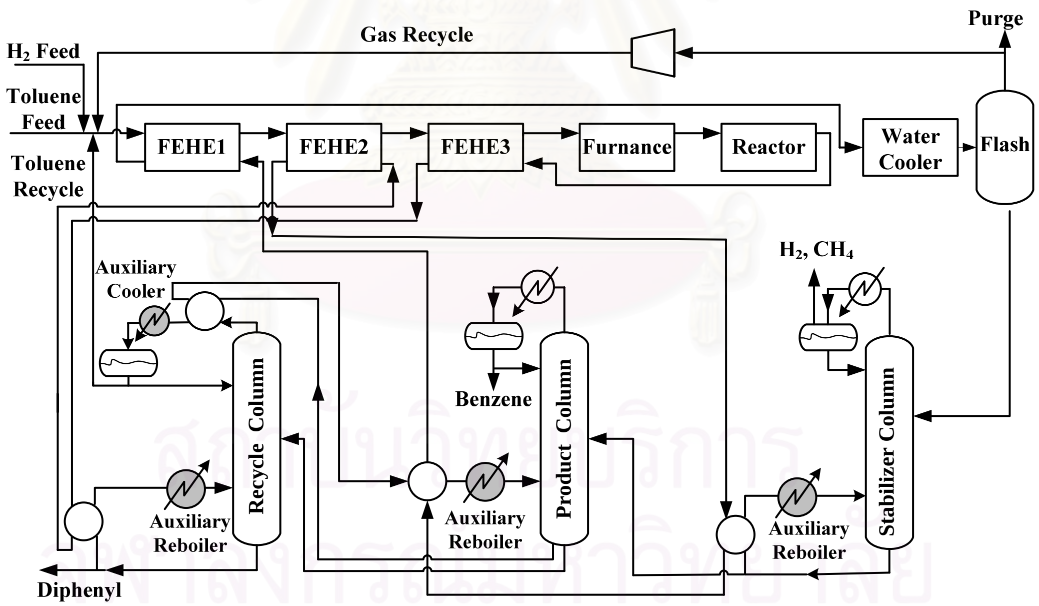


Figure 5.20 HDA Process Alternative 6 with three Auxiliary Reboilers and one Auxiliary Cooler

For our strategy, we found that only one auxiliary reboiler required to guarantee workable heat-integrated HDA process alternatives 5 and 6. The comparison the number of auxiliary utility unit between our strategy and Lyben for

HDA process with highly heat-integrated process alternatives 5 and 6 is summarized in table 5.8.

Table 5.8 Comparison the Number of Auxiliary Utility Unit between our Strategy and Lyben for HDA Process with Highly Heat-integrated Process Alternatives 5 and 6

auxiliary utility unit	Number of auxiliary utility unit			
	HDA alternative 5		HDA alternative 6	
	Lyben	Our strategy	Lyben	Our strategy
auxiliary reboiler	2	1	3	1
auxiliary cooler	1	-	1	-
total	3	1	4	1

In table 5.8, we found that our strategy can decrease number of auxiliary utility unit by 2 and 3 for HDA alternatives 5 and 6, respectively.

The new plantwide control structure for HDA alternatives 5 and 6 will propose in next charter. The control performance of HDA alternatives 5 and 6 with minimum auxiliary utility unit will be compared with HDA alternatives 5 and 6 with auxiliary utility unit suggested by Luyben.

Figures 5.20 and 5.21 show Hysy flowsheet of the steady state modeling of highly heat integrated HDA Process alternative 5 with three and minimum auxiliary utility unit, respectively.

Figures 5.22 and 5.23 show Hysy flowsheet of the steady state modeling of highly heat integrated HDA process alternative 6 with four and minimum auxiliary utility unit, respectively.

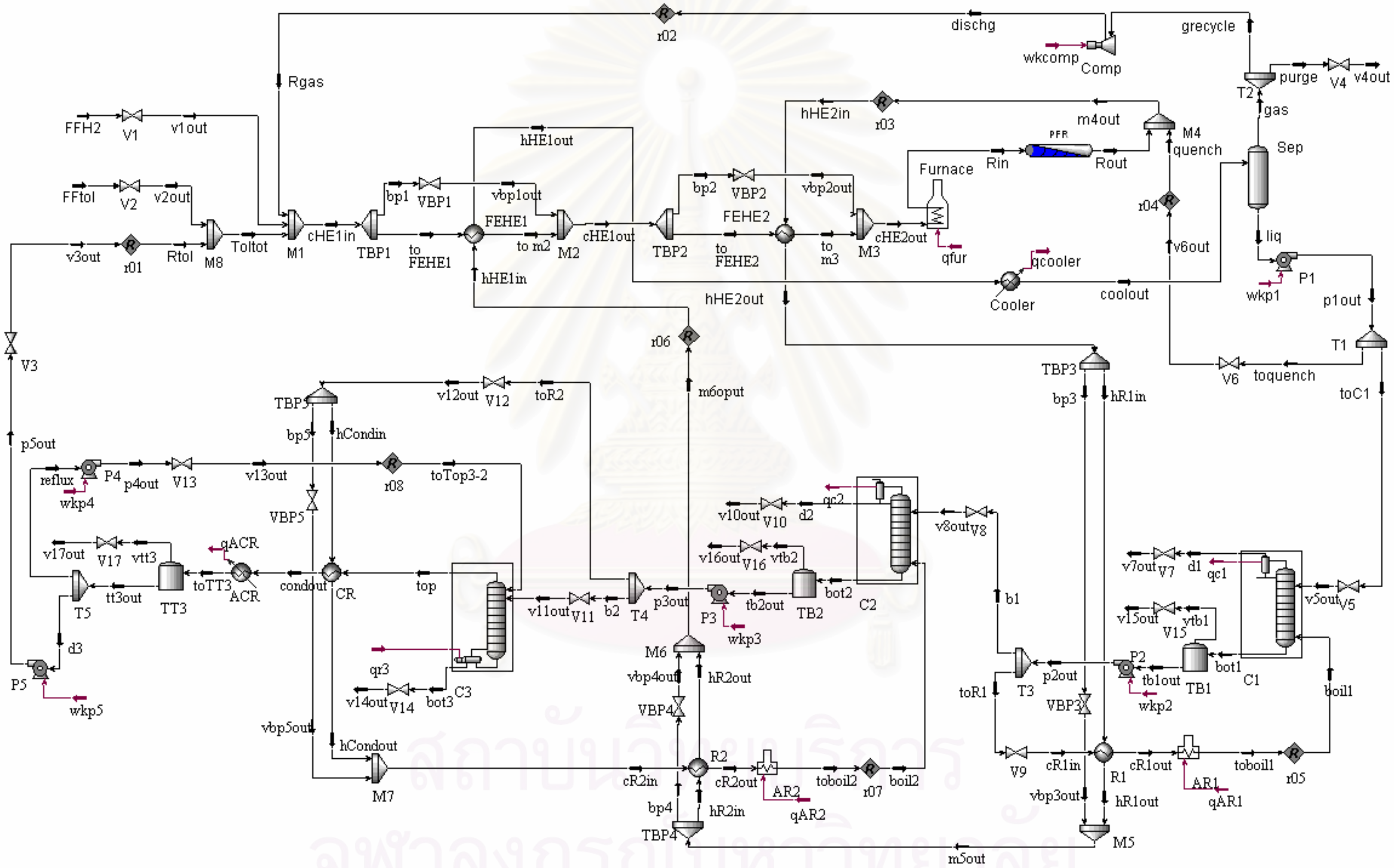


Figure 5.21 Hysy Flowsheet of the Steady State Modeling of HDA Process Alternative 5 with three Auxiliary Utility Unit

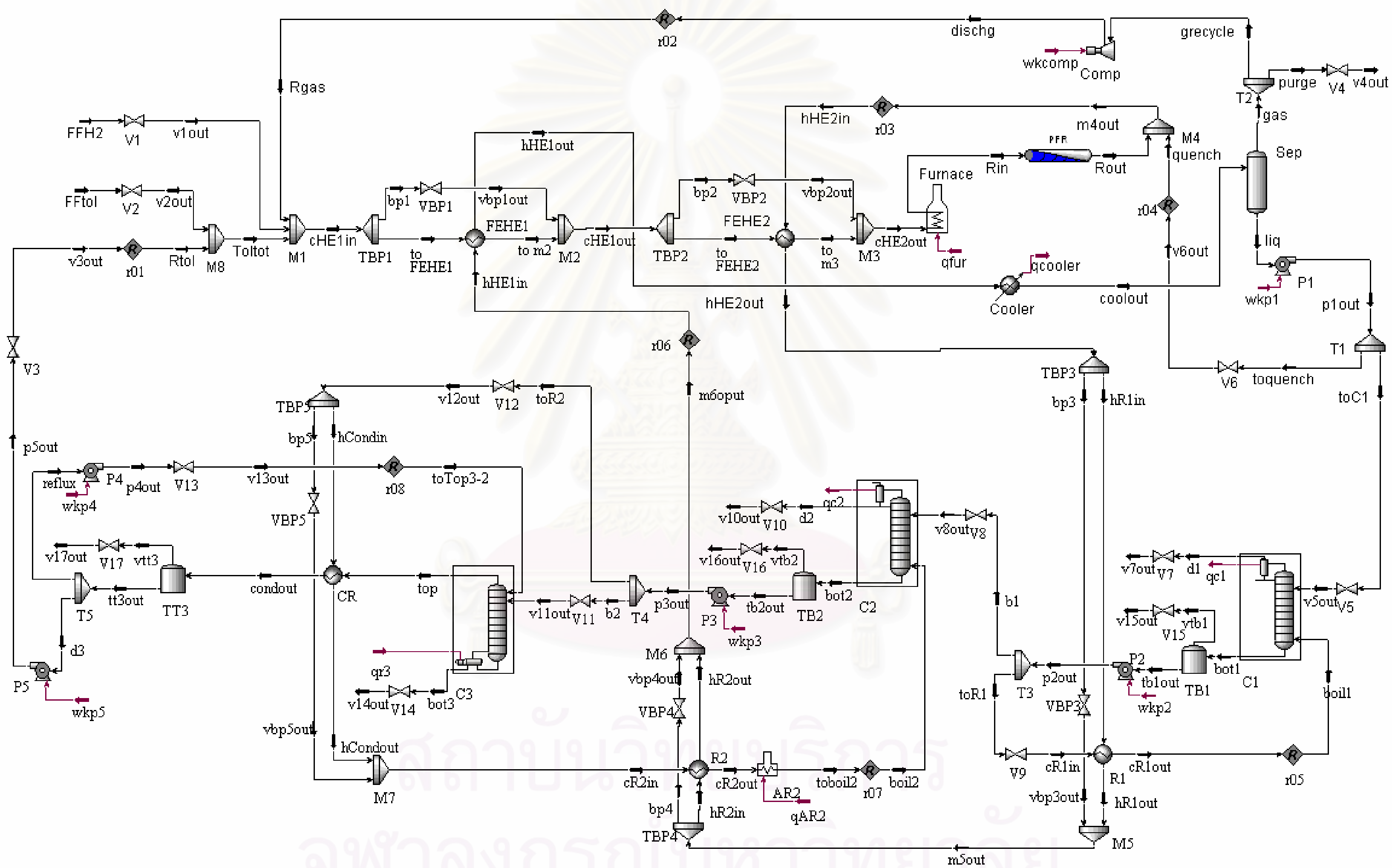


Figure 5.22 Hysy Flowsheet of the Steady State Modeling of HDA Process Alternative 5 with Minimum Auxiliary Utility Unit

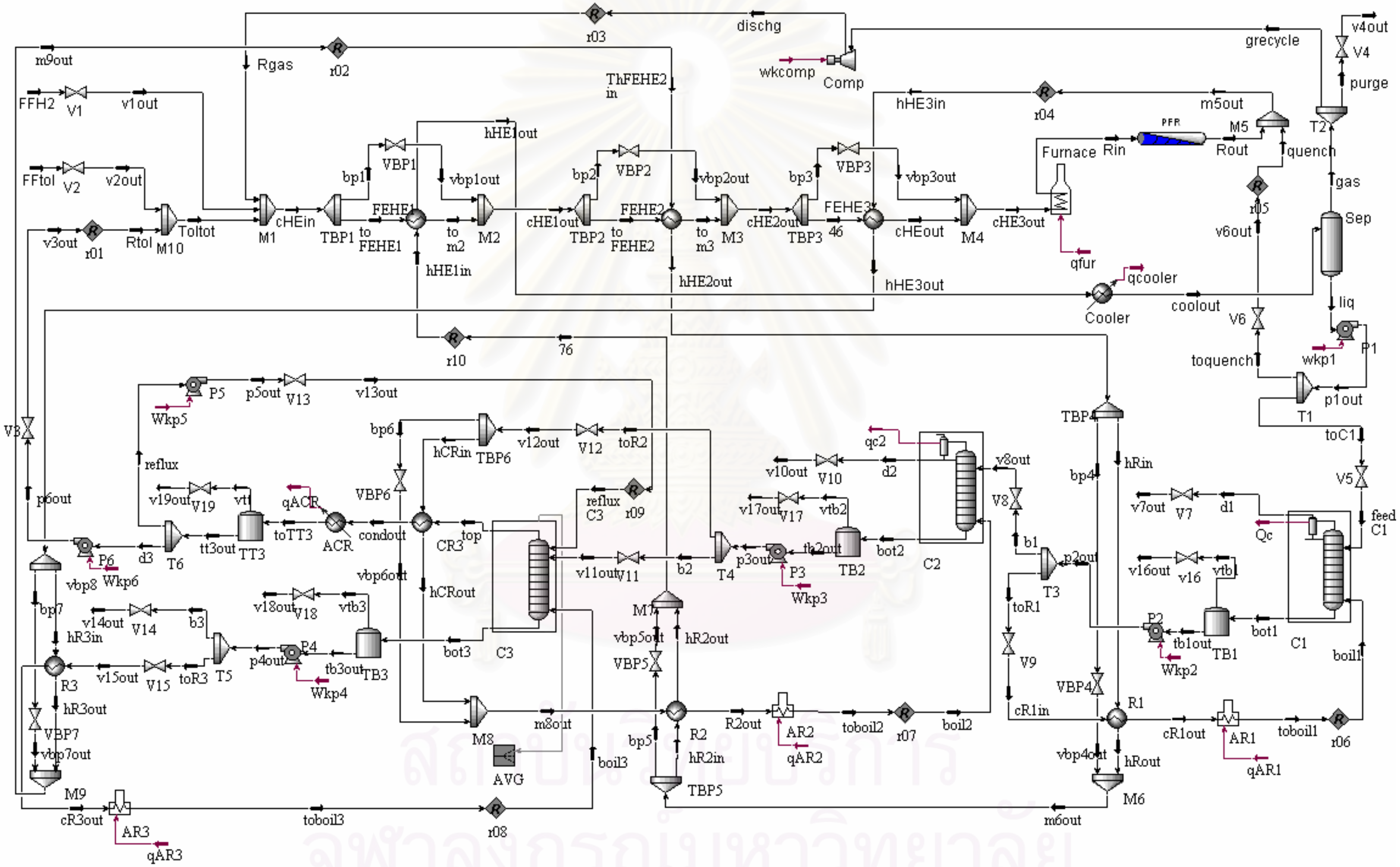


Figure 5.23 Hysy Flowsheet of the Steady State Modeling of HDA Process Alternative 6 with four Auxiliary Utility Unit

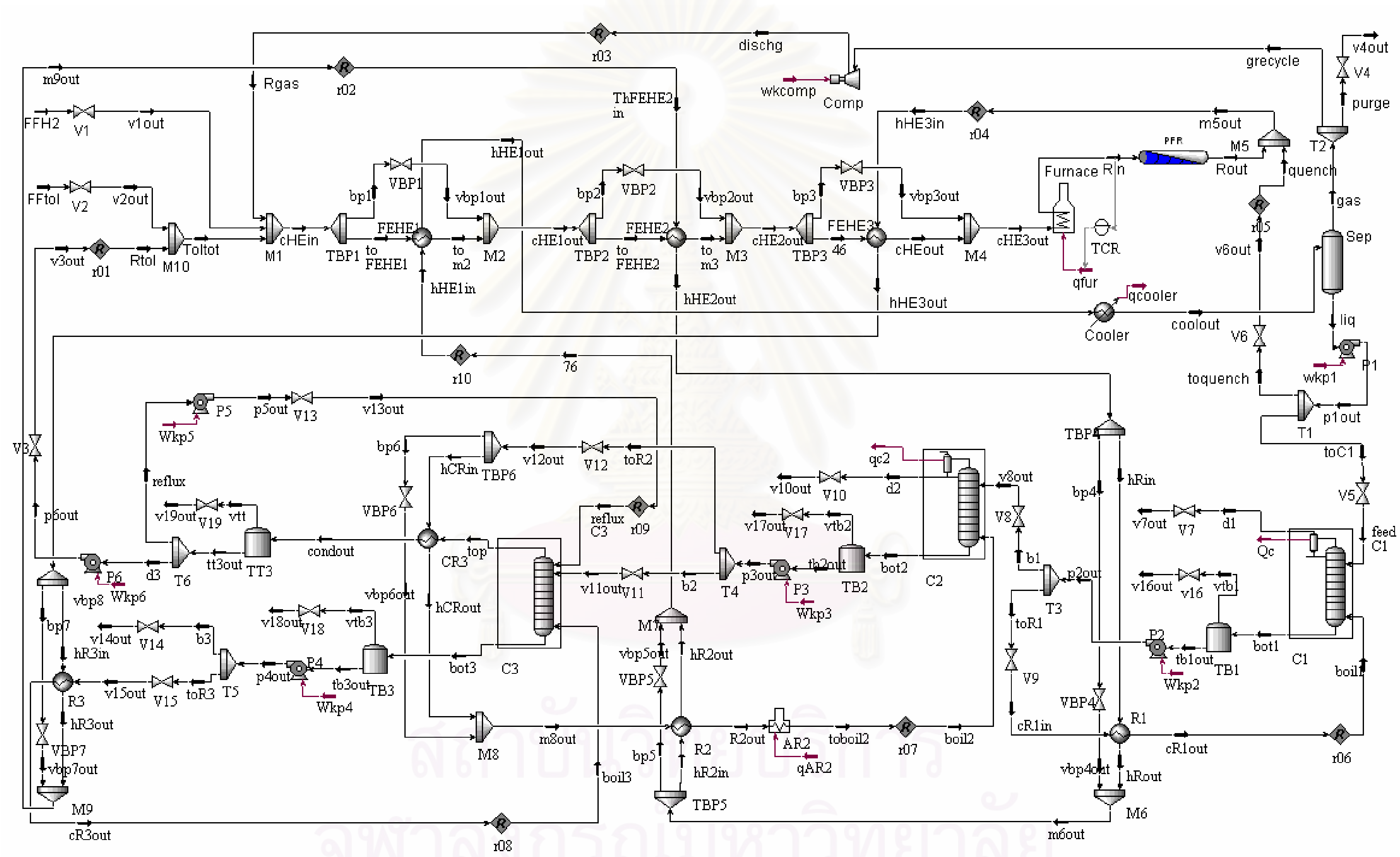


Figure 5.24 Hysy Flowsheet of the Steady State Modeling of HDA Process Alternative 6 with Minimum Auxiliary Utility Unit

CHAPTER VI

CONTROL STRUCTURES DESIGN
AND DYNAMIC SIMULATION

Maintaining the plant energy and mass balances are the essential task of plantwide for a complex plant consists of recycle streams and energy integration when the disturbance load come through the process. The control system is needed to reject loads and regulate an entire process into a design condition to achieve its objectives therefore our purpose of this chapter is to present the new control structures of highly energy integrated HDA process (alternatives 5 and 6) with auxiliary utility units that suggested by Luyben (1998) and with minimum auxiliary utility units that is proposed in previous chapter. Moreover, the three new designed control structures are also compared between minimum auxiliary utility units and auxiliary utility units that suggested by Luyben based on rigorous dynamic simulation by using the commercial software HYSYS.   .

6.1 Plantwide Control Design Procedure for HDA Process Alternatives 5 and 6

The plantwide control structures in the HDA process alternatives 5 and 6 are designed based on the heuristic design procedure given by Luyben et al.(1999) and discussed below.

HDA

Step 1 Establish Control Objectives.

For this process, the essential is to produce pure benzene while minimizing yield losses of hydrogen and diphenyl. The reactor effluent gas must be quenched to 621.1 ^oC to prevent coking and byproduct formation. The design a control structures for process associate with energy integration can be operated well.

Step 2 Determine Control Degree of Freedom.

There are 25 control degrees of freedom in HDA alternative 5. They include; two fresh feed valves for hydrogen and toluene; two bypass valve for two feed effluent heat exchangers (FEHE); purge valve; separator base and overhead valves, cooler cooling water valve; liquid quench valve; furnace fuel valve; stabilizer column bottoms, reboiler bypass valve, cooling water, and vapor product valves; product column bottoms, reboiler bypass valve, reflux, distillate, and cooling water valves; and recycle column bottoms, reboiler bypass valve, reflux, distillate, and cooling water valves. For HDA alternative 6, there have three feed effluent heat exchangers therefore the control degrees of freedom is 26.

Step 3 Establish Energy management system.

The reactor operates adiabatically, so for a given reactor design the exit temperature depends upon the heat capacities of the reactor gases, reactor inlet temperature, and reactor conversion. Heat from the adiabatic reactor is carried in the effluent stream and is not removed from the process until it is dissipated to utility in the separator cooler. Energy management of reaction section is handled by controlling the inlet and exit streams temperature of the reactor. Reactor inlet temperature must be controlled by adjusting fuel to the furnace and reactor exit temperature must be controlled by quench to prevent the benzene yield decreases from the side reaction.

In addition, a selective controller with low selector switch (LSS) is employed at FEHE no.1 to select an appropriate heat pathway to carry the associated load to a cooling or heating utility unit. The dynamic maximum energy recovery (DMER) can achieve.

In order to keep supplied heat load and guarantee workable HDA process alternative 5 with minimum auxiliary utility units, the bypass valve of FEHE no.2 is used to control the inlet hot stream of stabilizer column reboiler. The same concept is also applied to HDA process alternative 6, the bypass valve of FEHE no.2 is used to control the inlet hot stream of reboiler in stabilizer column(R1) and the bypass valve of FEHE no.3 is used to control the inlet hot stream of reboiler in recycle column(R3).

The auxiliary utility units are employed for highly heat integrated process. The minimum heat load of auxiliary utility unit is optimum operation therefore the split range control is used in this work to achieve optimum operation.

Step 4 Set Production Rate.

Many control structures, there are not constrained to set production either by supply or demand. Considering of the kinetics equation is found that the only three variables could be potentially dominant for the reaction rate; pressure, temperature and toluene concentration (limiting agent). Pressure is not a variable choice for production rate control because of the compressor has to operate at maximum capacity for yield purposes. This gives us two viable options: change reactor inlet temperature or inlet toluene composition. We select toluene composition. This allows us to control the total flow of toluene to reactor (recycle plus fresh). Fresh toluene feed flow is used to control toluene inventory reflected in the recycle column overhead receiver level as an indication of the need for reactant makeup. Controlling the total toluene flow sets the reactor composition indirectly.

Step 5 Control Product Quality and Handle Safety, Operational, and Environmental Constraints.

The distillate stream from the product column is salable benzene. Benzene quality can be affected primarily by two components, methane and toluene. Any methane that leaves in the bottoms of the stabilizer column contaminates the benzene product. The separation in the stabilizer column is used to prevent this problem by using a temperature to set column stream rate (boil-up). Toluene in the overhead of the product column affect to benzene quality. Benzene purity can be controlled by manipulating the column steam rate (boil up) to maintain temperature in the column.

Step 6 Control Inventories and Fix a Flow in Every Recycle Loop.

The recycle flow of toluene should be fixed. Four pressures must be controlled: in the three distillation columns and in the gas loop. The pressure in the gas loop is controlled with the fresh hydrogen feed flow since it indicates hydrogen inventory in the gas recycle loop. In the stabilizer column, vapor product flow is the most direct manipulator to control pressure. In the product and recycle columns, pressure control can be achieved by manipulating cooling water flow to regulate overhead condensation rate. In HDA process alternatives 5 and 6, since heat-integrated distillation system was used for both the product and recycle columns, the cold inlet stream of process-to-process-heat-exchanger (condenser/reboiler) is bypassed and manipulated to control pressure in the recycle column.

Seven liquid levels are in the process: separator and two (base and overhead receiver) in each column. The most direct way to control separator level is with the liquid flow to the stabilizer column. Then stabilizer column overhead receiver level is controlled with cooling water flow and base level is controlled with bottoms flow. In the product column and recycle column, distillate flow controls overhead receiver level and bottoms flow controls base level.

Step 7 Check Component Balances.

Methane is purged from the gas recycle loop to prevent it from accumulating, and its composition can be controlled with purge flow. Diphenyl is removed in the bottoms stream from the recycle column, where steam flow control the temperature. The inventory of benzene is accounted for via temperature and overhead receiver level control in the product column. Toluene inventory is accounted for via level control in the recycle column overhead receiver. Purge flow and gas-loop pressure control account for hydrogen inventory.

Step 8 Control Individual Unit Operations.

In this step, the rest degrees of freedom are assigned for control loops within individual units. Cooling water flow to the cooler controls the separator temperature. Reflux to the stabilizer, product, and recycle columns can be flow controlled because there is no requirement at the unit operations level to do anything beyond this.

Step 9 Optimize Economics or Improve Dynamic Controllability.

The basic regulatory strategy has now been established. Some freedom is used to select several controller set points to optimize economics and plant performance. Such as, the set point for the methane composition controller in the gas recycle loop must balance the trade-off between yield loss and reactor performance. Reflux flows to the stabilizer, product, and recycle columns must be determined based upon column energy requirement and potential yield losses of benzene (in the overhead of the stabilizer and recycle columns) and toluene (in the base of the recycle column).Since the separations are easy, in this system economics indicate that reflux flows would probably be constant.

6.2 Design of Plantwide Control for HDA Process Alternative 5 with three Auxiliary Utility Units and Minimum Auxiliary utility units

The three new control structures are designed for HDA process alternative 5 with three auxiliary utility units that is suggested by Luyben and the same three new design is also employed to HDA process alternative 5 with minimum auxiliary utility units to compare dynamic behavior.

6.2.1 Control Structure 1 (CS1) for HDA Process Alternative 5 with three Auxiliary Utility Units

This control structure is shown in Figure 6.1 and the controller parameters are given in table 6.1. For control structure, the all bypass valve of 2 feed effluent heat exchangers (FEHEs) are employed on cold side to control outlet temperature. A selector controller with low selector switch (LSS) is employed at FEHE1 to select an appropriate heat pathway. This control system involves one manipulated variable and two controlled variables and works as follows: The hot outlet temperature of FEHE1 is controlled at its nominal set point by manipulating bypass valve of FEHE1. At the same time, the cold outlet temperature of FEHE1 should not be allowed to drop below a lower limit value, which is necessary to keep the furnace duty at a good level. Whenever the cold outlet temperature of FEHE1 drops below the allowable limit due to, for example, a disturbance load entering the process, the LSS switches the control action to the cold outlet temperature control, and decrease the percent opening of bypass valve at FEHE1. 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. Whenever the cold outlet temperature of FEHE1 increases above a lower limit, i.e. a desired-condition during operation, due to the disturbance load entering the process, the LSS switches the control action to the hot outlet temperature of FEHE1. Consequently, the hot outlet temperature of FEHE1 will drop to its normal temperature and the cold outlet temperature of FEHE1 will be further increased, so the furnace duty will also be decreased.

Since the hot reactor product is used to drive reboiler in the two columns (R1 and R2), part of this stream is bypassed and manipulated to control the tray temperatures in the two columns. The hot outlet temperatures of FEHE2 (the temperature at the entrance of the reboilers at stabilizer column) are controlled by manipulating the bypass valves of FEHE2 in odder to prevent the propagation of thermal disturbance to the separation section.

Since the three auxiliary utility units consist of two auxiliary reboilers at stabilizer column and product column and one auxiliary condenser at recycle column. There are employed next to process to process heat exchanger as reboiler or condenser unit as show in figure 6.1. The optimum operation would be to minimize the heat load of auxiliary utility units. One way to do this is the using split range control. This control system involves two manipulated variable (the bypass valve of process to process heat exchanger and heat load of auxiliary utility unit) and one controlled variables (temperature or pressure) and works as for example: if the decreasing disturbance loads of hot stream occurs at reboiler and then the percent opening of bypass valve at reboiler will decrease in order to increase heat transfer at reboiler until the target tray temperature achieves. The heat load of the auxiliary utility units will be used when the bypass valve is full close but the target tray temperature still can't achieves the target tray temperature. Therefore the split range control is used for the tray temperature control in stabilizer column and product column by manipulating the bypass valve of reboiler and the heat load of the auxiliary reboiler. Beside for recycle column, the split control is also applied to control column pressure by manipulating the bypass valve of condenser (CR) and the heat load of the auxiliary condenser.

Since the temperature profile in the recycle column is very sharp because of temperature changes from tray to tray. This means that the process gain is very large when a single tray temperature is controlled. The standard solution for this problem is to use an average (AVG) temperature of several trays instead of a single tray (Luyben, 2002).

A heat exchanger (i.e. as a heat source or a heat sink) is artificially installed in the hot-side stream (i.e. the exchanger X1 in fig. 6.1) in order to make the disturbance loads of the hot stream (i.e. the hot reactor product). Note that, this exchanger is not used in the real plant, and the temperature controller TCX1 is set to be "off" whenever it is not used to make the disturbances.

The control structure and controller parameter are given in table 6.1. P controllers are used for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.2.2 Control Structure 2 (CS2) for HDA Process Alternative 5 with three Auxiliary Utility Units

This control structure is shown in figure 6.2 and the controller parameter is given in table 6.2. The major loops in this control structure are the same as CS1 except for control loop at FEHE. The all bypass valve of 2 FEHEs will be employed on hot side to control outlet temperature.

6.2.3 Control Structure 3 (CS3) for HDA Process Alternative 5 with three Auxiliary Utility Units

This control structure is shown in Figure 6.3 and the controller parameter is given in table 6.3. The major loops in this control structure are the same as CS1 except for temperature control in product distillation column. The temperature control in product distillation column is two point controls as the tray 12 and tray 17 temperature controls.

6.2.4 Control Structure 1 (CS1) for HDA Process Alternative 5 with Minimum Auxiliary Utility Units

The same control structure 1 for HDA process alternative 5 with three auxiliary utility units is employed to HDA process alternative 5 with minimum auxiliary utility units (figure 6.4). Since the number of minimum auxiliary utility unit for HDA process alternative 5 is only one unit therefore the split range control will be employed only one for product column to control the tray temperature. The control structure and controller parameter are given in table 6.4.

6.2.5 Control Structure 2 (CS2) for HDA Process Alternative 5 with Minimum Auxiliary utility units

The same control structure 2 for HDA process alternative 5 with three auxiliary utility units is employed to HDA process alternative 5 with minimum auxiliary utility units (figure 6.5). Since the number of minimum auxiliary utility unit for HDA process alternative 5 is only one unit therefore the split range control will be employed only one for product column to control the tray temperature. The control structure and controller parameter are given in table 6.5.

6.2.6 Control Structure 3 (CS3) for HDA Process Alternative 5 with Minimum Auxiliary utility units

The control structure 3 for HDA process alternative 5 with three auxiliary utility units is employed to HDA process alternative 5 with minimum auxiliary utility units (figure 6.6). Since the number of minimum auxiliary utility unit for HDA process alternative 5 is only one unit therefore the split range control will be employed only one for product column to control the tray temperature. The control structure and controller parameter are given in table 6.1.

Table 6.1 Control Structure and Controller Parameter for HDA Process Alternative 5 with three Auxiliary Utility Units: Control Structure 1

controller	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.399	0.404	0.0897
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.293	0.423	0.094
TCS	separator temperature	cooler duty (qcooler)	PID	0.898	0.359	0.0789
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.736	0.388	0.0863
TCE1h	cooler inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.730	0.379	0.0842
LSS	output of TCE1c and TCE1h	FEHE1 bypass cold stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	0.829	0.522	0.116
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3) and auxiliary reboiler 1 (AR1) duty	PID	2.57	1.95	0.433
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	5.34	4.19	0.930
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5) and auxiliary condenser(ACR) duty	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 reboiler duty (qr3)	PID	0.196	5.03	1.12
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum	toluene recycle valve (V3)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

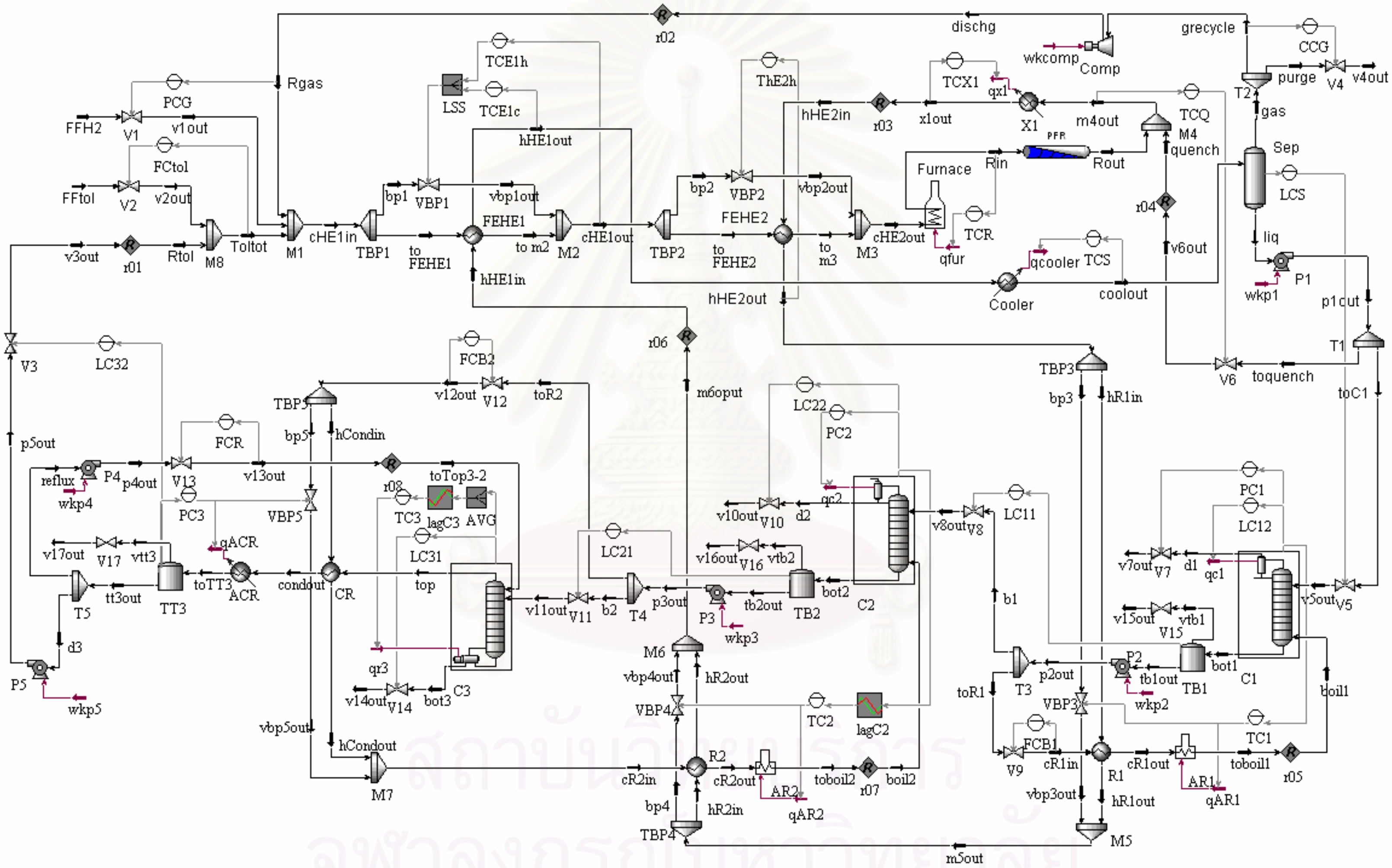


Figure 6.1 Control Structure 1 (CS1) for HDA Process Alternative 5 with three Auxiliary Utility Units

Table 6.2 Control structure and controller parameter for HDA Process Alternative 5 with three Auxiliary Utility Units: Control Structure 2

controller name	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.399	0.404	0.0897
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.293	0.423	0.094
TCS	separator temperature	cooler duty (qcooler)	PID	0.898	0.359	0.0789
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass hot stream valve (VBP1)	PID	0.751	0.378	0.0839
TCE1h	cooler inlet temperature	FEHE1 bypass hot stream valve (VBP1)	PID	0.725	0.379	0.0843
LSS	output of TCE1c and TCE1h	FEHE1 bypass hot stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass hot stream valve (VBP2)	PID	0.0815	0.260	0.0578
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3) and auxiliary reboiler 1 (AR1) duty	PID	2.57	1.95	0.433
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	5.34	4.19	0.930
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5) and auxiliary condenser(ACR) duty	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 reboiler duty (qr3)	PID	0.196	5.03	1.12
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum	toluene recycle valve (V3)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

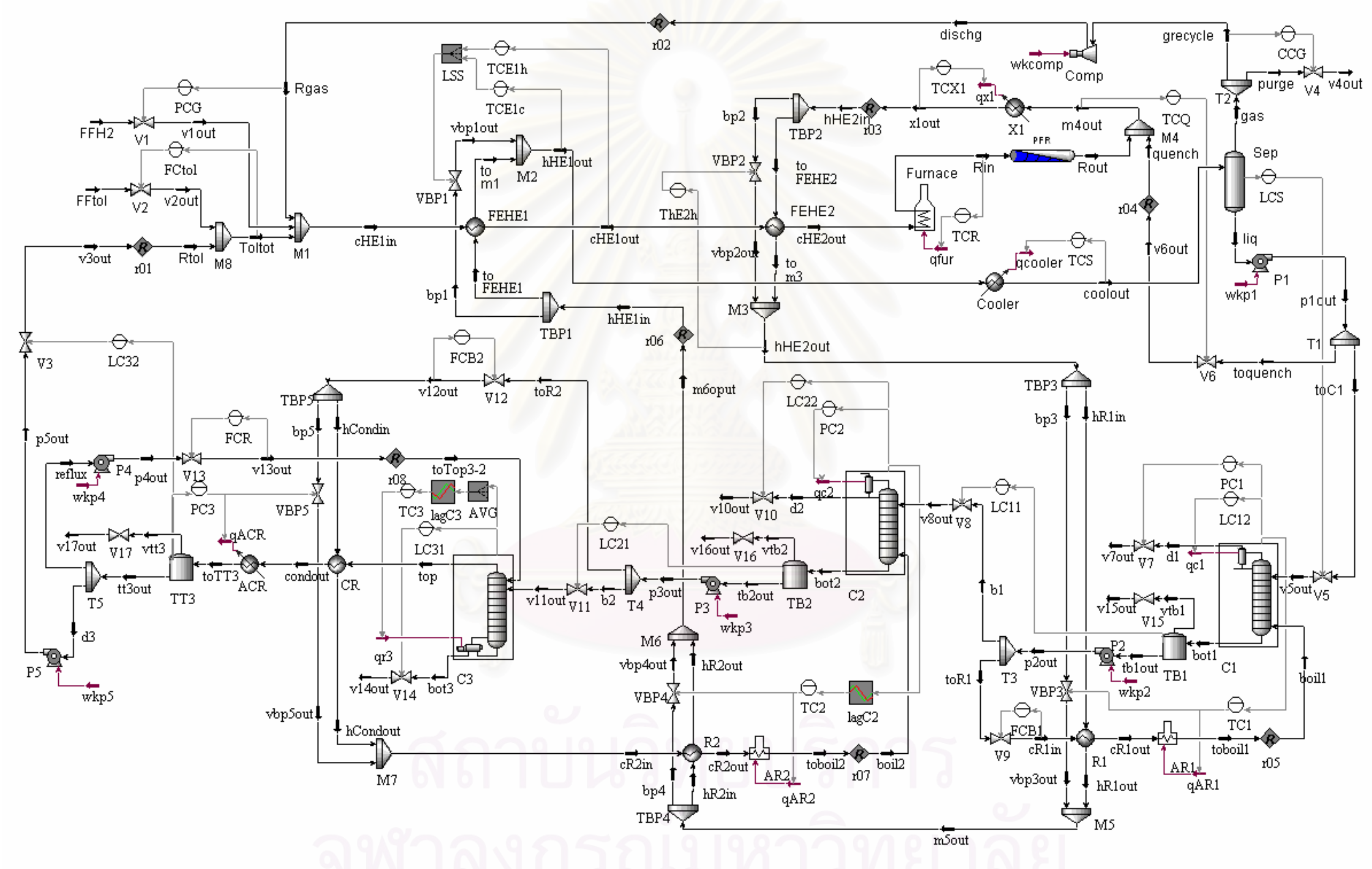


Figure 6.2 Control Structure 2 (CS2) for HDA Process Alternative 5 with three Auxiliary Utility Units

Table 6.3 Control Structure and Controller Parameter for HDA Process Alternative 5 with three Auxiliary Utility Units: Control Structure 3

controller name	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.399	0.404	0.0897
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.293	0.423	0.094
TCS	separator temperature	cooler duty (qcooler)	PID	0.898	0.359	0.0789
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.736	0.388	0.0863
TCE1h	cooler inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.730	0.379	0.0842
LSS	output of TCE1c and TCE1h	FEHE1 bypass cold stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	0.829	0.522	0.116
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC11	column C1 tray-6 temperature	R1 bypass valve (VBP3) and auxiliary reboiler 1 (AR1) duty	PID	2.57	1.95	0.433
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC21	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	5.34	4.19	0.930
TC22	column C2 tray-17 temperature	column C2 reflux flow rate	PID	2.37	11.8	2.63
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5) and auxiliary condenser(ACR) duty	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 reboiler duty (qr3)	PID	0.196	5.03	1.12
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum	toluene recycle valve (V3)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

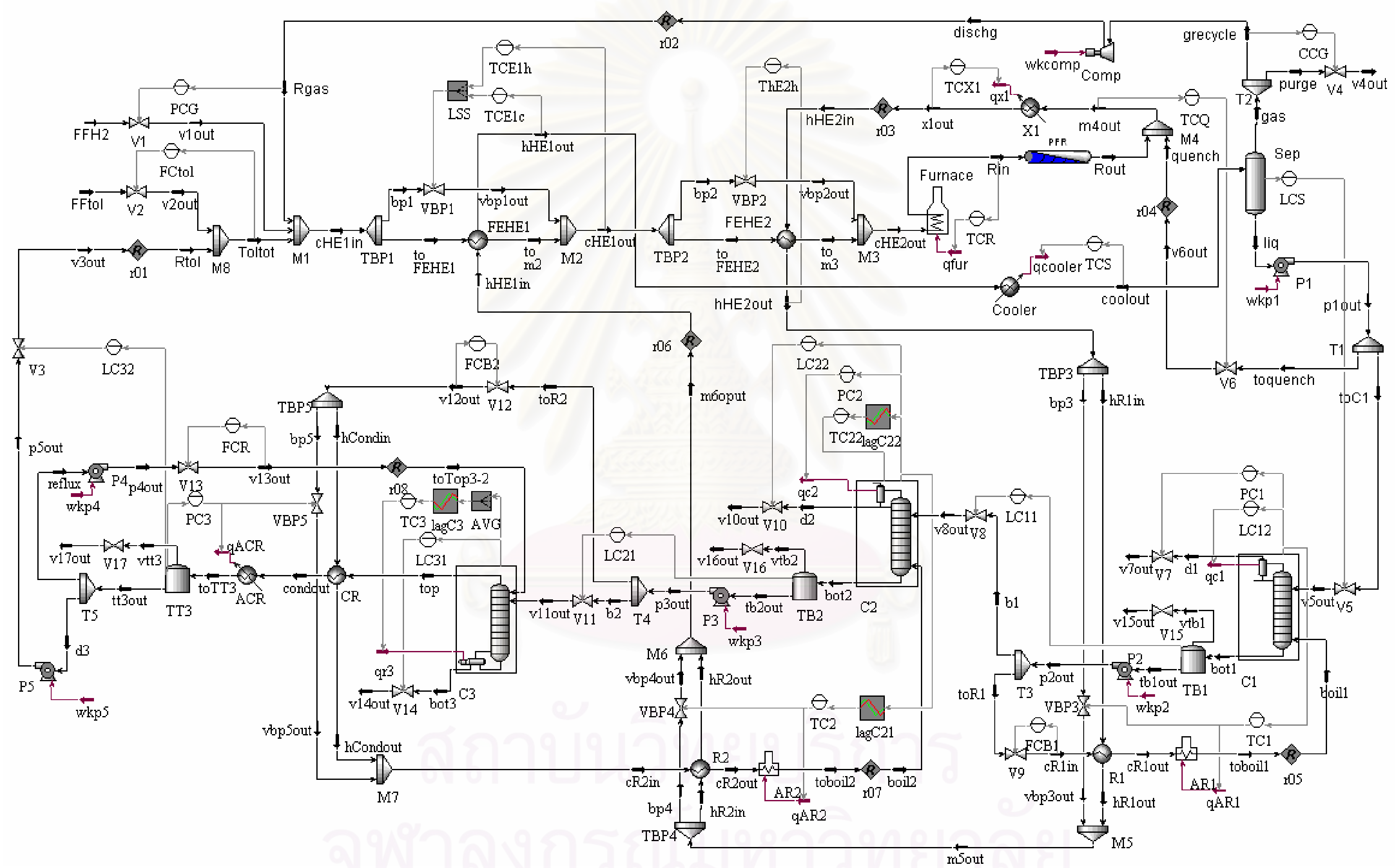


Figure 6.3 Control Structure 3 (CS3) for HDA Process Alternative 5 with three Auxiliary Utility Units

Table 6.4 Control Structure and Controller Parameter for HDA Process Alternative 5 with Minimum Auxiliary Utility Units: Control Structure 1

controller name	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.399	0.404	0.0897
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.293	0.423	0.094
TCS	separator temperature	cooler duty (qcooler)	PID	0.898	0.359	0.0789
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.736	0.388	0.0863
TCE1h	cooler inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.730	0.379	0.0842
LSS	output of TCE1c and TCE1h	FEHE1 bypass cold stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	0.829	0.522	0.116
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3)	PID	2.57	1.95	0.433
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	5.34	4.19	0.930
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 reboiler duty (qr3)	PID	0.196	5.03	1.12
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum	toluene recycle valve (V3)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

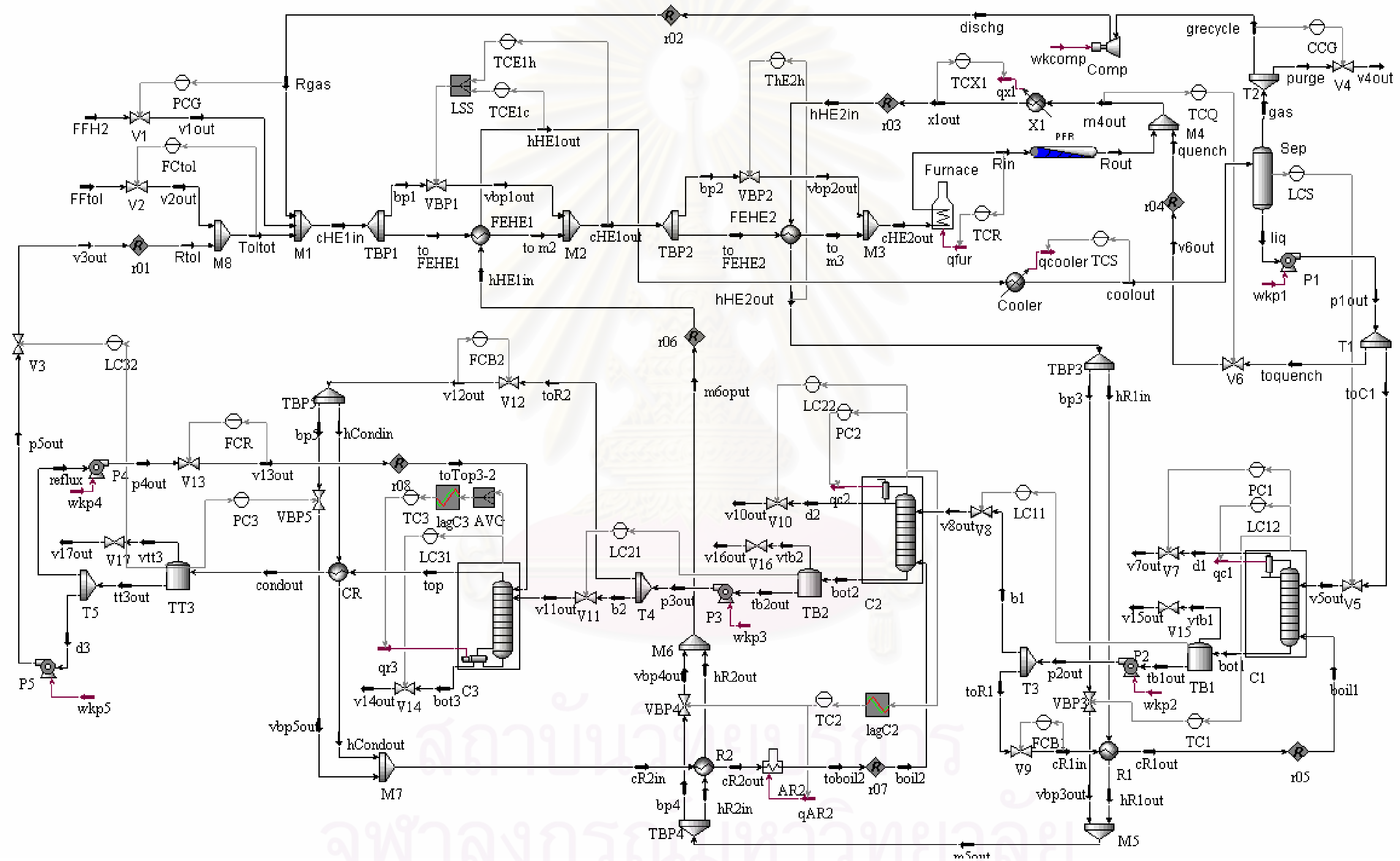


Figure 6.4 Control Structure 1 (CS1) for HDA Process Alternative 5 with Minimum Auxiliary Utility Units

Table 6.5 Control Structure and Controller Parameter for HDA Process Alternative 5 with Minimum Auxiliary Utility Units: Control Structure 2

controller name	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.399	0.404	0.0897
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.293	0.423	0.094
TCS	separator temperature	cooler duty (qcooler)	PID	0.898	0.359	0.0789
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass hot stream valve (VBP1)	PID	0.751	0.378	0.0839
TCE1h	cooler inlet temperature	FEHE1 bypass hot stream valve (VBP1)	PID	0.725	0.379	0.0843
LSS	output of TCE1c and TCE1h	FEHE1 bypass hot stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass hot stream valve (VBP2)	PID	0.0815	0.260	0.0578
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3)	PID	2.57	1.95	0.433
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	5.34	4.19	0.930
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 reboiler duty (qr3)	PID	0.196	5.03	1.12
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum	toluene recycle valve (V3)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

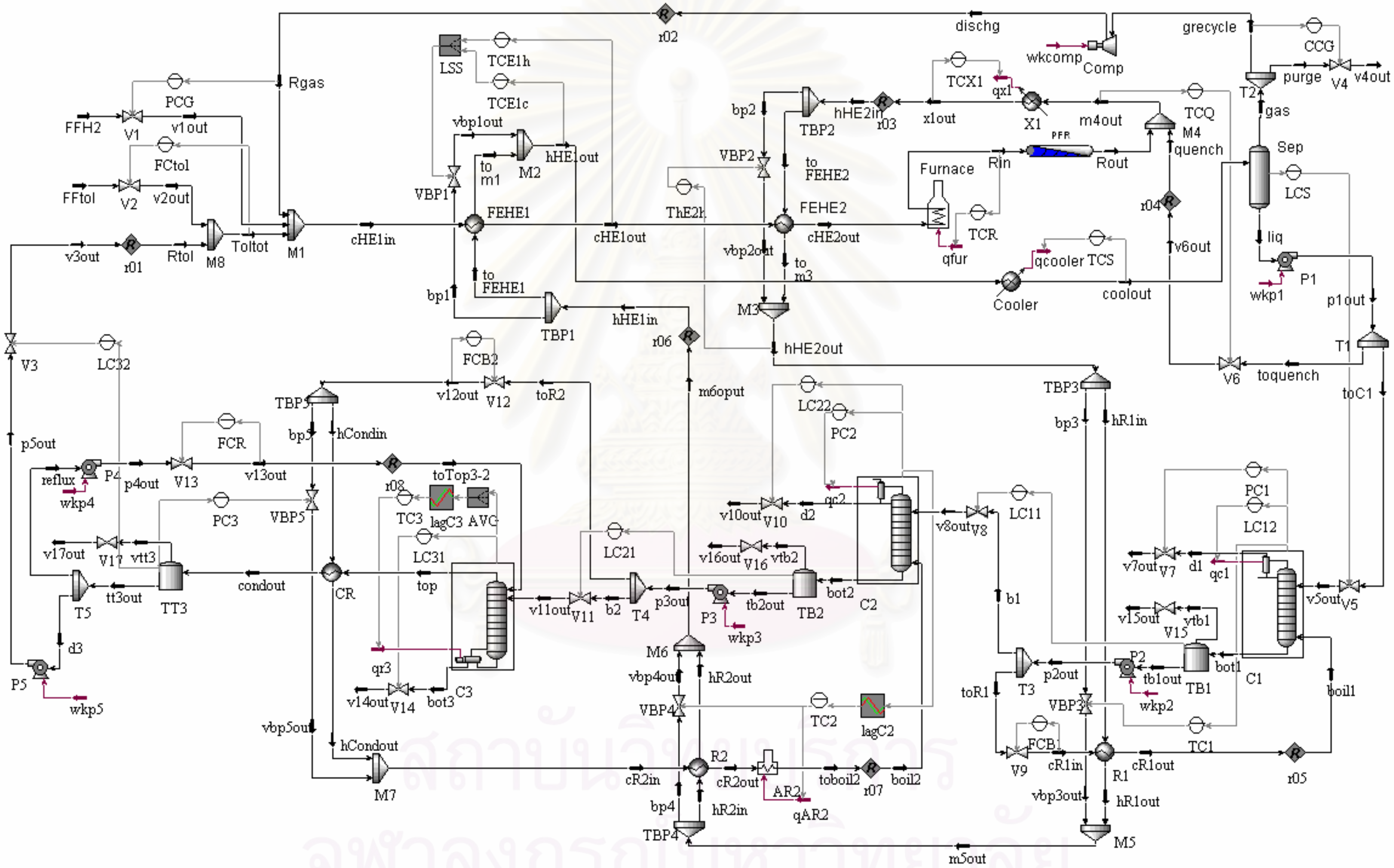


Figure 6.5 Control Structure 2 (CS2) for HDA Process Alternative 5 with Minimum Auxiliary Utility Units

Table 6.6 Control Structure and Controller Parameter for HDA Process Alternative 5 with Minimum Auxiliary Utility Units: Control Structure 3

controller name	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.399	0.404	0.0897
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.293	0.423	0.094
TCS	separator temperature	cooler duty (qcooler)	PID	0.898	0.359	0.0789
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.736	0.388	0.0863
TCE1h	cooler inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.730	0.379	0.0842
LSS	output of TCE1c and TCE1h	FEHE1 bypass cold stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	0.829	0.522	0.116
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3)	PID	2.57	1.95	0.433
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC21	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	5.34	4.19	0.930
TC22	column C2 tray-17 temperature	column C2 reflux flow rate	PID	2.37	11.8	2.63
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 reboiler duty (qr3)	PID	0.196	5.03	1.12
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum	toluene recycle valve (V3)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

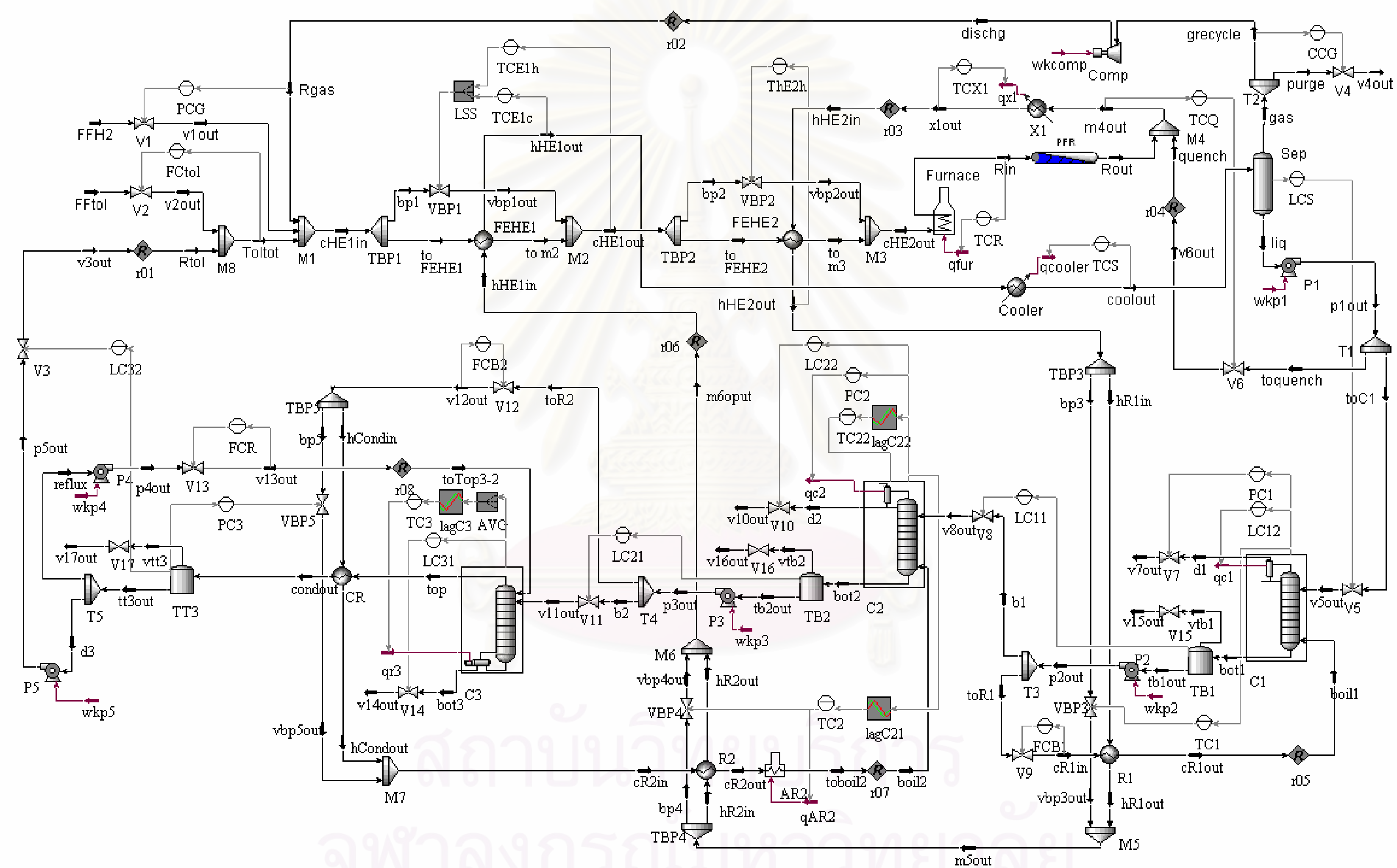


Figure 6.6 Control Structure 3 (CS3) for HDA Process Alternative 5 with Minimum Auxiliary Utility Units

6.3 Design of Plantwide Control for HDA Process Alternative 6 with four Auxiliary Utility Units and Minimum Auxiliary utility units

The three new control structures are designed for HDA process alternative 6 with three auxiliary utility units that is suggested by Luyben and the same three new design is also employed to HDA process alternative 6 with minimum auxiliary utility units to compare dynamic behavior.

6.3.1 Control Structure 1 (CS1) for HDA Process Alternative 6 with four Auxiliary Utility Units

This control structure is shown in Figure 6.7. In this control structure, the all bypass valve of 3 feed effluent heat exchangers (FEHE) are employed on hot side to control outlet temperature. A selector controller with low selector switch (LSS) for FEHE1 is employed to select an appropriate heat pathway. This control system involves one manipulated variable (bypass valve of FEHE1) and two controlled variable (hot and cold outlet temperature of FEHE1). The appropriate controlled variable is selected by LSS in order to achieve dynamic maximum energy recover (DMER).

Since the hot reactor product is used to drive reboiler in the three columns, part of this stream is bypassed and manipulated to control the tray temperatures in the three columns. In odder to prevent the propagation of thermal disturbance to separation section, the hot outlet temperatures of FEHE2 (the temperature at the entrance of the reboilers at stabilizer column,R1) are controlled by manipulating the bypass valves and the hot outlet temperatures of FEHE3 (the temperature at the entrance of the reboilers at recycle column,R3) are controlled by manipulating the bypass valves of FEHE3.

Since the four auxiliary utility units consist of three auxiliary reboilers at stabilizer column, product column and recycle column and one auxiliary condenser at recycle column. There are employed next to process to process heat exchanger as reboiler or condenser unit as show in figure 6.7. The optimum operation would be to minimize the heat load of auxiliary utility units. One way to do this is the using split range control. This control system involves two manipulated variable (the bypass valve of process to process heat exchanger and heat load of auxiliary utility unit) and one controlled variables (temperature or pressure) and works as for example: if the decreasing disturbance loads of hot stream occurs at reboiler and then the percent opening of bypass valve at reboiler will decrease in order to increase heat transfer at reboiler unit until the target tray temperature achieves. The heat load of the auxiliary utility units will be used when the bypass valve is full close but the target tray temperature still can't achieves the target tray temperature. The split range control is used for the tray temperature control in stabilizer column, product column and recycle column by manipulating the bypass valve of reboiler and the heat load of the auxiliary reboiler. For recycle column, the split control is also applied to control column pressure by manipulating the bypass valve of condenser and the heat load of the auxiliary condenser.

Since the temperature profile in the recycle column is very sharp because of temperature changes from tray to tray. This means that the process gain is very large when a single tray temperature is controlled. The standard solution for this problem is to use an average (AVG) temperature of several trays instead of a single tray (Luyben, 2002).

A heat exchanger (i.e. as a heat source or a heat sink) is artificially installed in the hot-side stream (i.e. the exchanger X1 in Fig. 6.7) in order to make the disturbance loads of the hot stream (i.e. the hot reactor product). Note that, this exchanger is not used in the real plant, and the temperature controller TCX1 is set to be "off" whenever it is not used to make the disturbances.

The control structure and controller parameter are given in table 6.7. P controllers are employed for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.3.2 Control Structure 2 (CS2) for HDA Process Alternative 6 with four Auxiliary utility unit

This control structure is shown in Figure 6.8. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass valve of 2 FEHEs will be employed on hot side to control the outlet temperature of FEHE. The controller parameter is given in table 6.8.

6.3.3 Control Structure 3 (CS3) for HDA Process Alternative 6 with four Auxiliary utility unit

This control structure is shown in Figure 6.9 and the control configuration is given in table 6.9. The major loops in this control structure are the same as CS1 except for temperature control in product distillation column. The temperature control in product distillation column is two point controls as the tray 12 and tray 17 temperatures.

6.3.4 Control Structure 1 (CS1) for HDA Process Alternative 6 with Minimum Auxiliary utility unit

The same control structure 1 for HDA process alternative 6 with four auxiliary utility units is employed to HDA process alternative 6 with minimum auxiliary utility units (figure 6.10). Since the number of minimum auxiliary utility unit for HDA process alternative 6 is only one unit therefore the split range control will employ only one for product column to control the tray temperature. The control structure and controller parameter are given in table 6.10.

6.3.5 Control Structure 2 (CS2) for HDA Process Alternative 6 with Minimum Auxiliary utility unit

The same control structure 2 for HDA process alternative 6 with four auxiliary utility units is employed to HDA process alternative 6 with minimum auxiliary utility units (figure 6.11). Since the number of minimum auxiliary utility unit for HDA process alternative 5 is only one unit therefore the split range control will employ only one for product column to control the tray temperature. The control structure and controller parameter are given in table 6.11.

6.3.6 Control Structure 3 (CS3) for HDA Process Alternative 5 with Minimum Auxiliary utility unit

The same control structure 3 for HDA process alternative 6 with four auxiliary utility units is employed to HDA process alternative 6 with minimum auxiliary utility units (figure 6.12). Since the number of minimum auxiliary utility unit for HDA process alternative 5 is only one unit therefore the split range control will employ only one for product column to control the tray temperature. The control structure and controller parameter are given in table 6.12.

Table 6.7 Control Structure and Controller Parameter for HDA Process Alternative 6 with four Auxiliary Utility Units: Control Structure 1

controller	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.375	0.406	0.090
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.333	0.420	0.093
TCS	separator temperature	cooler duty (qcooler)	PID	0.923	0.367	0.081
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.578	0.433	0.096
TCE1h	cooler inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.644	0.275	0.061
LSS	output of TCE1c and TCE1h	FEHE1 bypass cold stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	1.24	0.207	0.046
TCE3h	FEHE3 hot-outlet temperature	FEHE3 bypass cold stream valve (VBP3)	PID	0.566	0.893	0.199
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3) and auxiliary reboiler 1 (AR1) duty	PID	5.25	2.04	0.452
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	9.83	4.23	0.941
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5) and auxiliary condenser(ACR) duty	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	R3 bypass valve (VBP4) and auxiliary reboiler 3 (AR3) duty	PID	0.641	5.52	1.23
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-
FCB3	column C3 boil up flow rate	R3 cold-inlet valve (V15)	PI	0.5	0.3	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

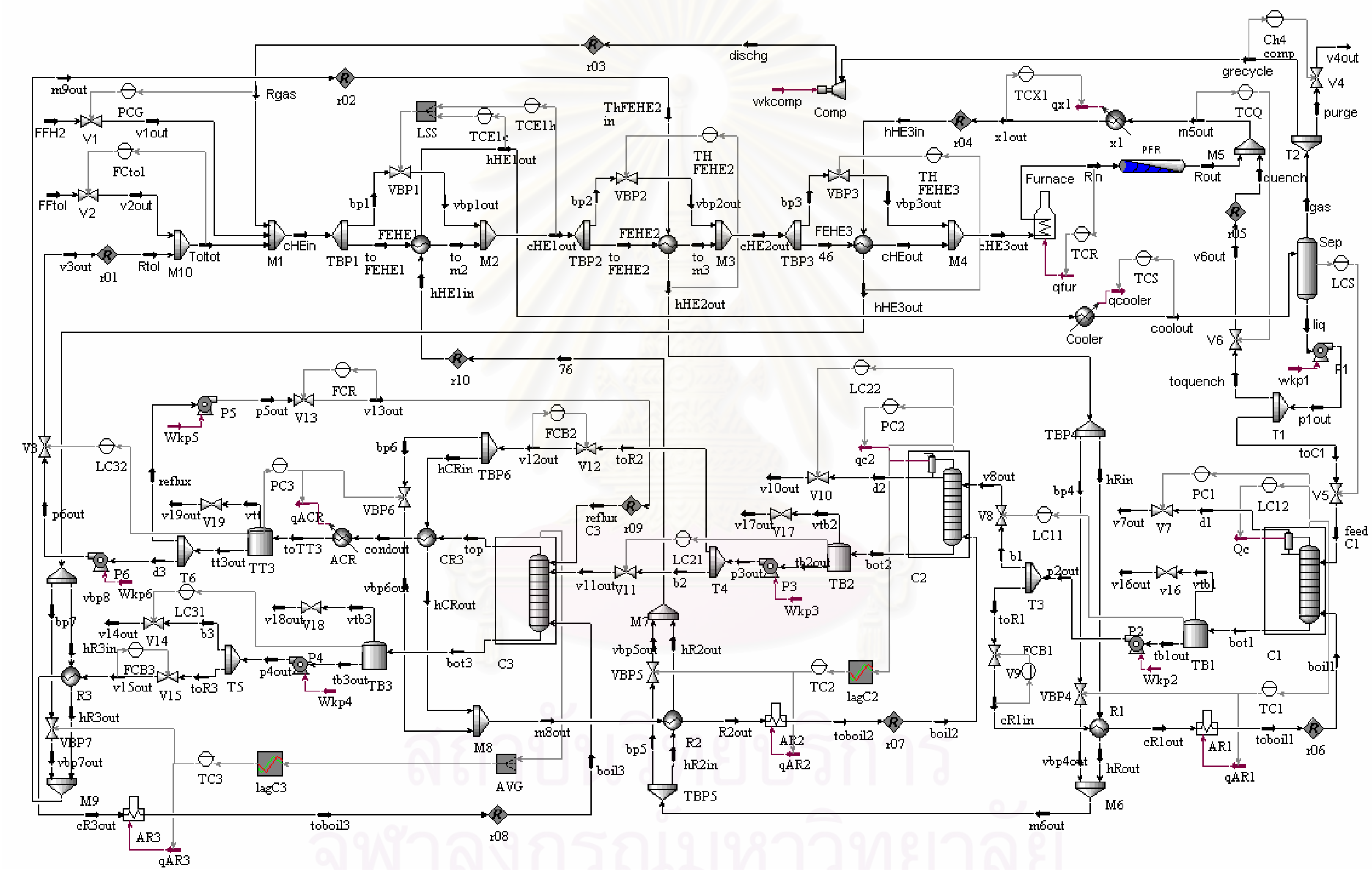


Figure 6.7 Control Structure 1 (CS1) for HDA Process Alternative 6 with four Auxiliary Utility Units

Table 6.8 Control Structure and Controller Parameter for HDA Process Alternative 6 with four Auxiliary Utility Units: Control Structure 2

controller	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.375	0.406	0.090
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.333	0.420	0.093
TCS	separator temperature	cooler duty (qcooler)	PID	0.923	0.367	0.081
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass hot stream valve (VBP1)	PID	0.782	0.212	0.047
TCE1h	cooler inlet temperature	FEHE1 bypass hot stream valve (VBP1)	PID	1.07	0.227	0.05
LSS	output of TCE1c and TCE1h	FEHE1 bypass hot stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass hot stream valve (VBP2)	PID	0.417	0.312	0.069
TCE3h	FEHE3 hot-outlet temperature	FEHE3 bypass hot stream valve (VBP3)	PID	0.124	0.387	0.086
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3) and auxiliary reboiler 1 (AR1) duty	PID	5.25	2.04	0.452
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	9.83	4.23	0.941
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5) and auxiliary condenser(ACR) duty	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	R3 bypass valve (VBP4) and auxiliary reboiler 3 (AR3) duty	PID	0.641	5.52	1.23
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-
FCB3	column C3 boil up flow rate	R3 cold-inlet valve (V15)	PI	0.5	0.3	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

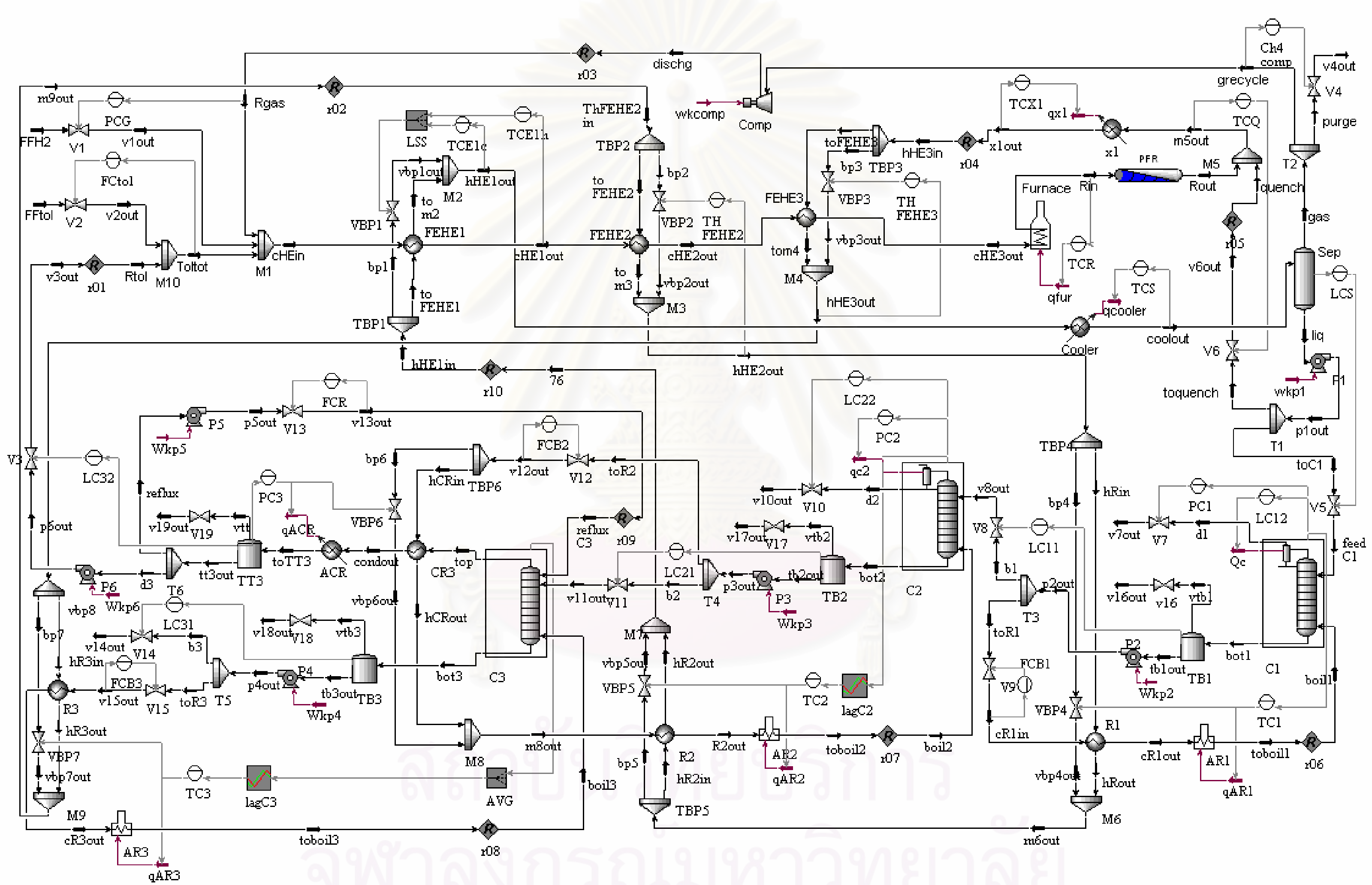


Figure 6.8 Control Structure 2 (CS2) for HDA Process Alternative 6 with four Auxiliary Utility Units

Table 6.9 Control Structure and Controller Parameter for HDA Process Alternative 6 with four Auxiliary Utility Units: Control Structure 3

controller	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.375	0.406	0.090
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.333	0.420	0.093
TCS	separator temperature	cooler duty (qcooler)	PID	0.923	0.367	0.081
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.578	0.433	0.096
TCE1h	cooler inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.644	0.275	0.061
LSS	output of TCE1c and TCE1h	FEHE1 bypass cold stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	1.24	0.207	0.046
TCE3h	FEHE3 hot-outlet temperature	FEHE3 bypass cold stream valve (VBP3)	PID	0.566	0.893	0.199
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3) and auxiliary reboiler 1 (AR1) duty	PID	5.25	2.04	0.452
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC21	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	9.83	4.23	0.941
TC22	column C2 tray-17 temperature	column 2 reflux flow rate	PID	9.83	4.23	0.941
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum level	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5) and auxiliary condenser(ACR) duty	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	R3 bypass valve (VBP4) and auxiliary reboiler 3 (AR3) duty	PID	0.641	5.52	1.23
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-
FCB3	column C3 boil up flow rate	R3 cold-inlet valve (V15)	PI	0.5	0.3	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

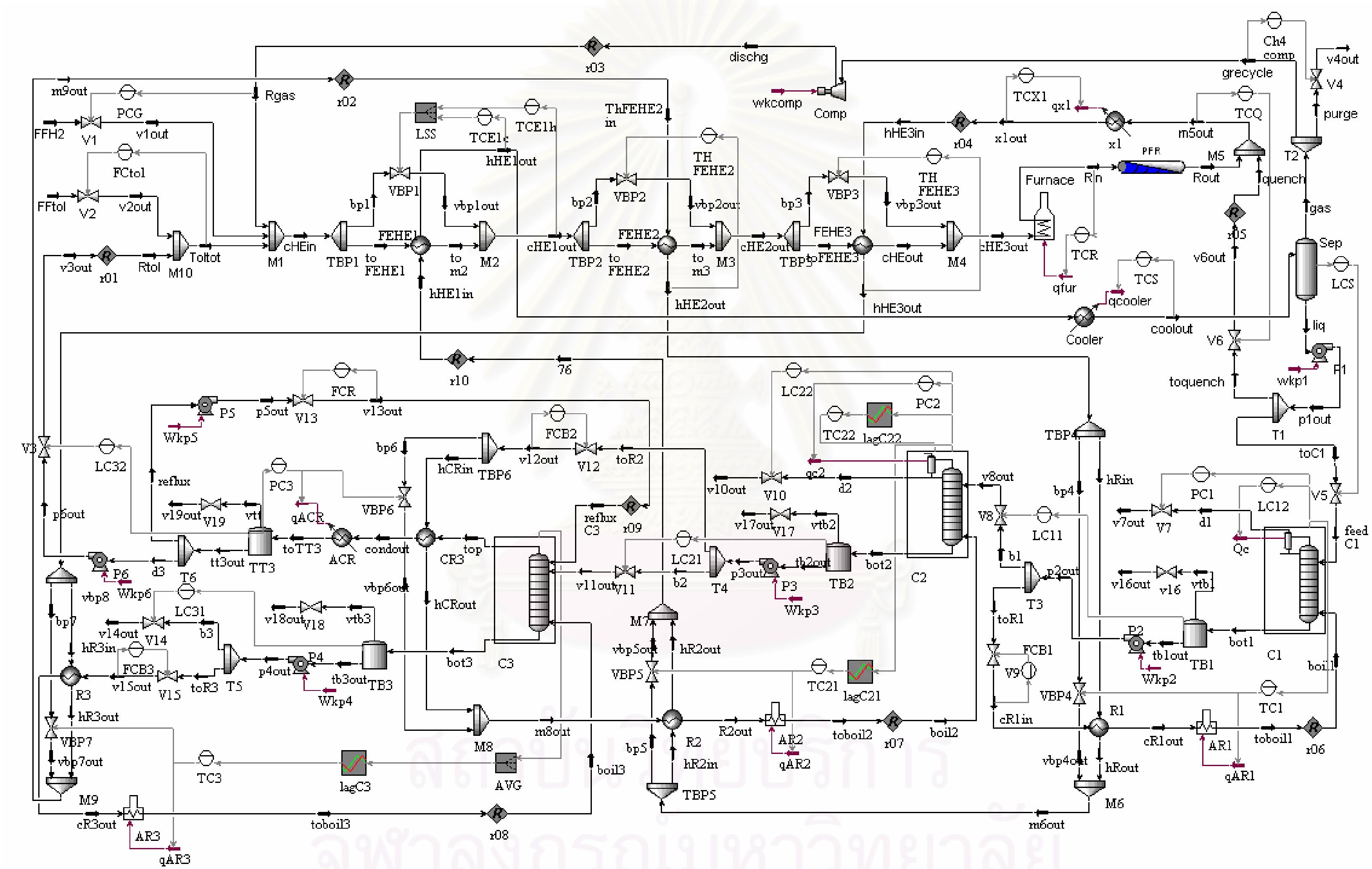


Figure 6.9 Control Structure 2 (CS2) for HDA Process Alternative 6 with four Auxiliary Utility Units

Table 6.10 Control Structure and Controller Parameter for HDA Process Alternative 6 with Minimum Auxiliary Utility Units: Control Structure 1

controller	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.375	0.406	0.090
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.333	0.420	0.093
TCS	separator temperature	cooler duty (qcooler)	PID	0.923	0.367	0.081
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.578	0.433	0.096
TCE1h	cooler inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.644	0.275	0.061
LSS	output of TCE1c and TCE1h	FEHE1 bypass cold stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	1.24	0.207	0.046
TCE3h	FEHE3 hot-outlet temperature	FEHE3 bypass cold stream valve (VBP3)	PID	0.566	0.893	0.199
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3)	PID	5.25	2.04	0.452
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	9.83	4.23	0.941
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	R3 bypass valve (VBP4)	PID	0.641	5.52	1.23
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-
FCB3	column C3 boil up flow rate	R3 cold-inlet valve (V15)	PI	0.5	0.3	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

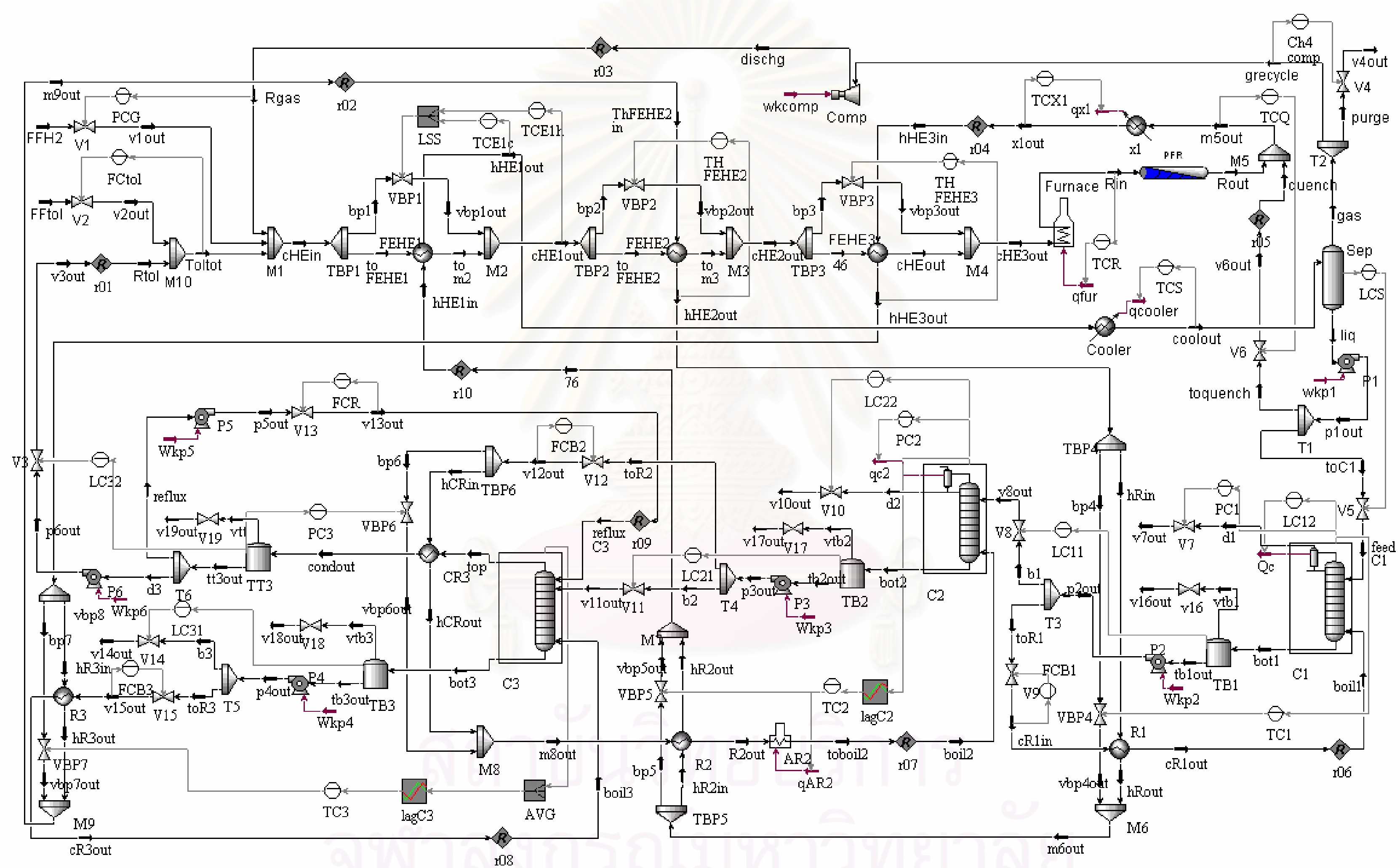


Figure 6.10 Control Structure 1 (CS1) for HDA Process Alternative 6 with Minimum Auxiliary Utility Units

Table 6.11 Control Structure and Controller Parameter for HDA Process Alternative 6 with Minimum Auxiliary Utility Units: Control Structure 2

controller	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.375	0.406	0.090
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.333	0.420	0.093
TCS	separator temperature	cooler duty (qcooler)	PID	0.923	0.367	0.081
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass hot stream valve (VBP1)	PID	0.782	0.212	0.047
TCE1h	cooler inlet temperature	FEHE1 bypass hot stream valve (VBP1)	PID	1.07	0.227	0.05
LSS	output of TCE1c and TCE1h	FEHE1 bypass hot stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass hot stream valve (VBP2)	PID	0.417	0.312	0.069
TCE3h	FEHE3 hot-outlet temperature	FEHE3 bypass hot stream valve (VBP3)	PID	0.124	0.387	0.086
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3)	PID	5.25	2.04	0.452
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	9.83	4.23	0.941
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	R3 bypass valve (VBP4)	PID	0.641	5.52	1.23
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-
FCB3	column C3 boil up flow rate	R3 cold-inlet valve (V15)	PI	0.5	0.3	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

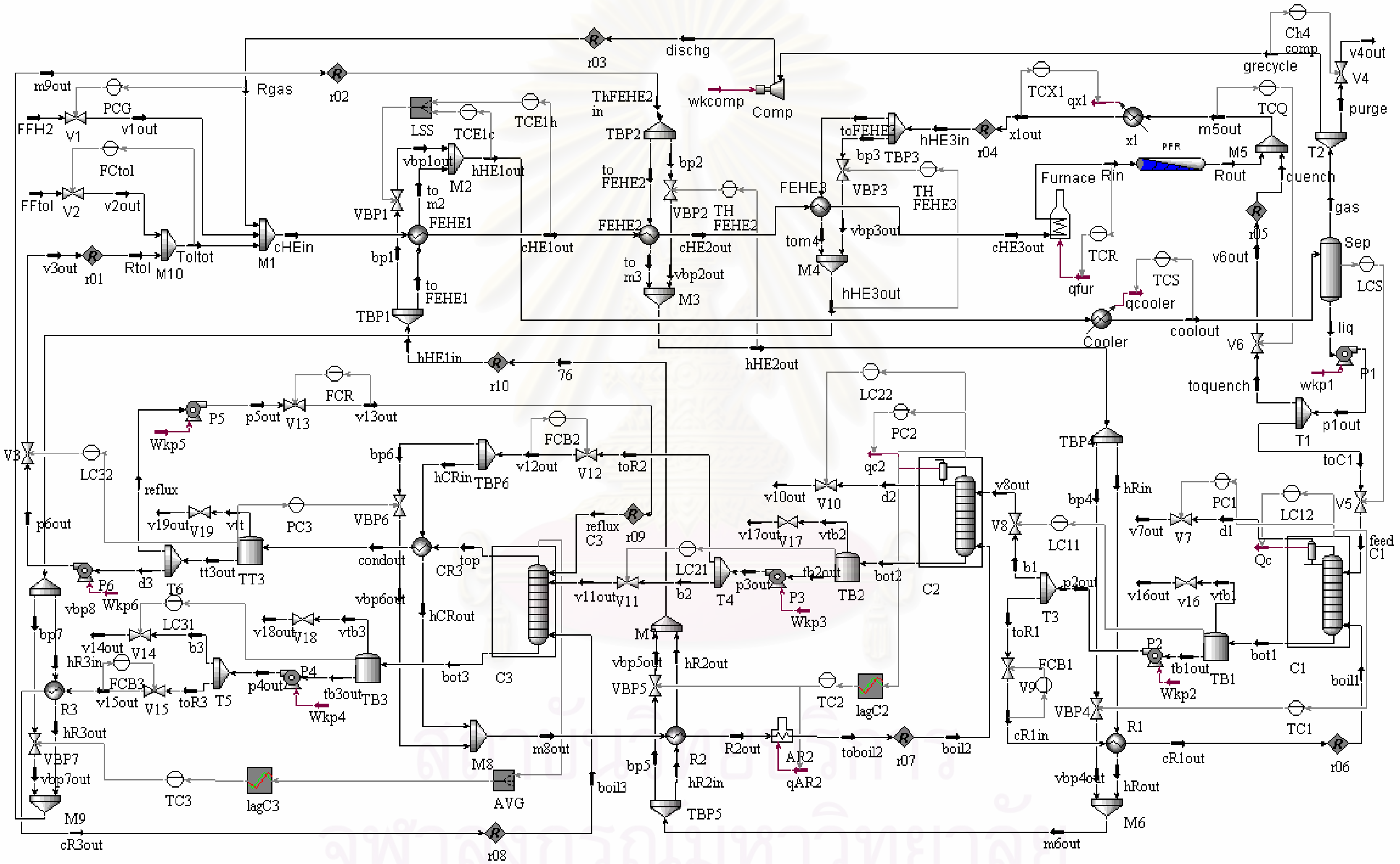


Figure 6.11 Control Structure 2 (CS2) for HDA Process Alternative 6 with Minimum Auxiliary Utility Units

Table 6.12 Control Structure and Controller Parameter for HDA Process Alternative 6 with Minimum Auxiliary Utility Units: Control Structure 3

controlle r name	controlled variable	manipulated variable	type	Kc	Ti	Td
FCtol	total toluene flow rate	fresh feed toluene valve (V2)	PI	0.5	0.3	-
PCG	gas recycle stream pressure	fresh feed hydrogen valve (V1)	PI	2	10	-
CCG	methane in gas recycle	purge valve (V4)	PI	0.5	15	-
TCQ	quenched temperature	quench valve (V6)	PID	0.375	0.406	0.090
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.333	0.420	0.093
TCS	separator temperature	cooler duty (qcooler)	PID	0.923	0.367	0.081
TCE1c	FEHE2 cold inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.578	0.433	0.096
TCE1h	cooler inlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	0.644	0.275	0.061
LSS	output of TCE1c and TCE1h	FEHE1 bypass cold stream valve (VBP1)	Min	-	-	-
TCE2h	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	1.24	0.207	0.046
TCE3h	FEHE3 hot-outlet temperature	FEHE3 bypass cold stream valve (VBP3)	PID	0.566	0.893	0.199
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
PC1	column C1 pressure	column C1 gas valve (V7)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3)	PID	5.25	2.04	0.452
LC11	column C1 base level	column C2 feed valve (V8)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	3	-	-
FCB1	column C1 boil up	cold-inlet valve of R1 flow rate (V9)	PI	0.5	0.3	-
PC2	column C2pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC21	column C2 tray-12 temperature	R2 bypass valve (VBP4) and auxiliary reboiler 2 (AR2) duty	PID	9.83	4.23	0.941
TC22	column C2 tray-17 temperature	column 2 reflux flow rate	PID	9.83	4.23	0.941
LC21	column C2 base level	column C3 feed valve (V11)	P	2	-	-
LC22	column C2 reflux drum level	column C2 product valve level (V10)	P	2	-	-
FCB2	column C2 boil up flow rate	R2 cold-inlet valve (V12)	PI	0.5	0.3	-
PC3	column C3 pressure	CR bypass valve (VBP5)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	R3 bypass valve (VBP4)	PID	0.641	5.52	1.23
LC31	column C3 base level	C3 bottom valve (V14)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-
FCB3	column C3 boil up flow rate	R3 cold-inlet valve (V15)	PI	0.5	0.3	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

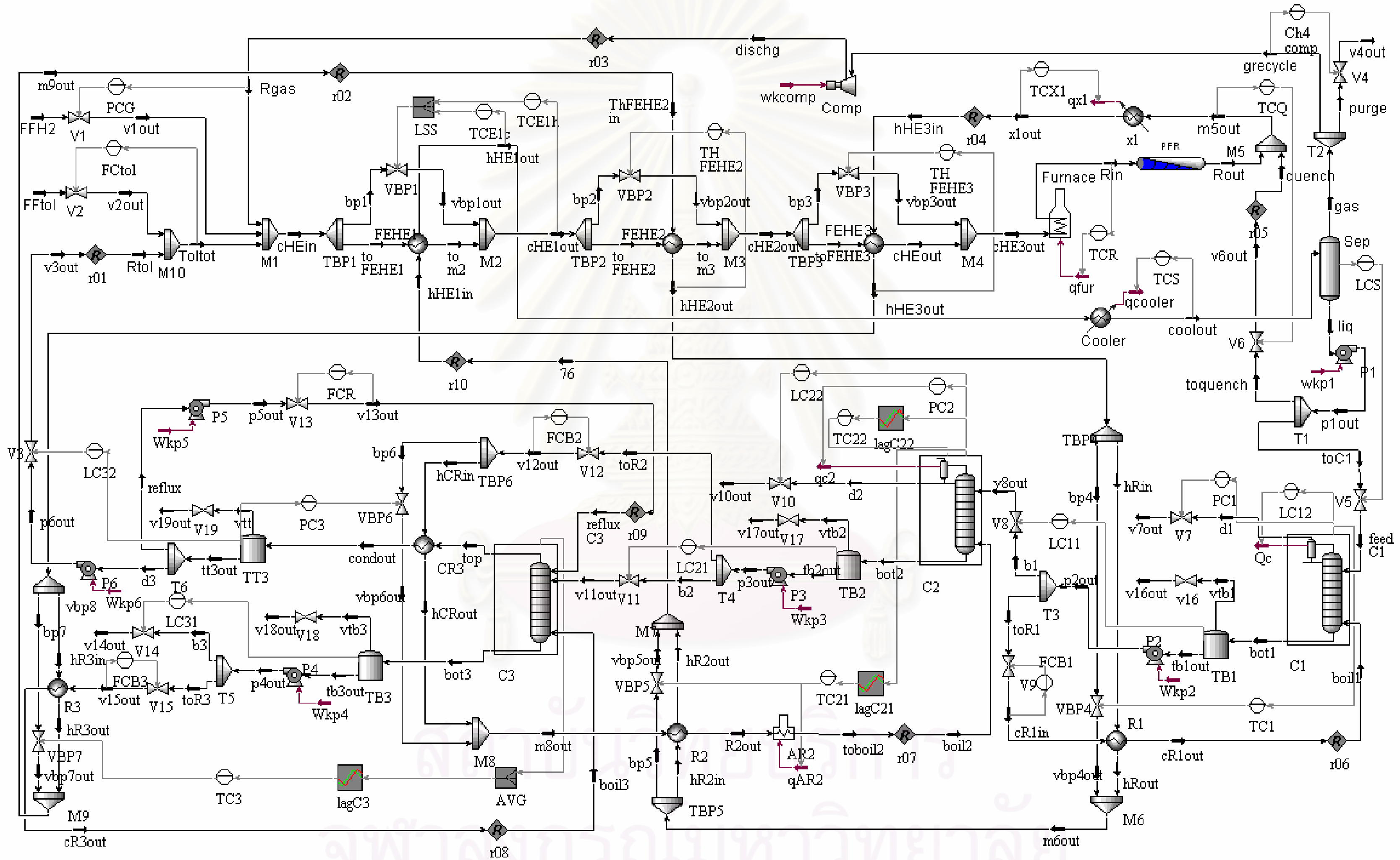


Figure 6.12 Control Structure 3 (CS3) for HDA Process Alternative 6 with Four Auxiliary Utility Units

6.4 Dynamic Simulation Results for HDA Process Alternative 5 with three Auxiliary Utility Units and Minimum Auxiliary Utility Units

In order to illustrate the dynamic behavior of the control structure in HDA process alternative 5 with three auxiliary utility units and minimum auxiliary utility units, several disturbance loads are made. The dynamic results are explained in this part.

6.4.1 Dynamic Simulation Results for HDA Process Alternative 5 with three Auxiliary Utility Units

Three disturbance loads are used to evaluate the dynamic performance of the new control structures 1, 2 and 3 for HDA process alternative 5 with three auxiliary utility units.

6.4.1.1 Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream)

Figures 6.13, 6.16 and 6.19 show the dynamic responses of the control systems of HDA process alternative 5 with three auxiliary utility units to a change in the heat load disturbance of cold stream (reactor feed stream). In order to make this disturbance, first the fresh toluene feed temperature is decreased from 30 to 20^{o}C at time equals 10 minutes, and the temperature is increased from 20 to 40^{o}C at time equals 100 minutes, then its temperature is returned to its nominal value of 30^{o}C at time equals 200 minutes (Figures 6.13.a, 6.16.a and 6.1.a).

The three new control structures (CS1, CS2 and CS3) give the same result for shifting the thermal disturbance load in cold steam (reactor feed stream) to furnace or cooler as follows.

In the first the cold inlet temperature of FEHE1 is decreased and then both the cold and hot outlet temperatures of FEHE1 decrease suddenly. The hot outlet temperature decreasing is a desired condition, hence the LSS switches the control action to control the cold outlet temperature of FEHE1 at its set point value of 141.1^{o}C (Figures 6.13.e, 6.16.e and 6.19.e). As a result, the hot outlet temperature of FEHE1 rapidly drops to a new steady state value (Figures 6.13.c, 6.16.c and 6.19.c), and the cooler duty decreases as DMER achieve (Figures 6.13.j, 6.16.j and 6.19.j). When the cold inlet temperature of FEHE1 increases, both the cold and hot outlet temperatures of FEHE1 increase. In order to achieve maximum energy recovery, the increasing cold outlet temperature is a desired condition. So the LSS switches the control action to control the hot outlet temperature of FEHE1 at its set point value of 108.3 ^{o}C (Figures 6.13.c, 6.16.c and 6.19.c). As a result, the cold outlet temperature of FEHE1 temperature quickly increases a new steady state value (Figures 6.13.e, 6.16.e and 6.19.e), and the furnace duty decreases as DMER achieve (Figures 6.13.i, 6.16.i and 6.19.i).

The hot outlet temperature of FEHE2 is slightly well controlled to prevent the thermal disturbance load propagation to the heat exchanger R1 (reboiler of stabilizer).

As can be seen, this disturbance load has a little bit effect to the tray temperatures in the product and recycle columns except the tray temperature in stabilizer column however the three new control structures can control the tray temperature in stabilizer column slightly well (Figures 6.13.n, 6.16.n and 6.19.n).

The reactor inlet temperature, the quench temperature, the separator temperature are also slightly well controlled but the dynamic response of CS2 is smoother than CS1 and CS3.

6.4.1.2 Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream)

Figures 6.14, 6.17 and 6.20 shows the dynamic responses of the control systems of HDA process alternative 5 with three auxiliary utility units to a change in the heat load disturbance of hot stream (the hot reactor product). This disturbance is made as follows: first the set point of FEHE3-hot-inlet temperature controller (i.e. TCX1 in Figure 6.1) is decreased from 621.1 to 611.1oC at time equals 10 minutes, and the temperature is increased from 611.1 to 631.1oC at time equals 100 minutes, then its temperature is returned to its nominal value of 621.1oC at time equals 200 minutes (Figure 6.14.a, 6.17.a and 6.20.a). As can be seen, this temperature response is very fast (Figure 6.14.a, 6.17.a and 6.20.a), the new steady state is reached quickly (Figures 6.14.a, 6.17.a and 6.20.a).

Since the hot outlet temperature of FEHE2 (i.e. the hot temperature at the entrance of reboiler at stabilizer column, R1) is controlled (Figures 6.14.g, 6.17.g and 6.20.g) at its set point value of 300 ^{o}C to prevent the propagation of the thermal disturbance, both the positive and negative disturbance loads of the hot stream are shifted to the furnace utility. Therefore, whenever the negative disturbance load comes with the hot stream (i.e. the hot inlet temperature of FEHE2 decreases Figures

6.14.a, 6.17.a and 6.20.a), this disturbance load is shifted to the furnace utility. The furnace duty will be increased in this case (Figures 6.14.h, 6.17.h and 6.20.h). Consider the case when the hot inlet temperature of FEHE2 increases (Figures 6.14.a, 6.17.a and 6.20.a), this is a desired condition to shift the disturbance load to the cold stream. Therefore, the furnace duty decreases to a new steady state value (Figures 6.14.h, 6.17.h and 6.20.h).

For the tray temperatures in the stabilizer, product and recycle columns are slightly well controlled but CS2 is more oscillation than CS1 and CS3.

The reactor inlet temperature, the quench temperature, the separator temperature are slightly well controlled but the dynamic response of CS2 is smoother than CS1 and CS3.

6.4.1.3 Change in the Total Toluene Feed Flowrates

Figure 6.15, 6.18 and 6.21 shows the dynamic responses of the control systems of HDA process alternative 5 with three auxiliary utility units to a change in the total toluene flowrates. This disturbance is made by decreasing toluene flowrates from 172.3 to 162.3 kgmole/h at time equals 10 minutes, and the flowrates is increased from 162.3 to 182.3 kgmole/h at time equals 100 minutes, then its flowrates is returned to its nominal value of 172.3 kgmole/h at time equals 200 (Figures 6.15.a , 6.18.a and 6.21.a).

The dynamic result can be seen that the drop in total toluene feed flowrates reduces the reaction rate, so the benzene product flowrates drops (Figures 6.15.c, 6.18.c and 6.21.c).and the benzene product quality increases (Figures 6.15.d, 6.18.d and 6.21.d).Consider the case when the total toluene feed flowrates increase (Figures 6.15.a, 6.18.a and 6.21.a), and the benzene product flowrates increase because of the reaction rate enlargement. The benzene product quality will increase in this case (Figures 6.15.d, 6.18.d and 6.21.d). The deviation of benzene product quality from nominal value in CS1, CS2 and CS3 are same when total toluene feed flowrates change.

As can be seen, this disturbance has height effect to trey temperatures in the stabilizer, product and recycle columns. However all control structure can control the tray temperatures in the three columns slightly well.

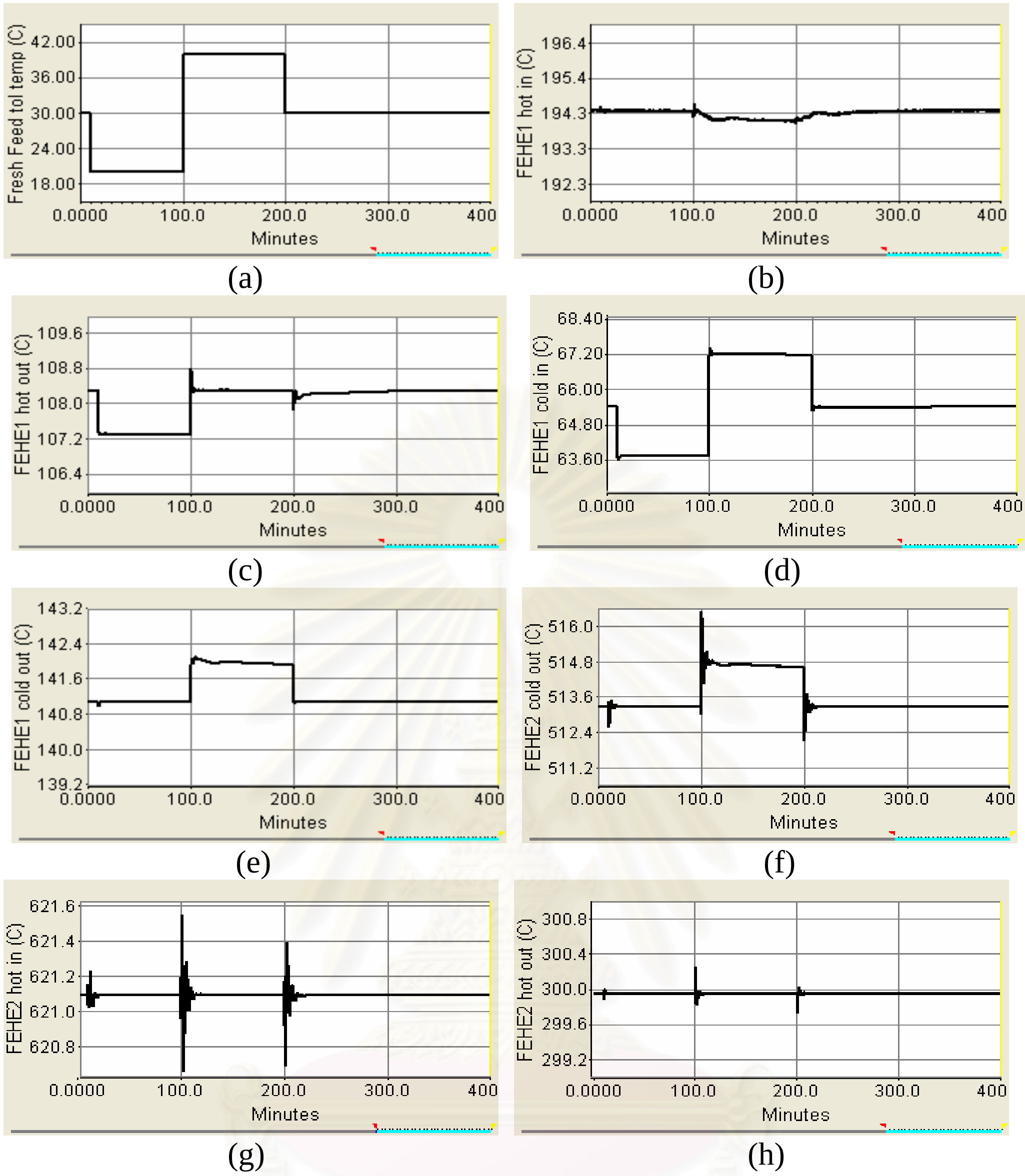
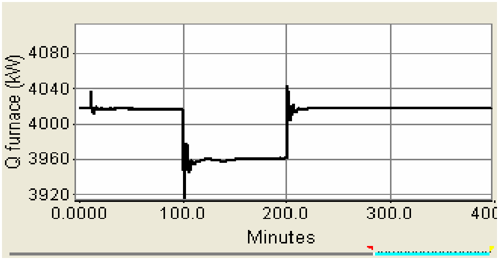
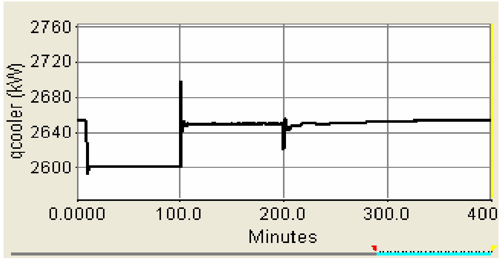
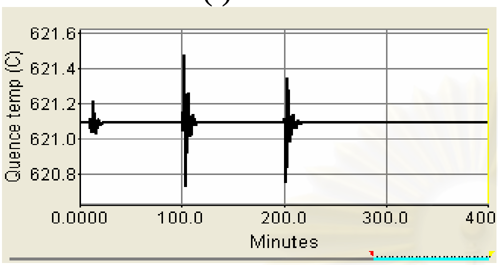


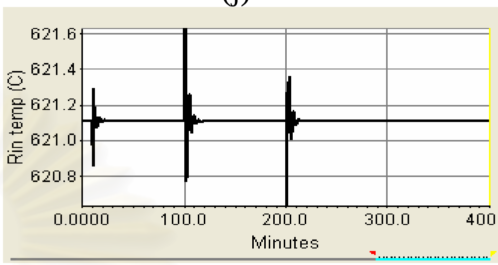
Figure 6.13 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature, (i) furnace duty, (j) cooler duty, (k) quench temperature, (l) reactor inlet temperature, (m) separator temperature, (n) stabilizer column tray temperature, (o) product column tray temperature, (p) recycle column tray temperature





(i) (j)





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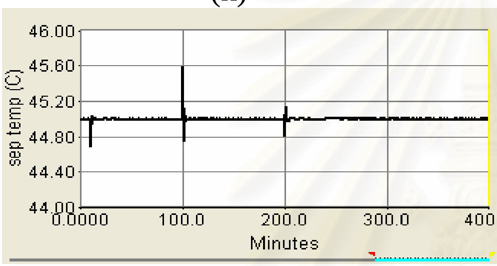

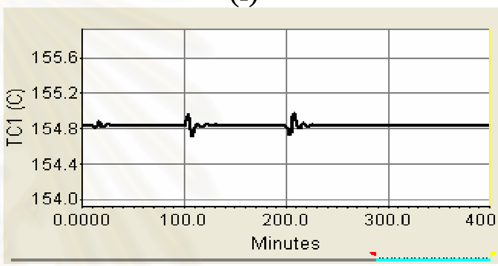

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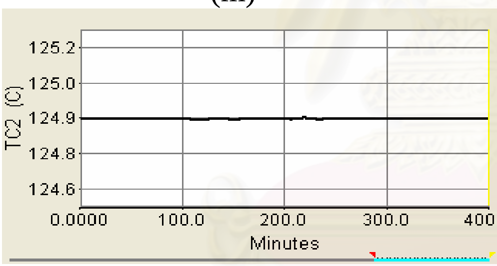
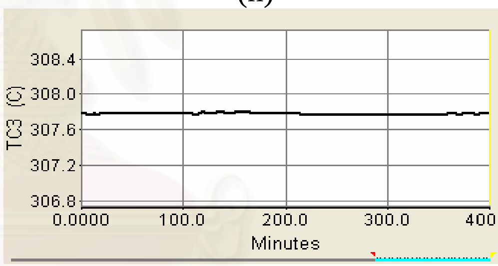




(o) (p)

Figure 6.13 (continue)

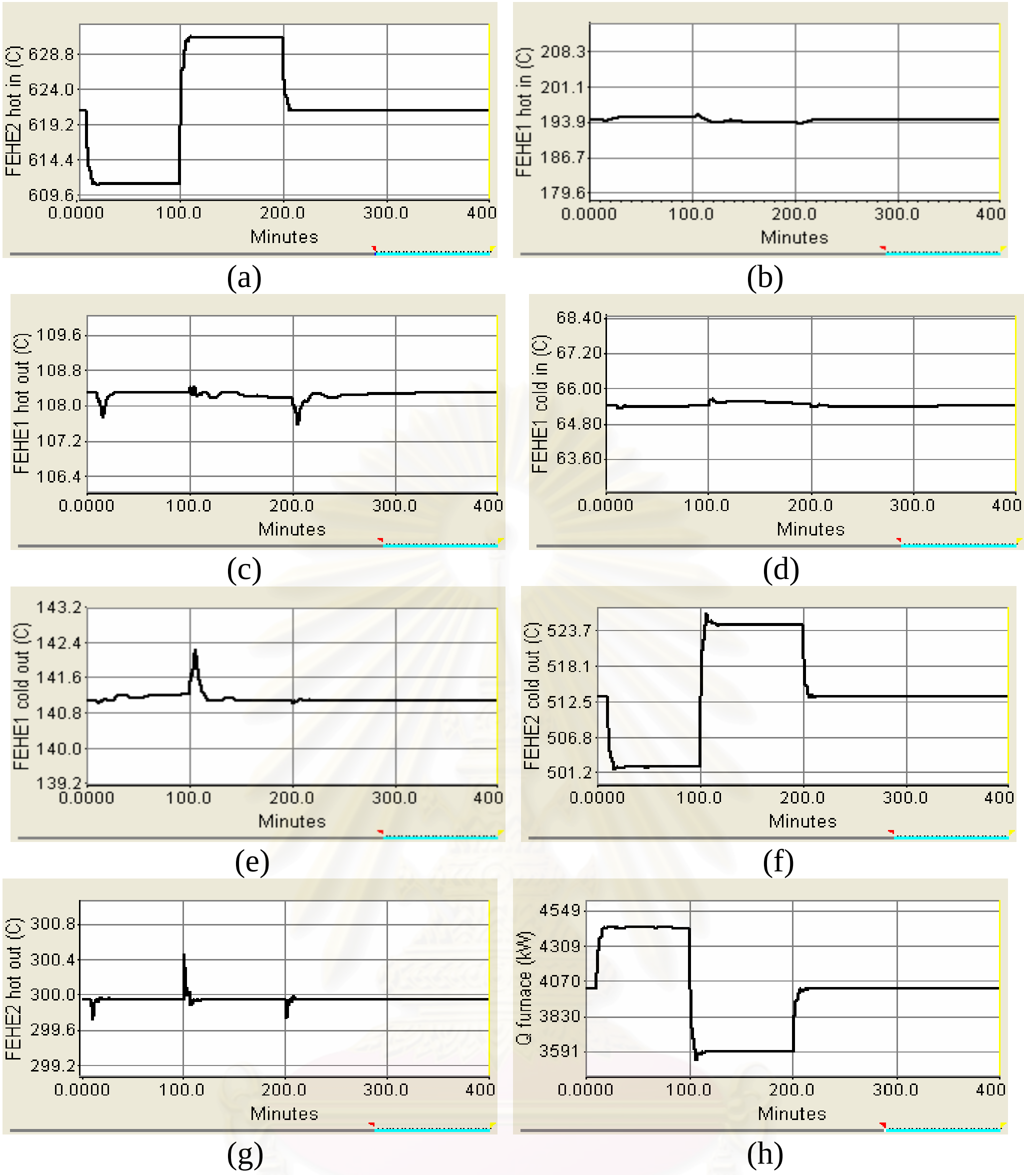
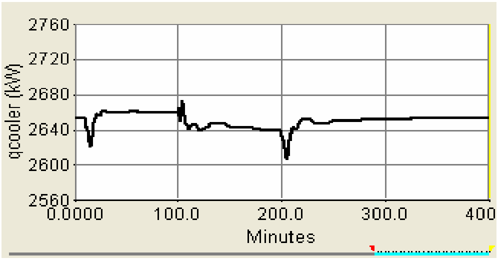


Figure 6.14 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot outlet temperature, (g) FEHE2 cold outlet temperature, (h) furnace duty, (i) cooler duty, (j) quench temperature, (k) reactor inlet temperature, (l) separator temperature, (m) stabilizer column tray temperature, (n) product column tray temperature, (o) recycle column tray temperature



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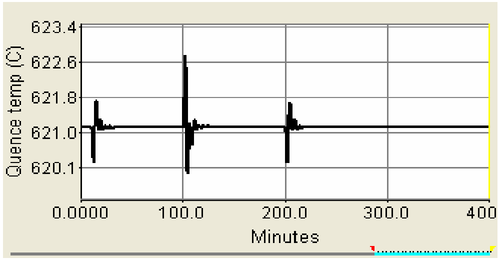

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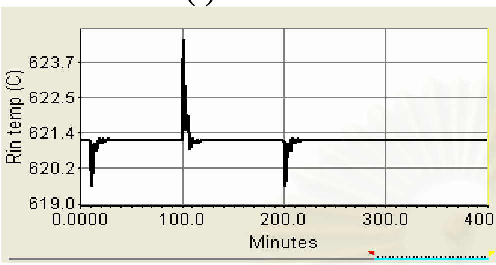

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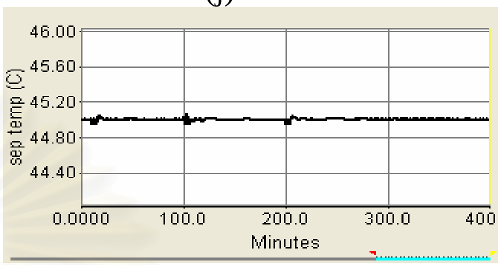

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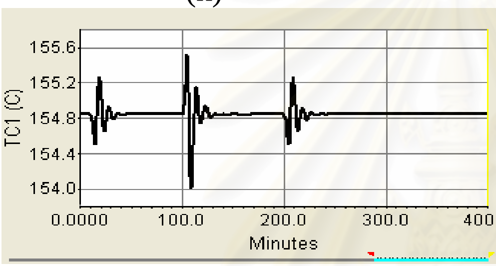

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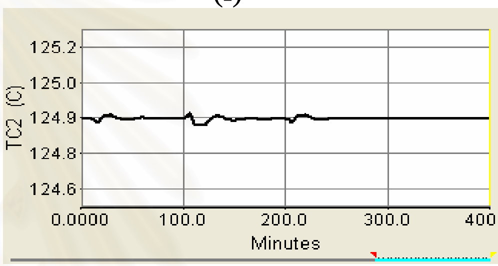

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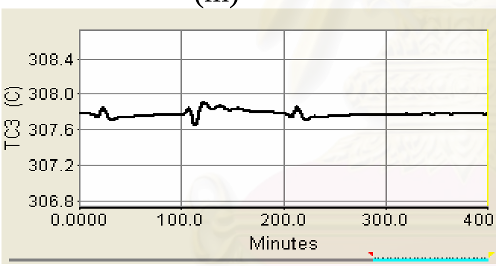

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Figure 6.14 (continue)

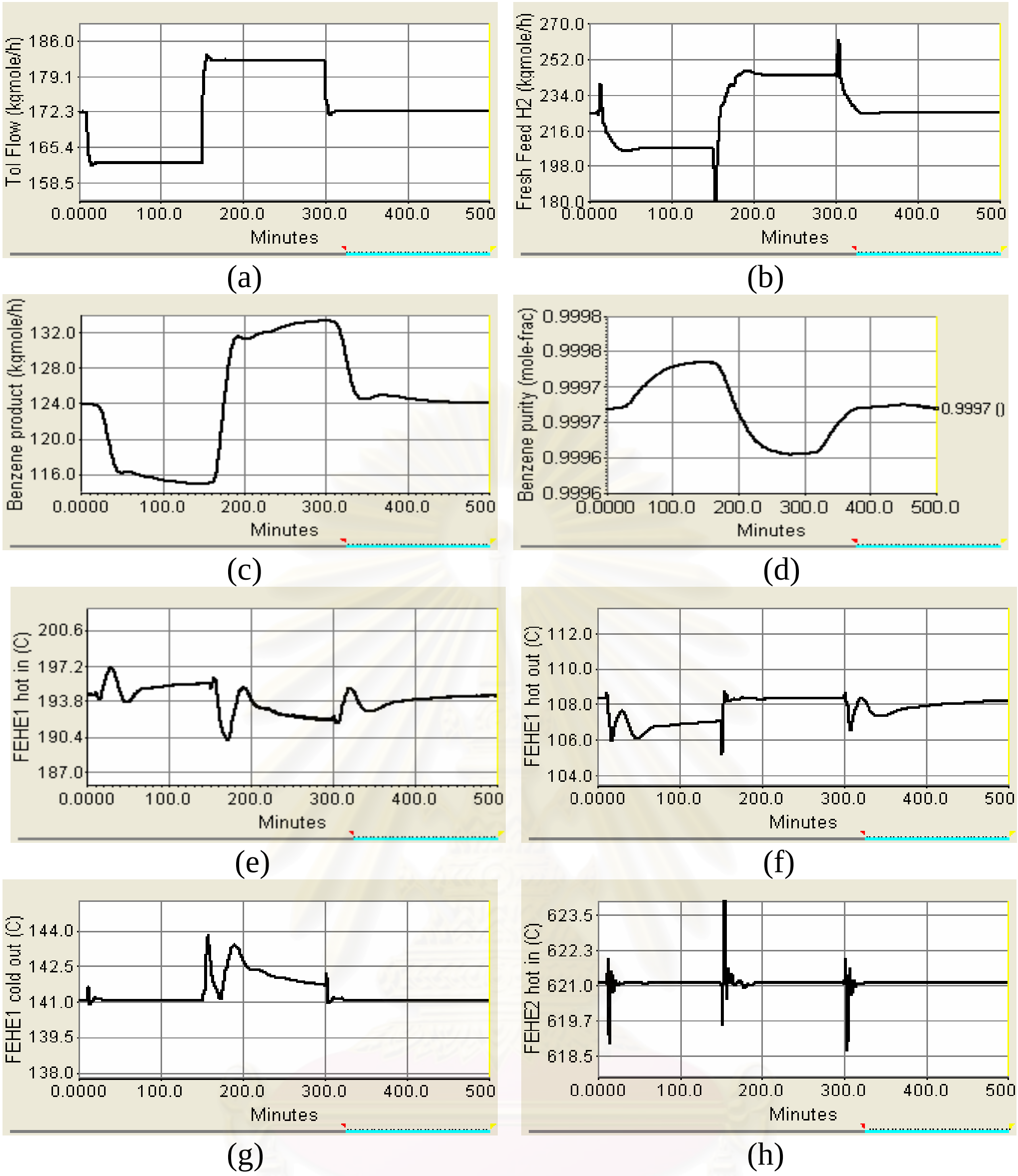
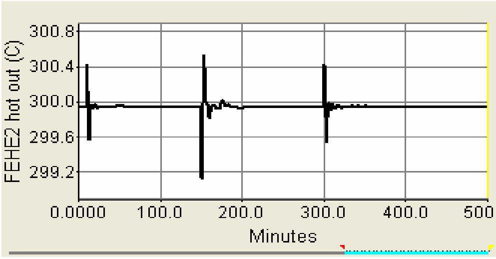


Figure 6.15 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) furnace duty, (k) cooler duty, (l) quench temperature, (m) reactor inlet temperature, (n) separator temperature, (o) stabilizer column tray temperature, (p) product column tray temperature, (q) recycle column tray temperature



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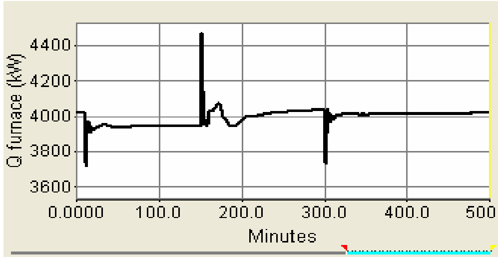

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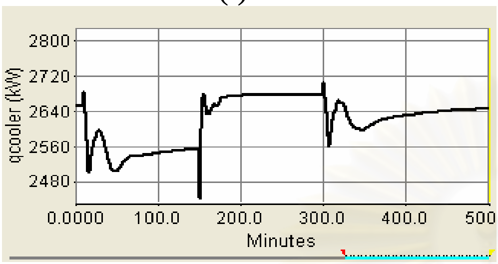

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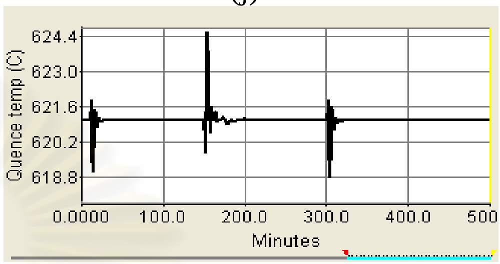

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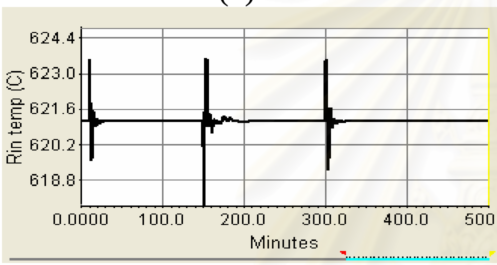
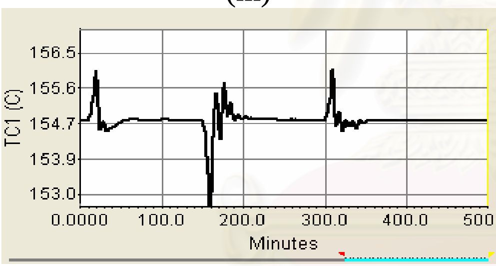


(m)

(n)



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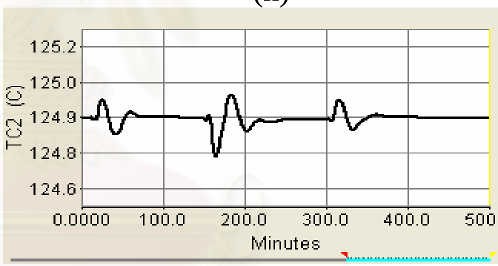

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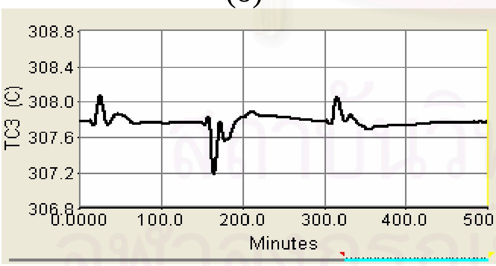


(q)

Figure 6.15 (continue)

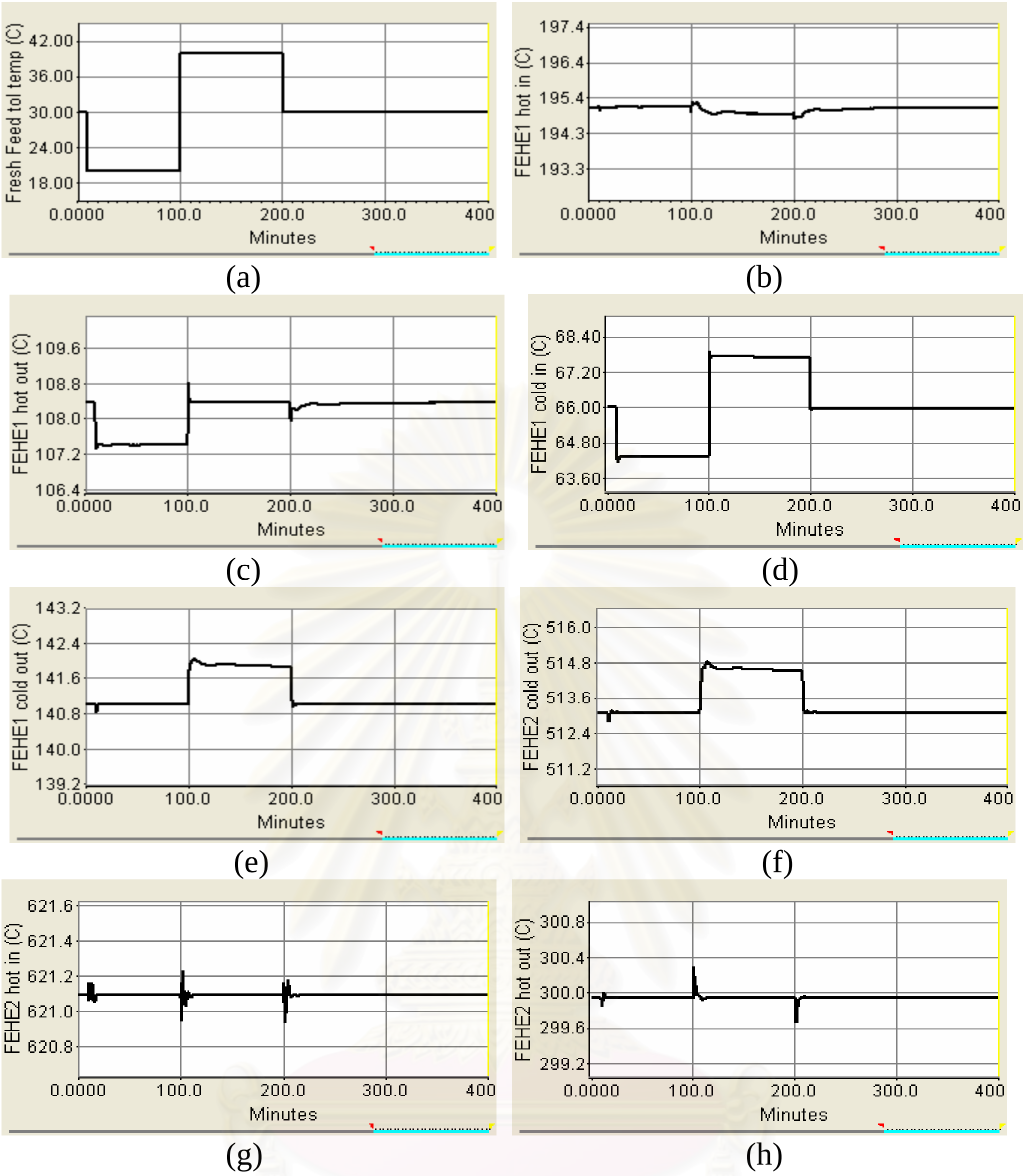


Figure 6.16 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature, (i) furnace duty, (j) cooler duty, (k) quench temperature, (l) reactor inlet temperature, (m) separator temperature, (n) stabilizer column tray temperature, (o) product column tray temperature, (p) recycle column tray temperature

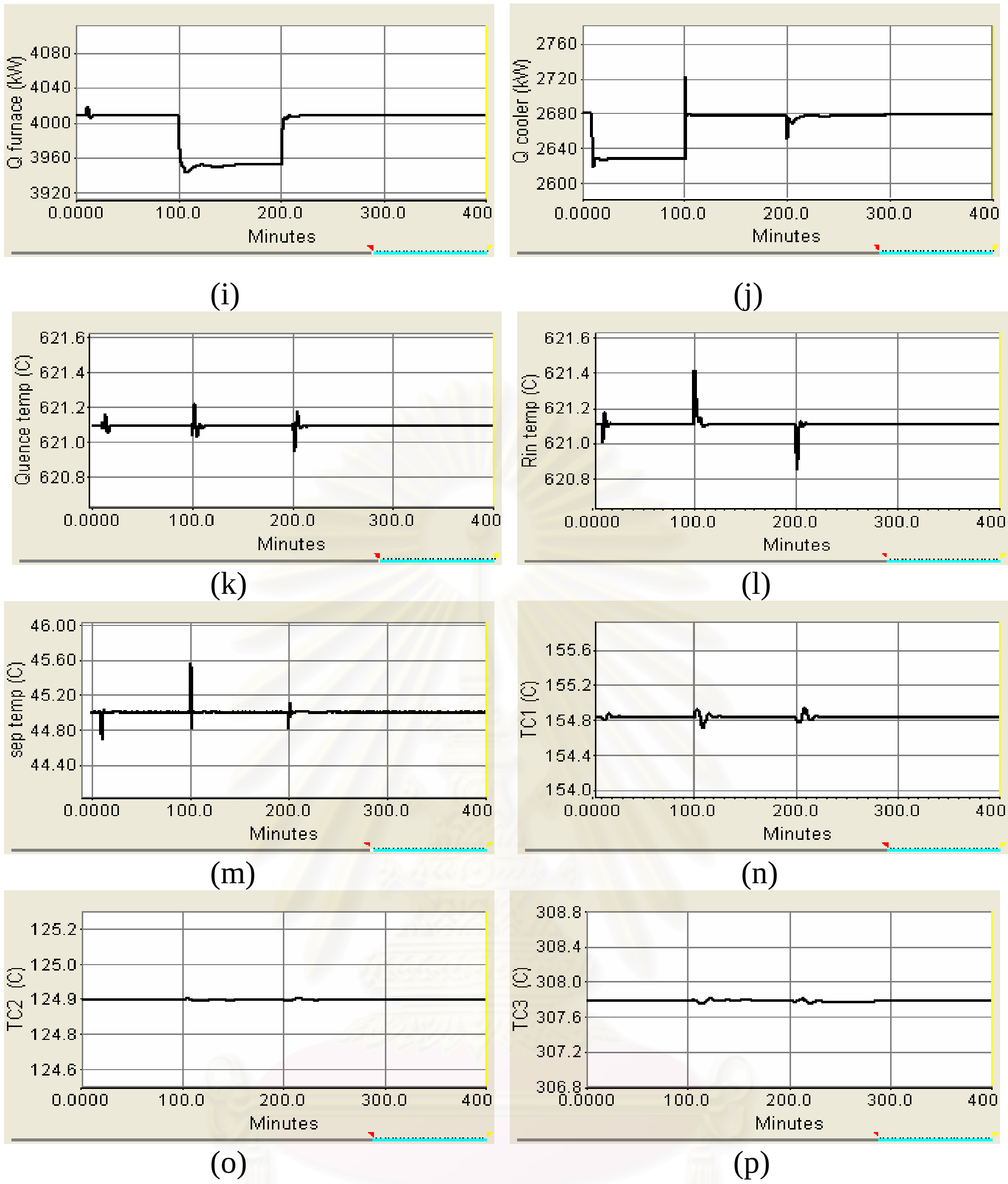


Figure 6.16 (continue)

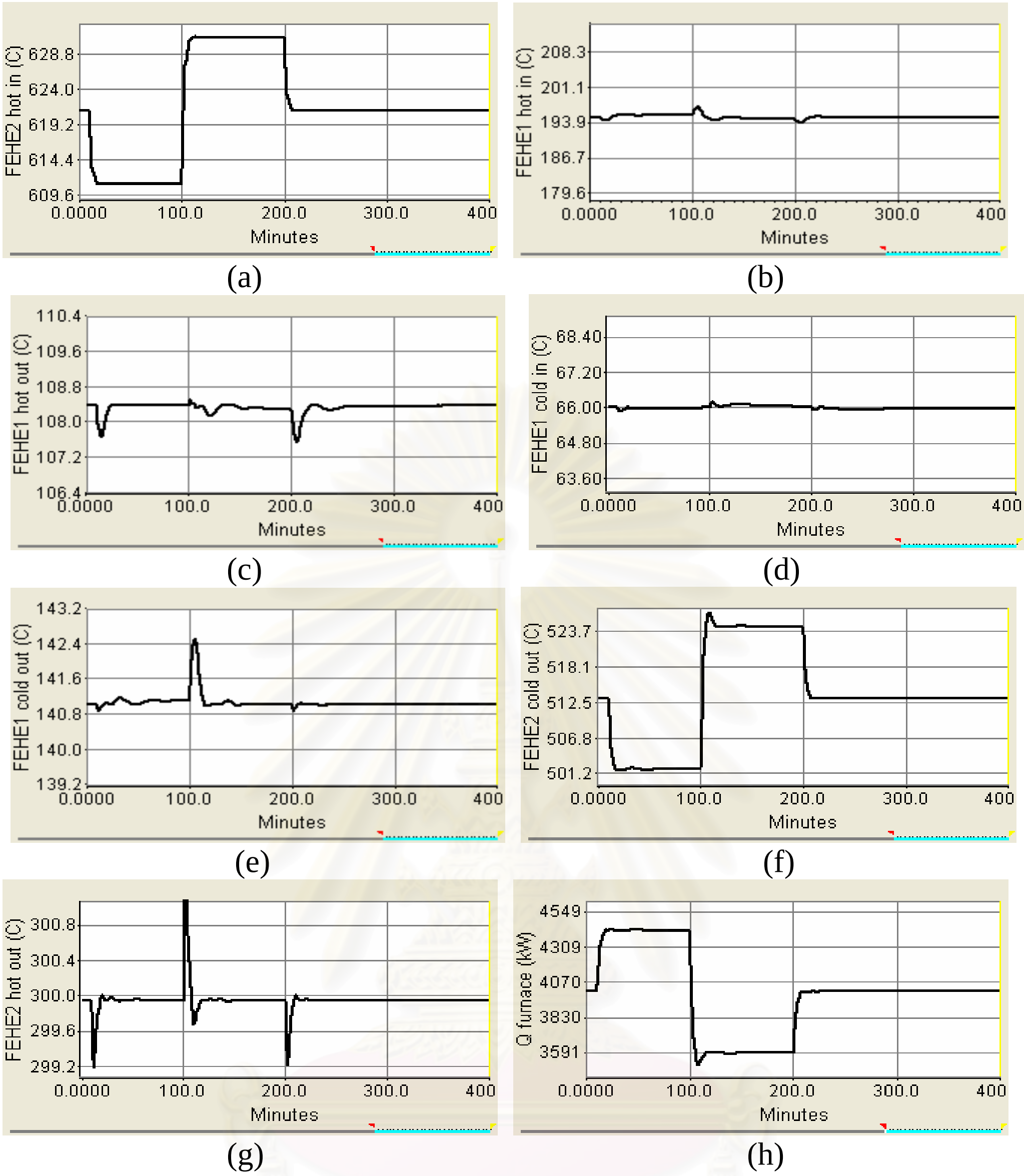
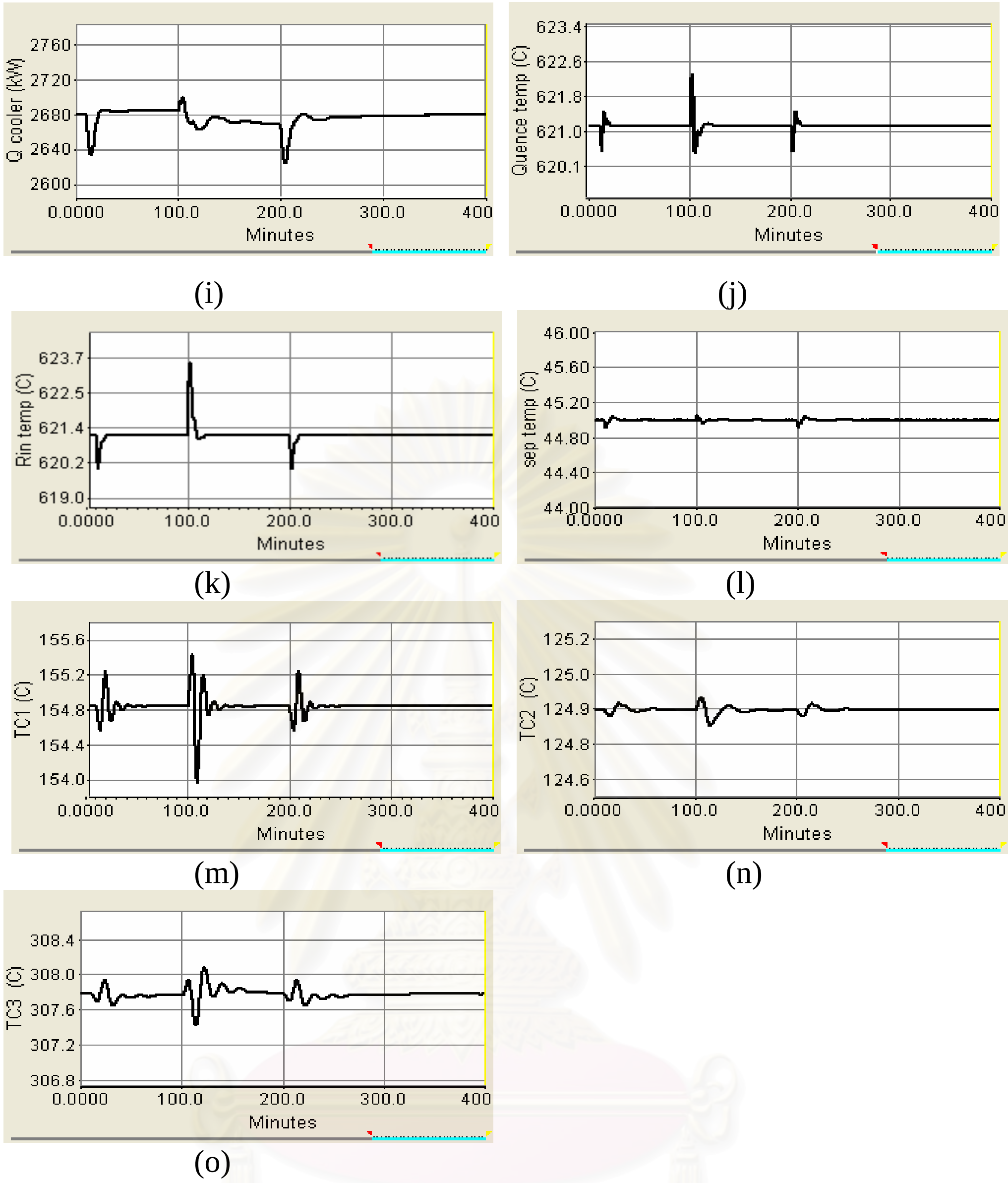


Figure 6.17 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot outlet temperature, (g) FEHE2 cold outlet temperature, (h) furnace duty, (i) cooler duty, (j) quench temperature, (k) reactor inlet temperature, (l) separator temperature, (m) stabilizer column tray temperature, (n) product column tray temperature, (o) recycle column tray temperature



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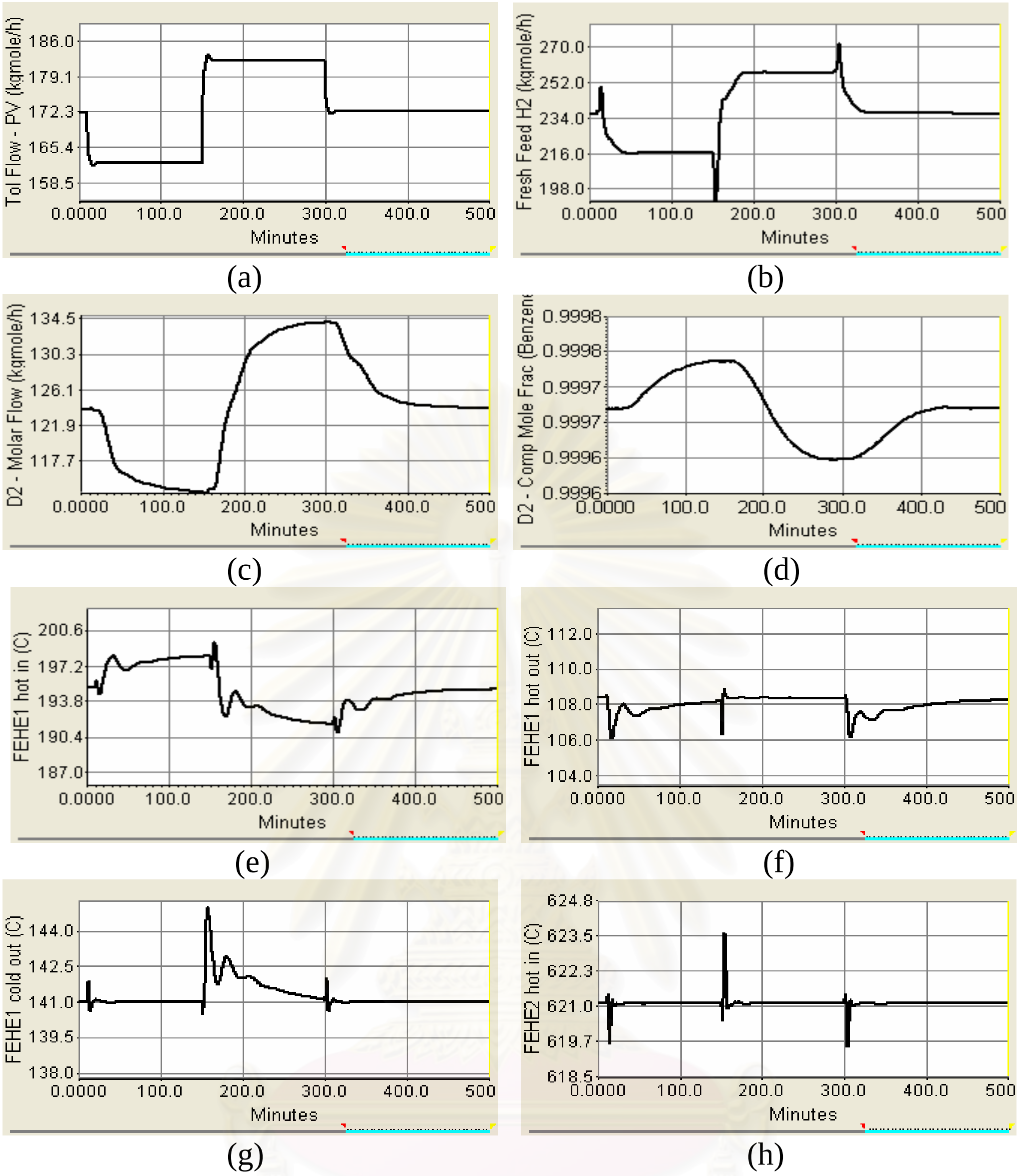
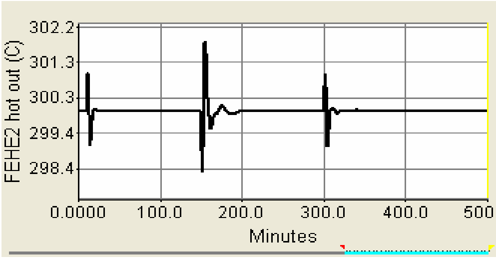


Figure 6.18 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates: CS2, where: (a) total toluene feed flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) furnace duty, (k) cooler duty, (l) quench temperature, (m) reactor inlet temperature, (n) separator temperature, (o) stabilizer column tray temperature, (p) product column tray temperature, (q) recycle column tray temperature



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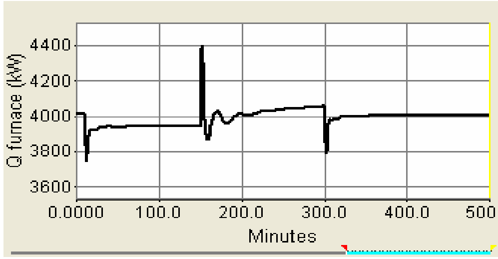

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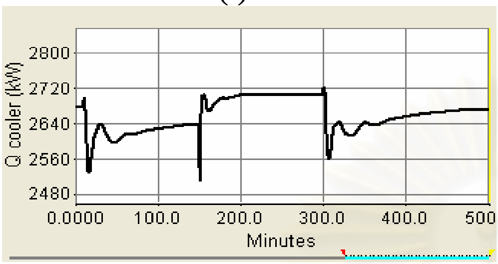

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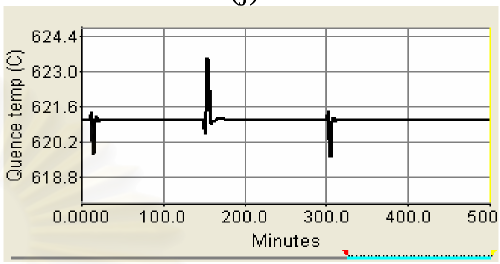

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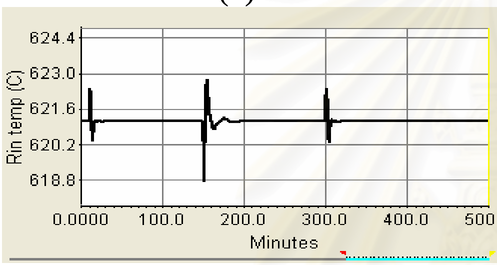

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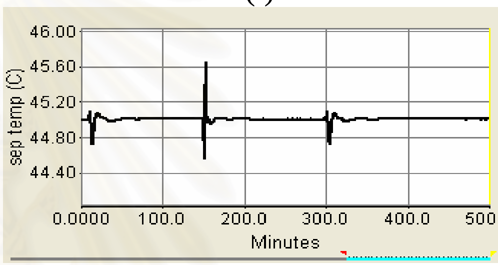

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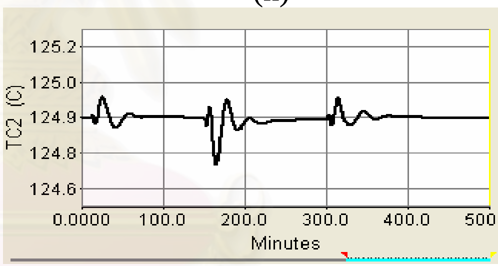
(o)



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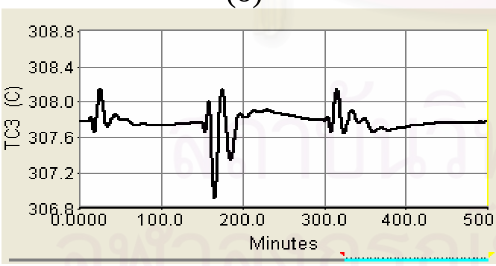

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Figure 6.18 (continue)

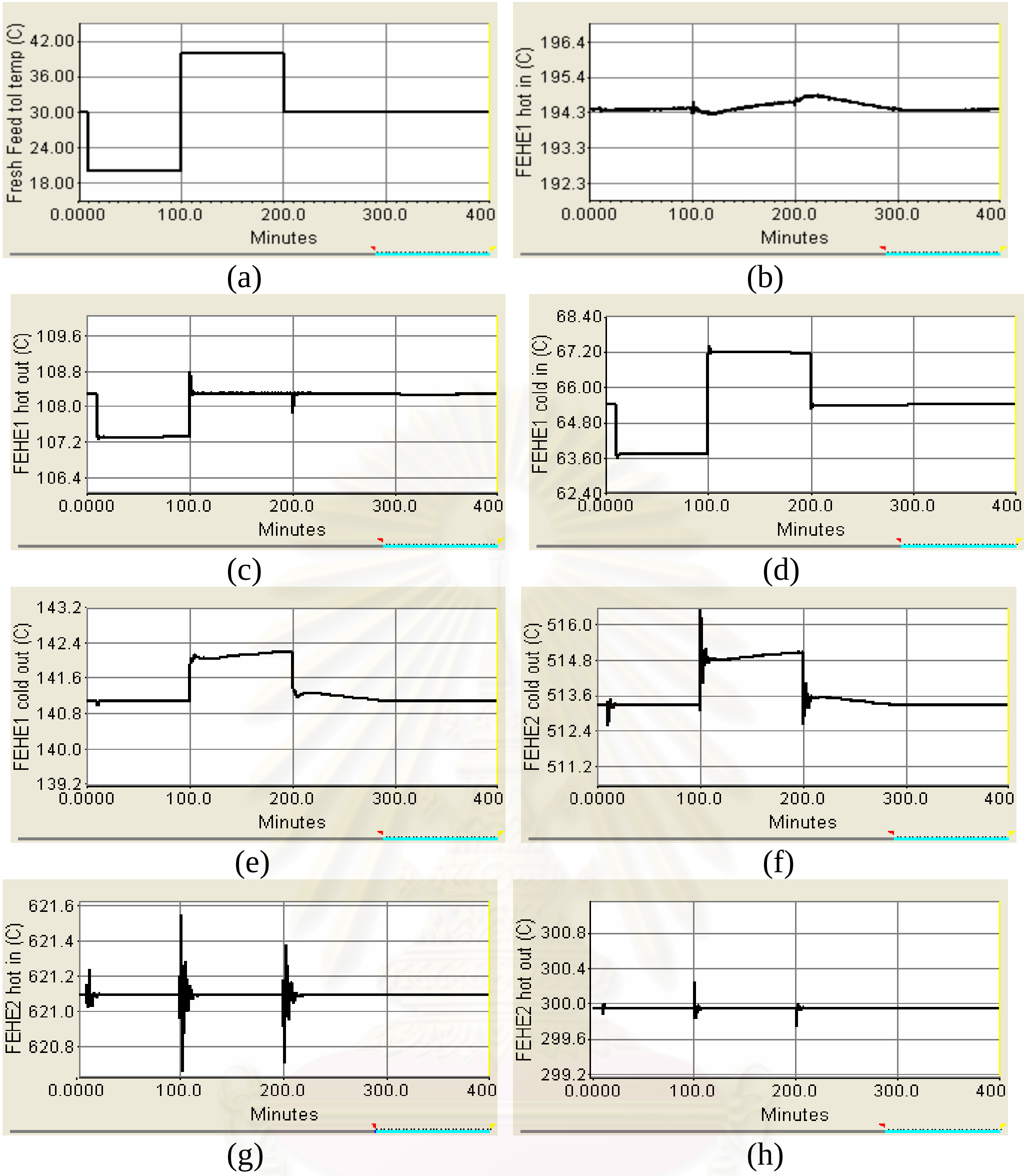
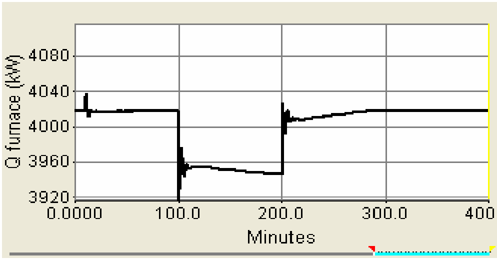


Figure 6.19 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature, (i) furnace duty, (j) cooler duty, (k) quench temperature, (l) reactor inlet temperature, (m) separator temperature, (n) stabilizer column tray temperature, (o) product column tray 12 temperature, (p) product column tray 17 temperature, (q) recycle column tray temperature



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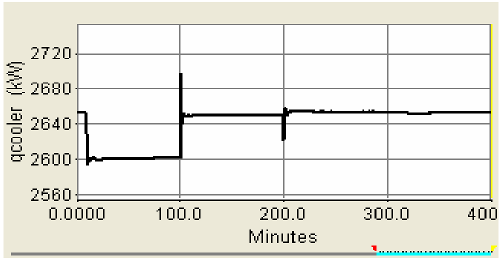

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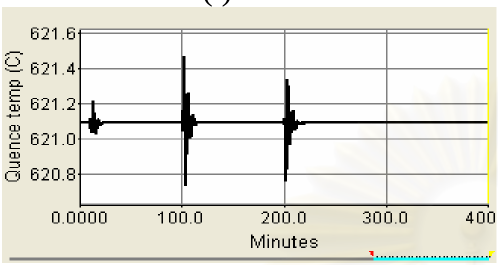

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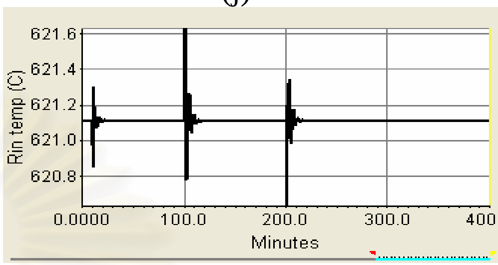

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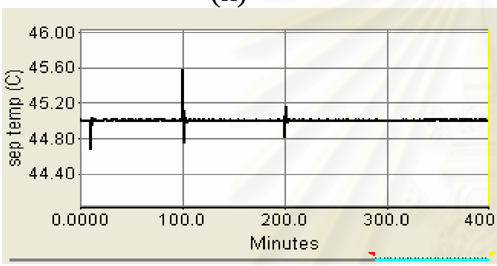

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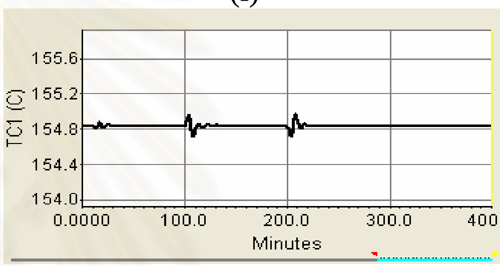

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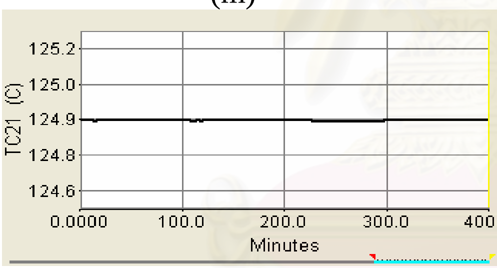

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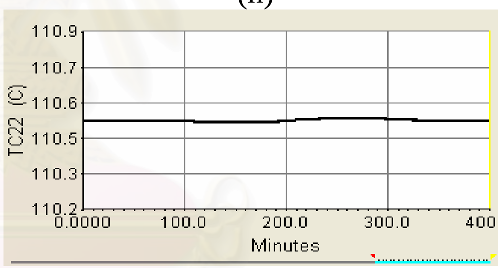

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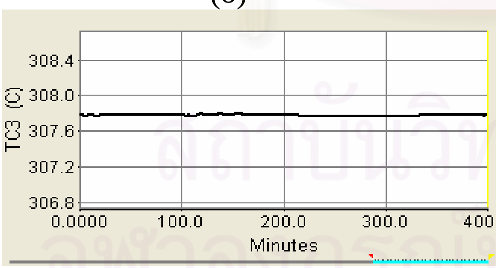

(q)

Figure 6.19 (continue)

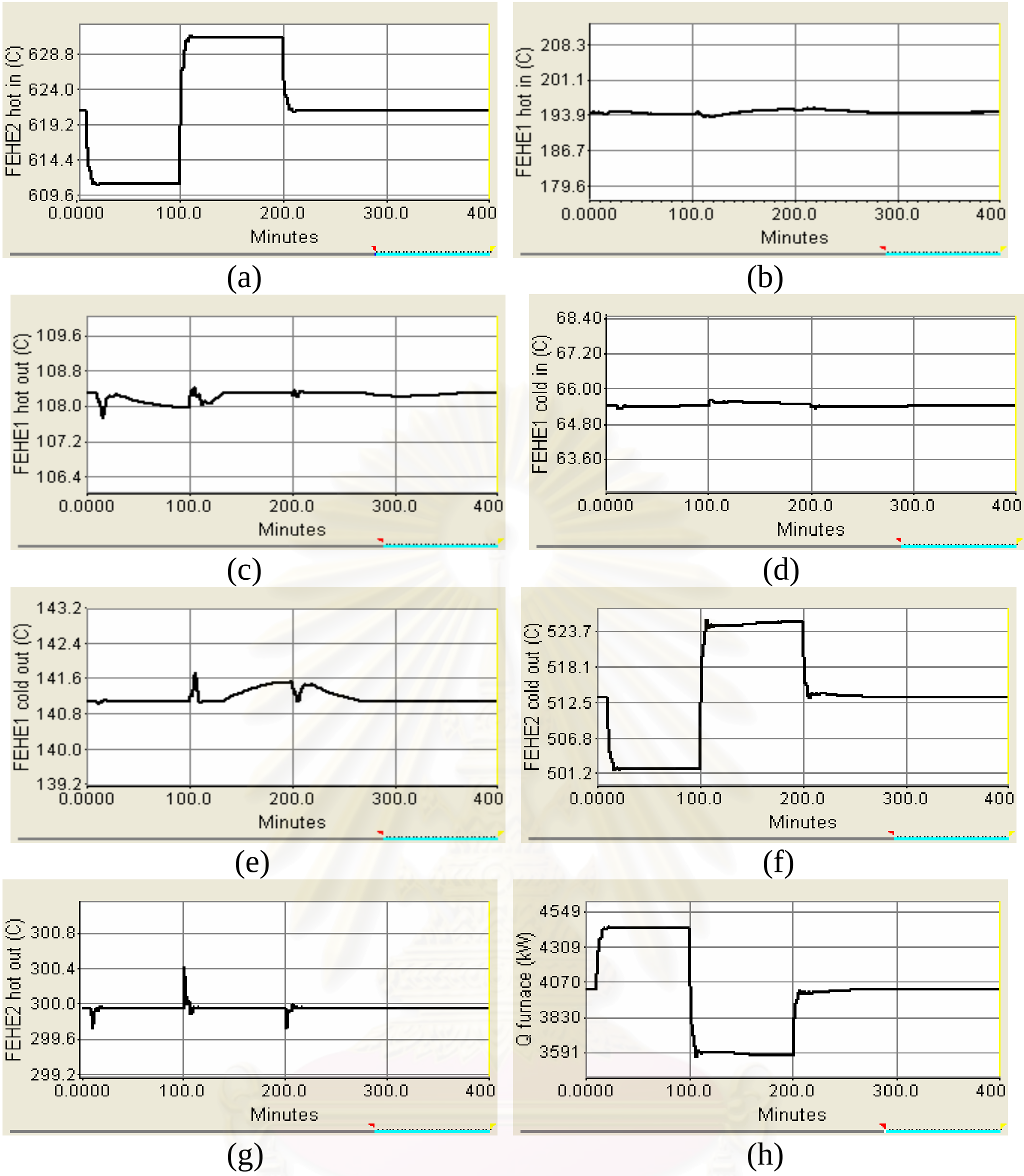

Figure 6.20 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot outlet temperature, (g) FEHE2 cold outlet temperature, (h) furnace duty, (i) cooler duty, (j) quench temperature, (k) reactor inlet temperature, (l) separator temperature, (m) stabilizer column tray temperature, (n) product column tray 12 temperature, (o) product column tray 17 temperature, (p) recycle column tray temperature

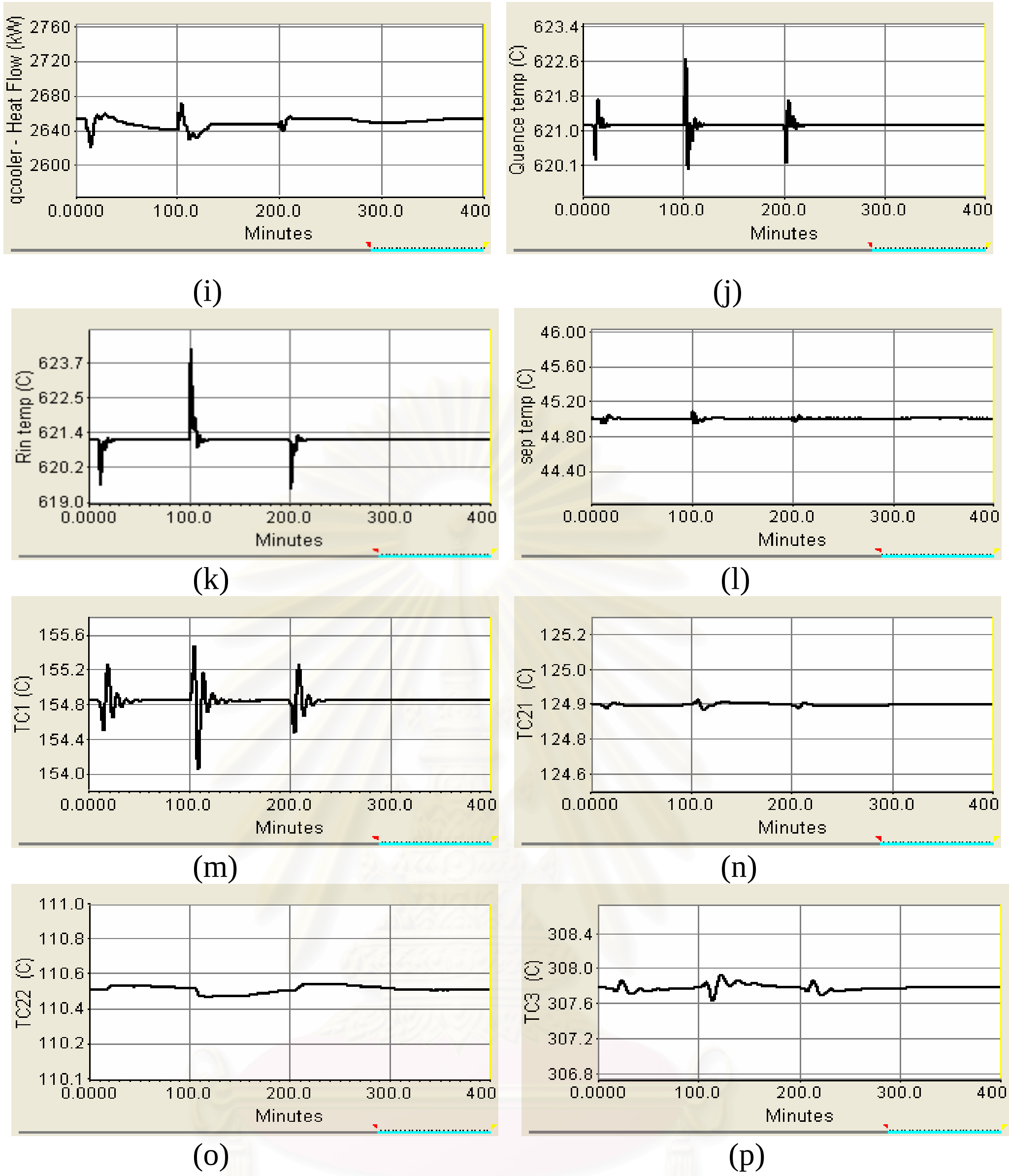

Figure 6.20 (continue)

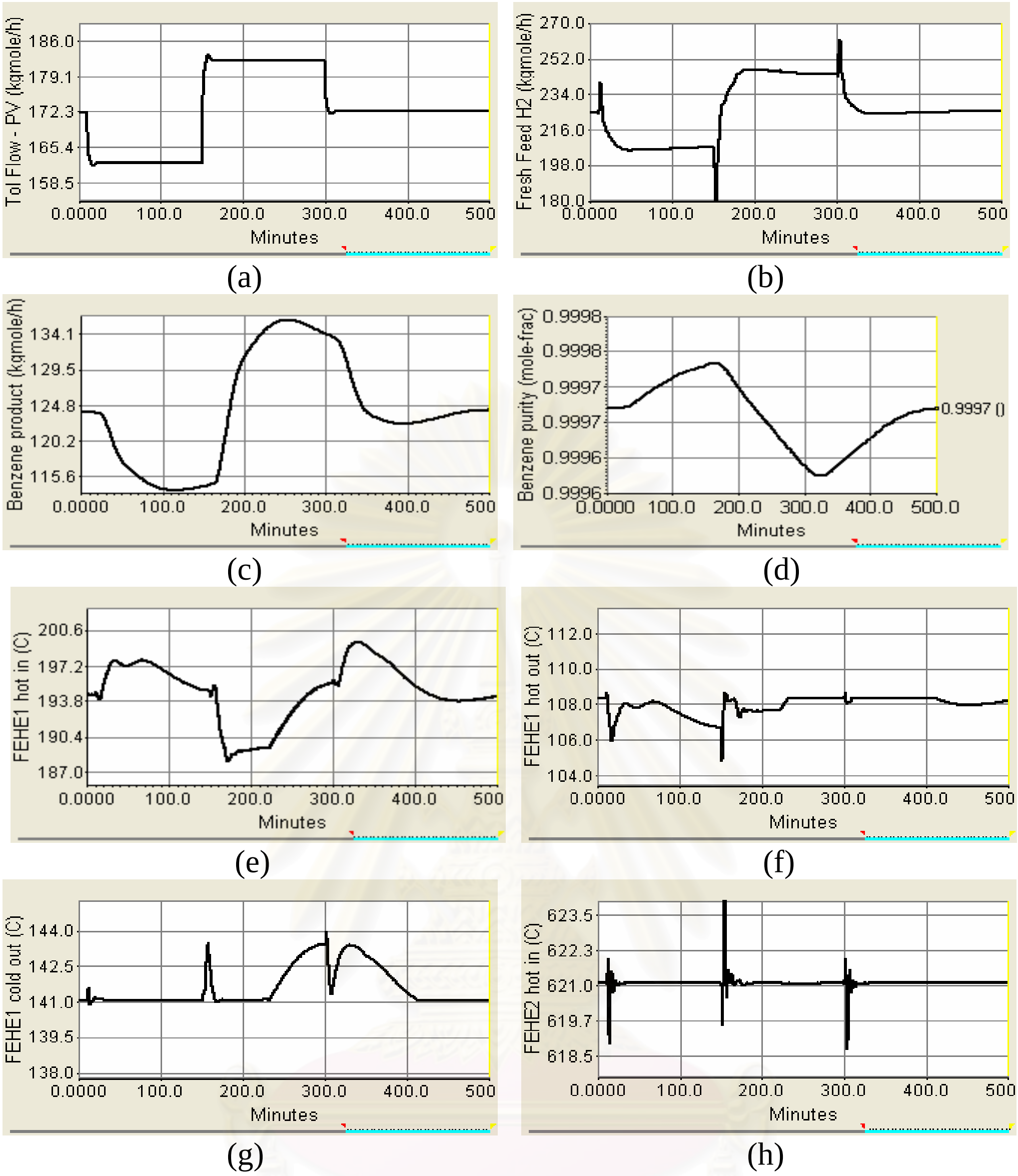

Figure 6.21 Dynamic Responses of the HDA Process Alternative 5 with three Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates: CS3, where: (a) total toluene feed flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) furnace duty, (k) cooler duty, (l) quench temperature, (m) reactor inlet temperature, (n) separator temperature, (o) stabilizer column tray temperature, (p) product column tray 12 temperature, (q) product column tray 17 temperature, (r) recycle column tray temperature

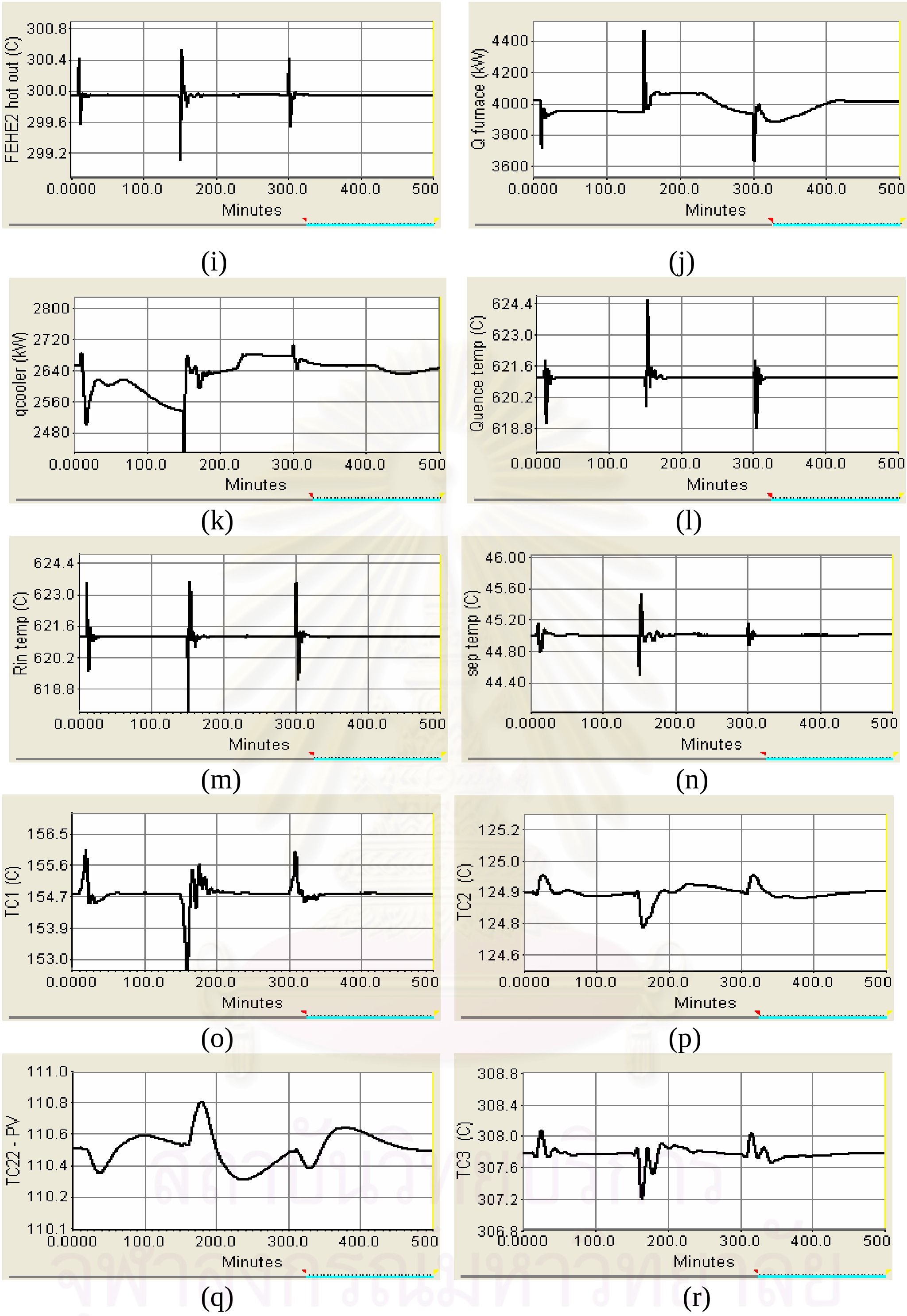

Figure 6.21 (continue)

6.4.2 Dynamic Simulation Results for HDA Process Alternative 5 with Minimum Auxiliary Utility Units

Three disturbance loads are also used to evaluate the dynamic performance of the control structures (CS1, CS2 and CS3) for HDA process alternative 5 with minimum auxiliary utility units.

6.4.2.1 Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream)

Figures 6.22, 6.25 and 6.28 show the dynamic responses of the control systems of HDA process alternative 5 with minimum auxiliary utility units to a change in the heat load disturbance of cold stream (reactor feed stream). In order to make this disturbance, first the fresh toluene feed temperature is decreased from 30 to 20^{o}C at time equals 10 minutes, and the temperature is increased from 20 to 40^{o}C at time equals 100 minutes, then its temperature is returned to its nominal value of 30^{o}C at time equals 200 minutes (Figures 6.22.a, 6.25.a and 6.28.a).

As can be seen, the dynamic responses for HDA process alternative 5 with minimum auxiliary utility units and with three auxiliary utility units are same in CS1, CS2 and CS3.

6.4.2.2 Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream)

Figure 6.23, 6.26 and 6.29 shows the dynamic responses of the control systems of HDA process alternative 5 with three auxiliary utility units to a change in the heat load disturbance of hot stream (the hot reactor product). This disturbance is made as follows: first the set point of FEHE3-hot-inlet temperature controller (i.e. TCX1 in Figure 6.4 6.5 and 6.6) is decreased from 621.1 to 611.1oC at time equals 10 minutes, and the temperature is increased from 611.1 to 631.1oC at time equals 100 minutes, then its temperature is returned to its nominal value of 621.1oC at time equals 200 minutes (Figure 6.23.a, 6.26.a and 6.29.a). As can be seen, this temperature response is very fast (Figure 6.23.a, 6.26.a and 6.29.a), the new steady state is reached quickly (Figure 6.23.a, 6.26.a and 6.29.a).

As can be seen, the dynamic responses for HDA process alternative 5 with minimum auxiliary utility units and with three auxiliary utility units are same in CS1, CS2 and CS3.

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6.4.2.3 Change in the Total Toluene Feed Flowrates

Figure 6.24, 6.27 and 6.23 shows the dynamic responses of the control systems of HDA process alternative 5 with three auxiliary utility units to a change in the total toluene flowrates. This disturbance is made by decreasing toluene flowrates from 172.3 to 162.3 kgmole/h at time equals 10 minutes, and the flowrates is increased from 162.3 to 182.3 kgmole/h at time equals 100 minutes, then its flowrates is returned to its nominal value of 172.3 kgmole/h at time equals 200 (Figure 6.24.a , 6.27.a and 6.23.a).

As can be seen, the dynamic responses for HDA process alternative 5 with minimum auxiliary utility units and with three auxiliary utility units are same in CS1, CS2 and CS3.

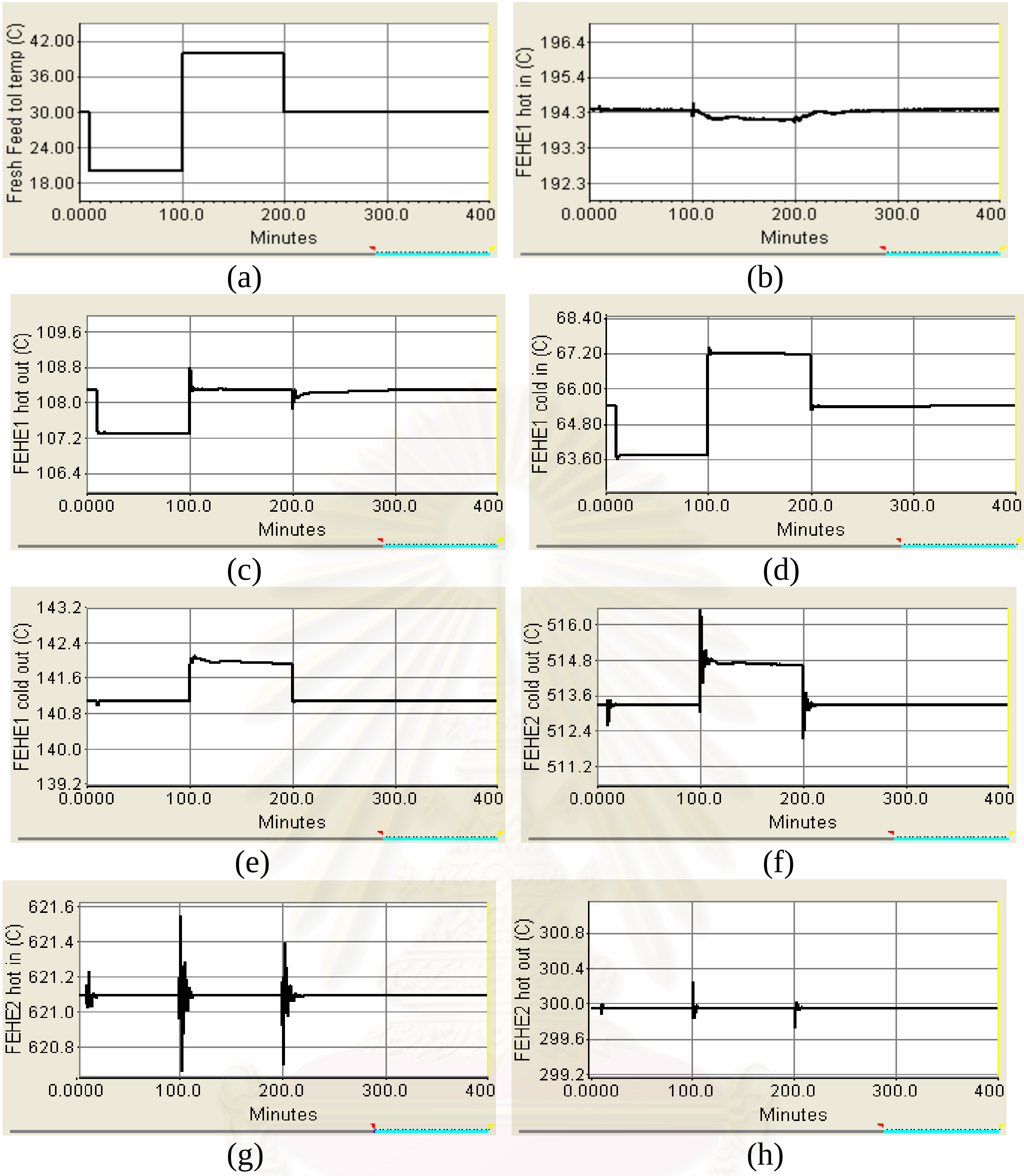


Figure 6.22 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature, (i) furnace duty, (j) cooler duty, (k) quench temperature, (l) reactor inlet temperature, (m) separator temperature, (n) stabilizer column tray temperature, (o) product column tray temperature, (p) recycle column tray temperature

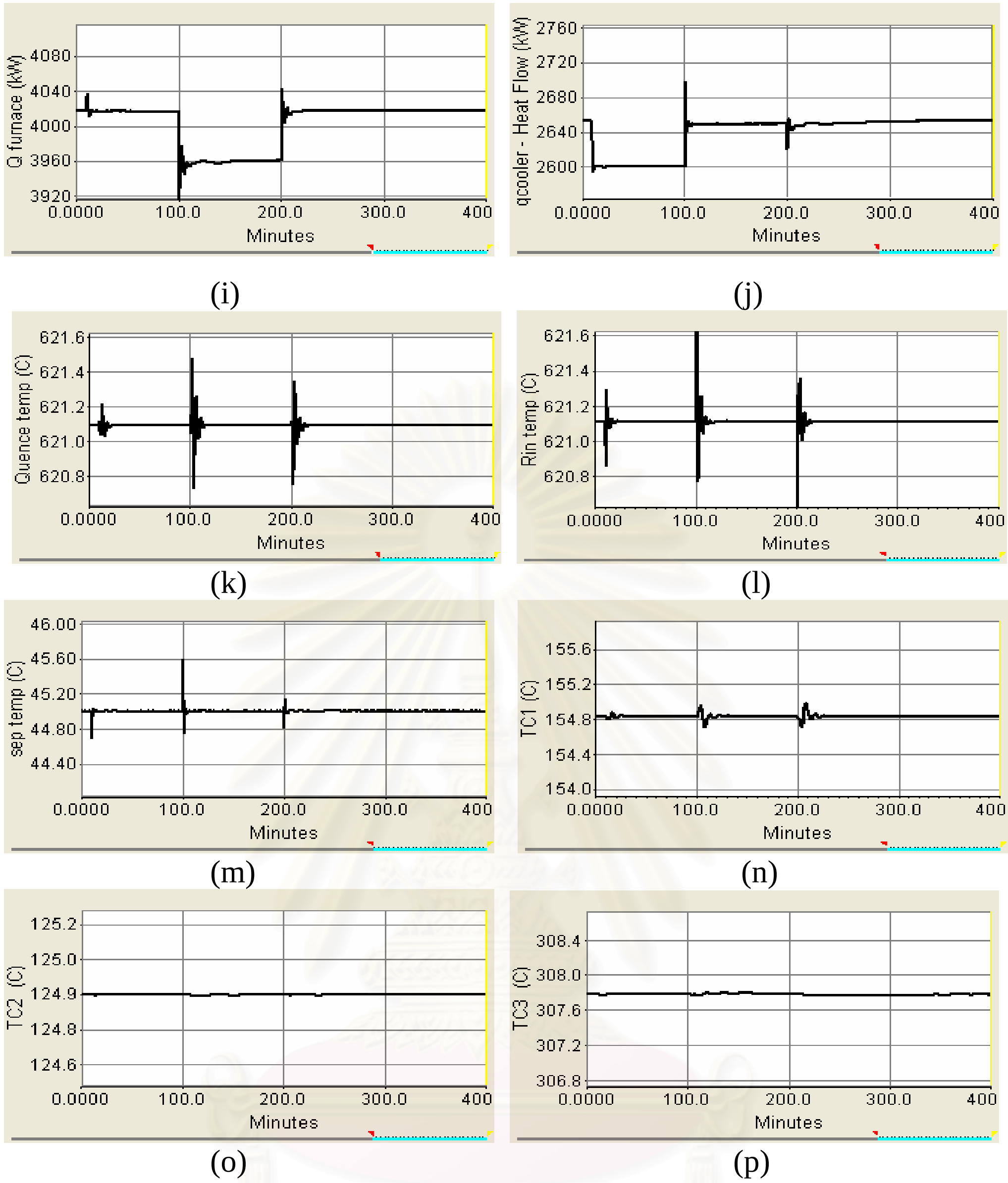


Figure 6.22 (continue)

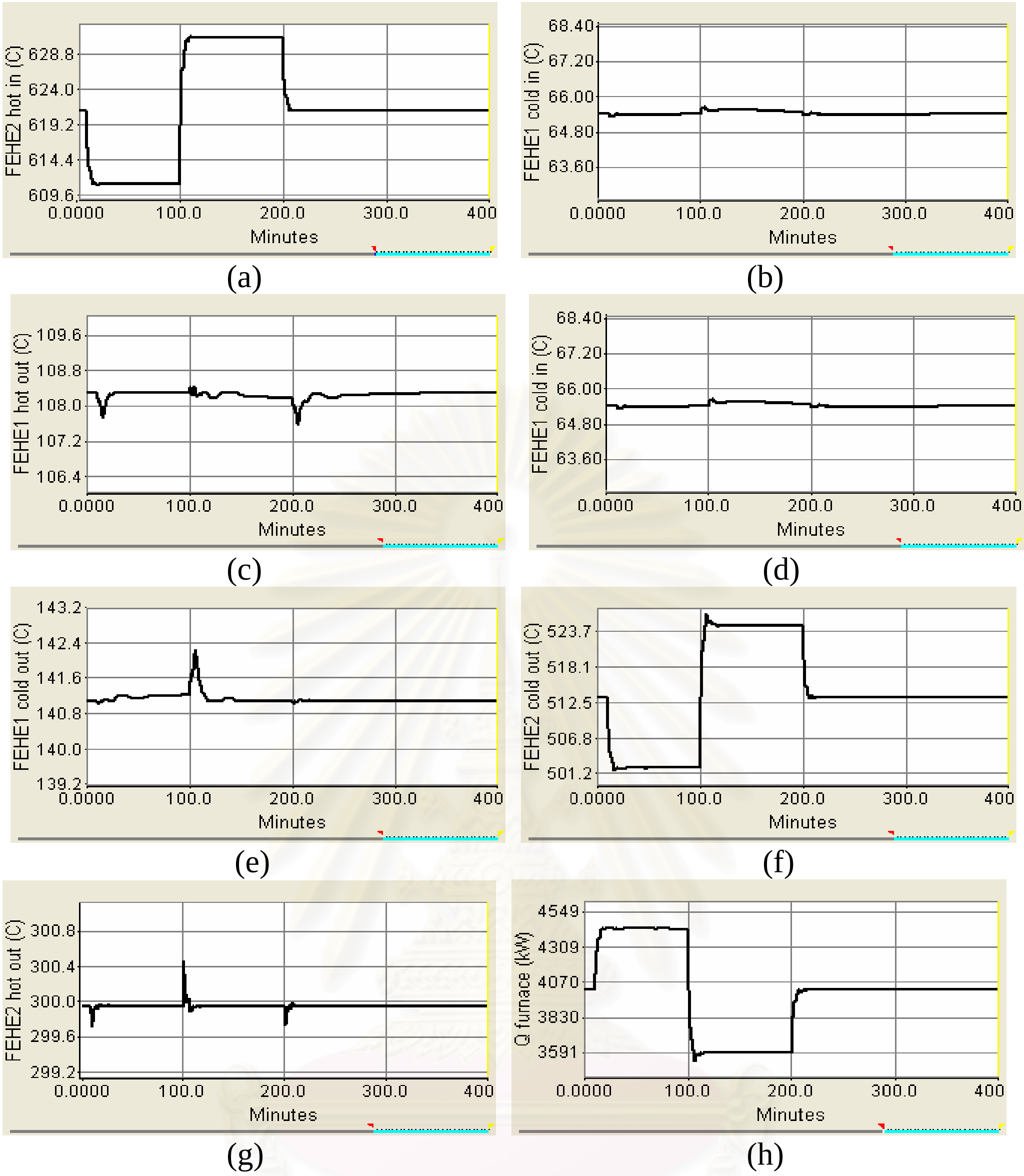
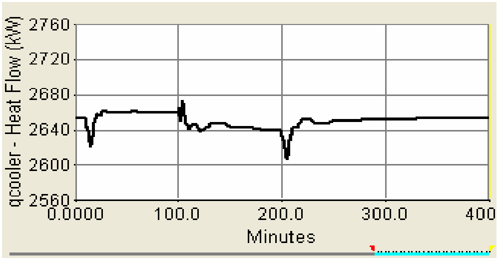


Figure 6.23 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot outlet temperature, (g) FEHE2 cold outlet temperature, (h) furnace duty, (i) cooler duty, (j) quench temperature, (k) reactor inlet temperature, (l) separator temperature, (m) stabilizer column tray temperature, (n) product column tray temperature, (o) recycle column tray temperature



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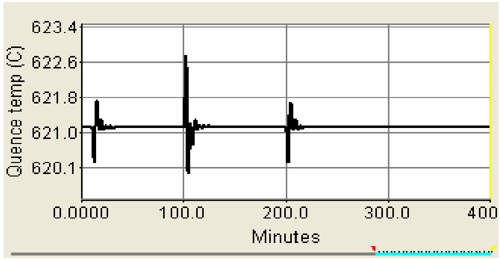

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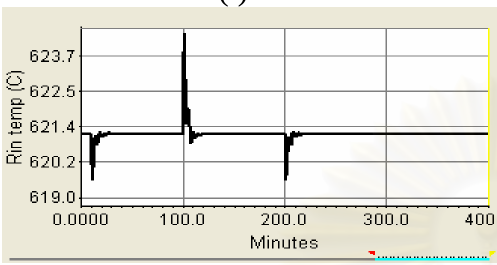

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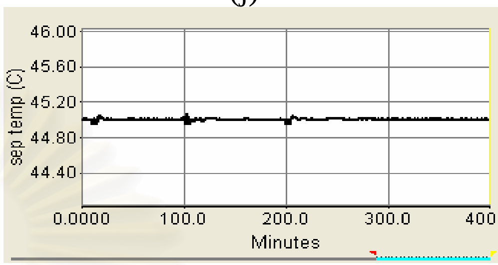

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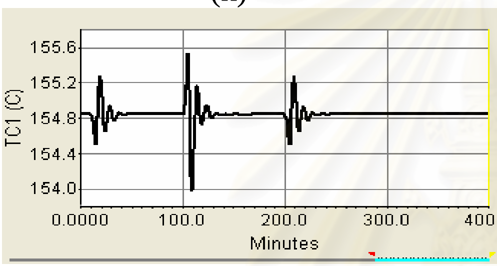

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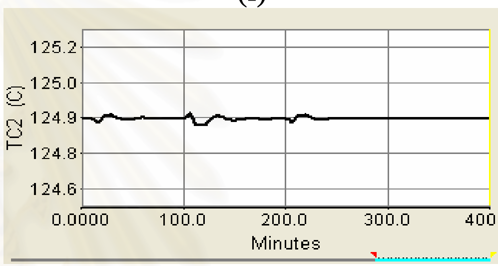

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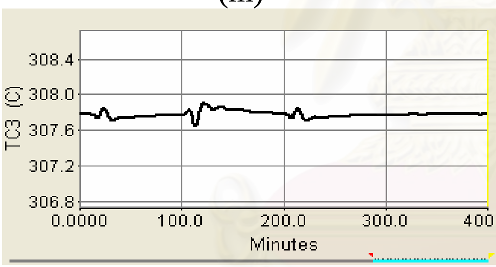


(o)

Figure 6.23 (continue)

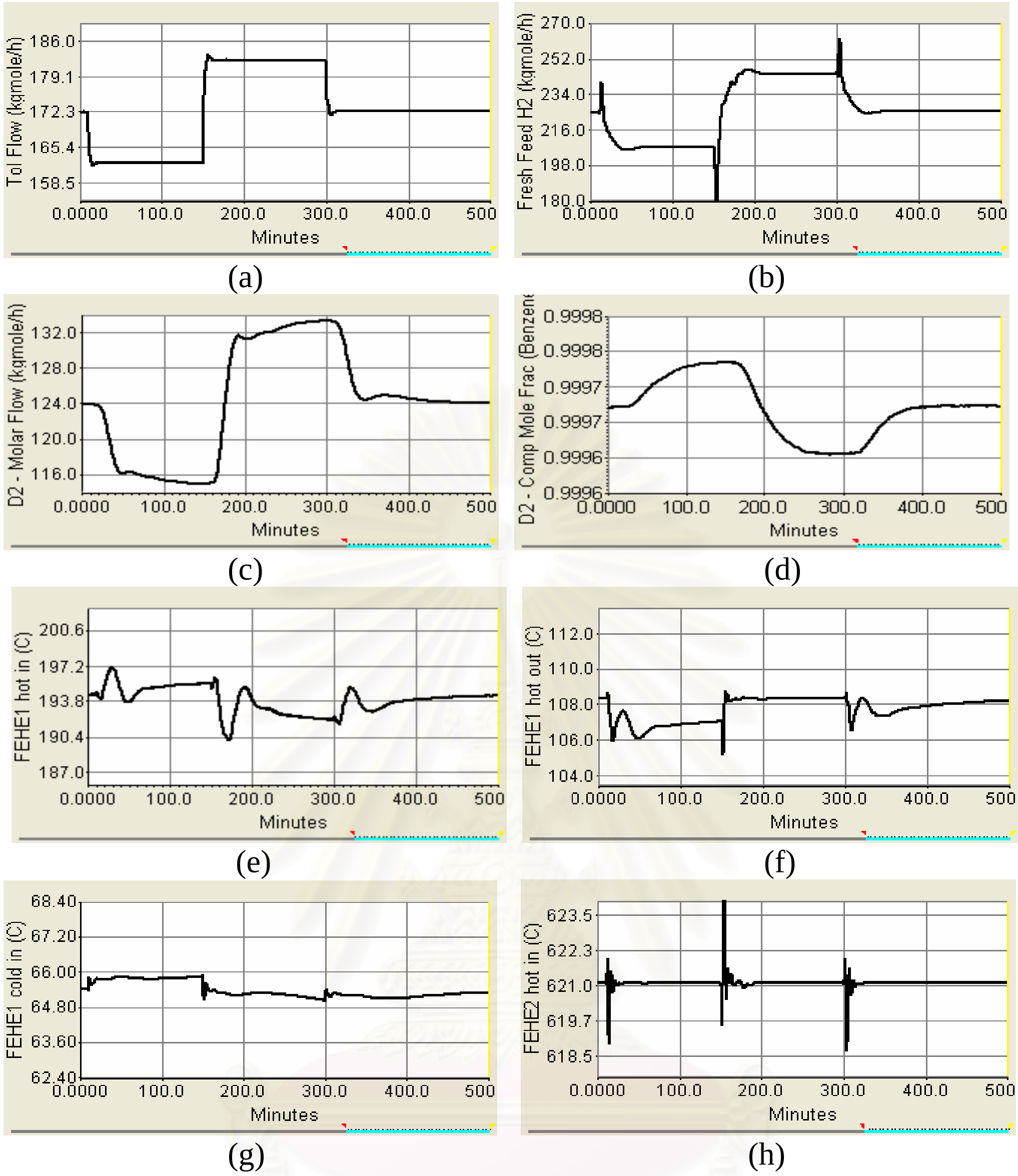
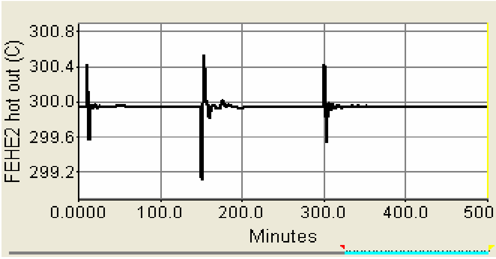


Figure 6.24 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) furnace duty, (k) cooler duty, (l) quench temperature, (m) reactor inlet temperature, (n) separator temperature, (o) stabilizer column tray temperature, (p) product column tray temperature, (q) recycle column tray temperature



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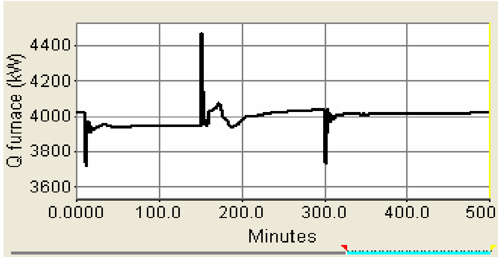

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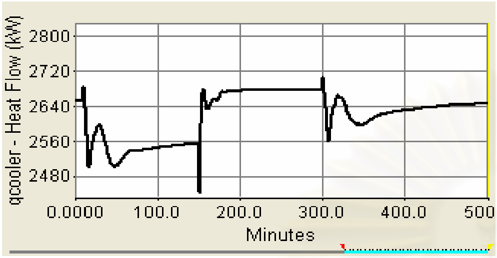

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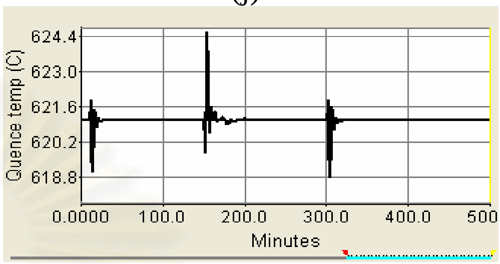

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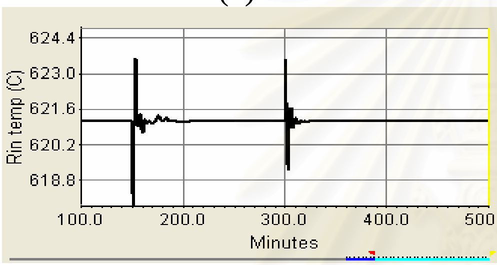

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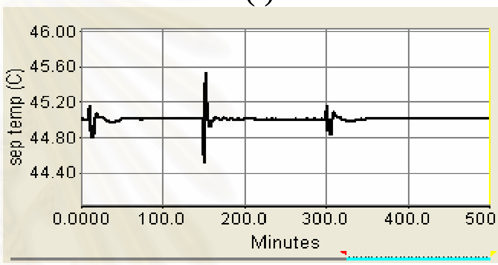

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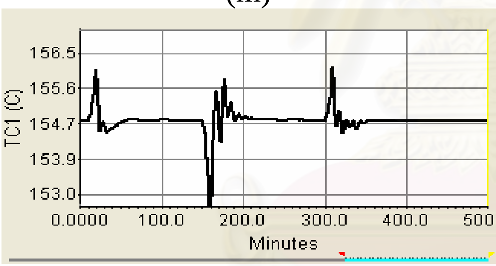

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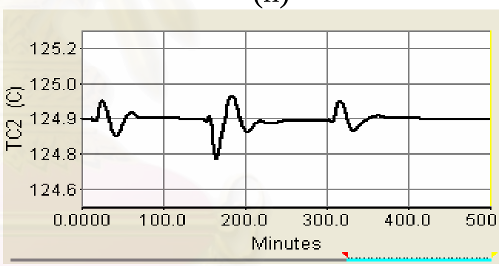

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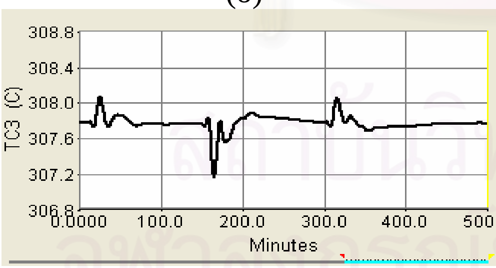


(s)

Figure 6.24 (continue)

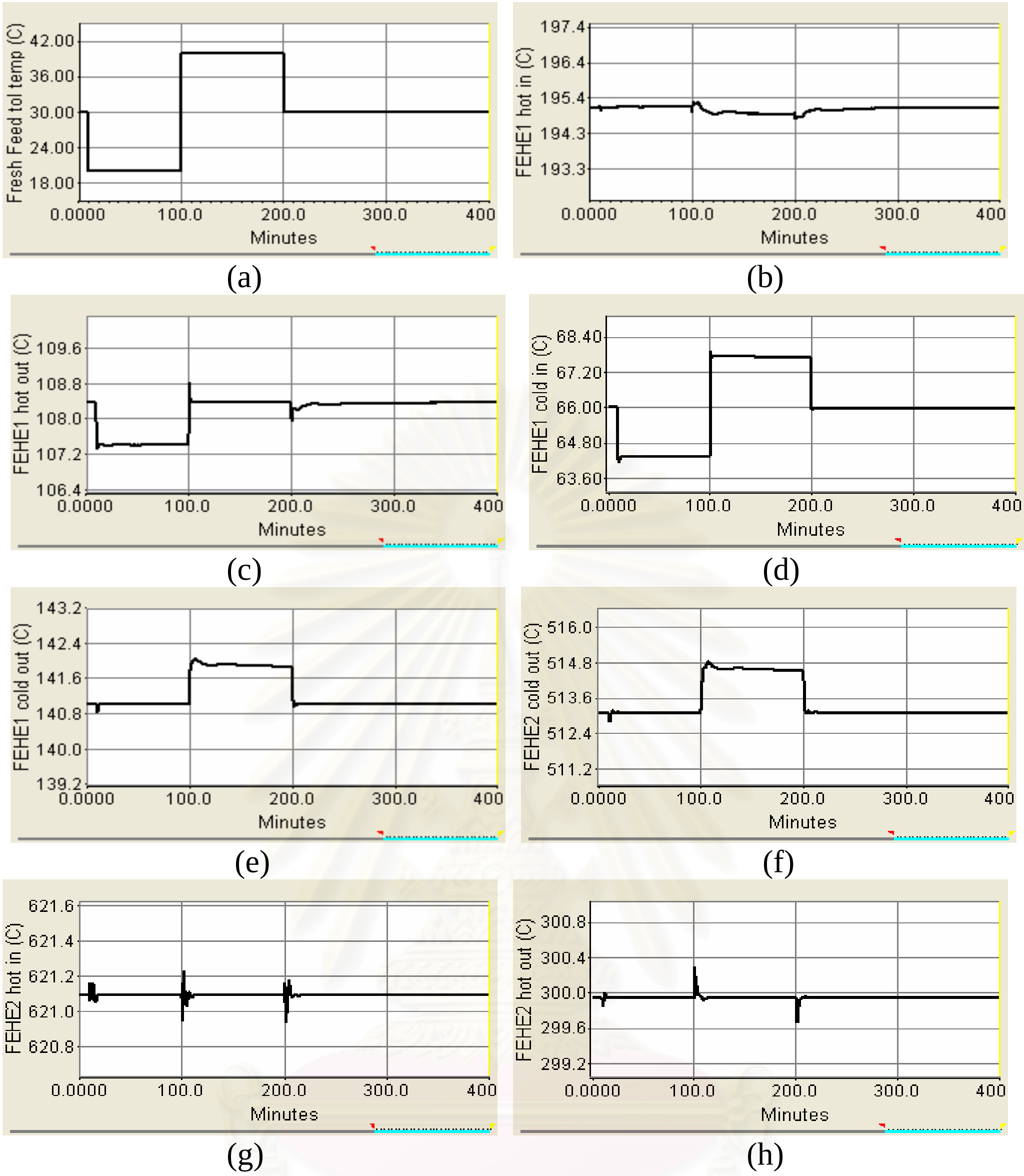


Figure 6.25 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature, (i) furnace duty, (j) cooler duty, (k) quench temperature, (l) reactor inlet temperature, (m) separator temperature, (n) stabilizer column tray temperature, (o) product column tray temperature, (p) recycle column tray temperature

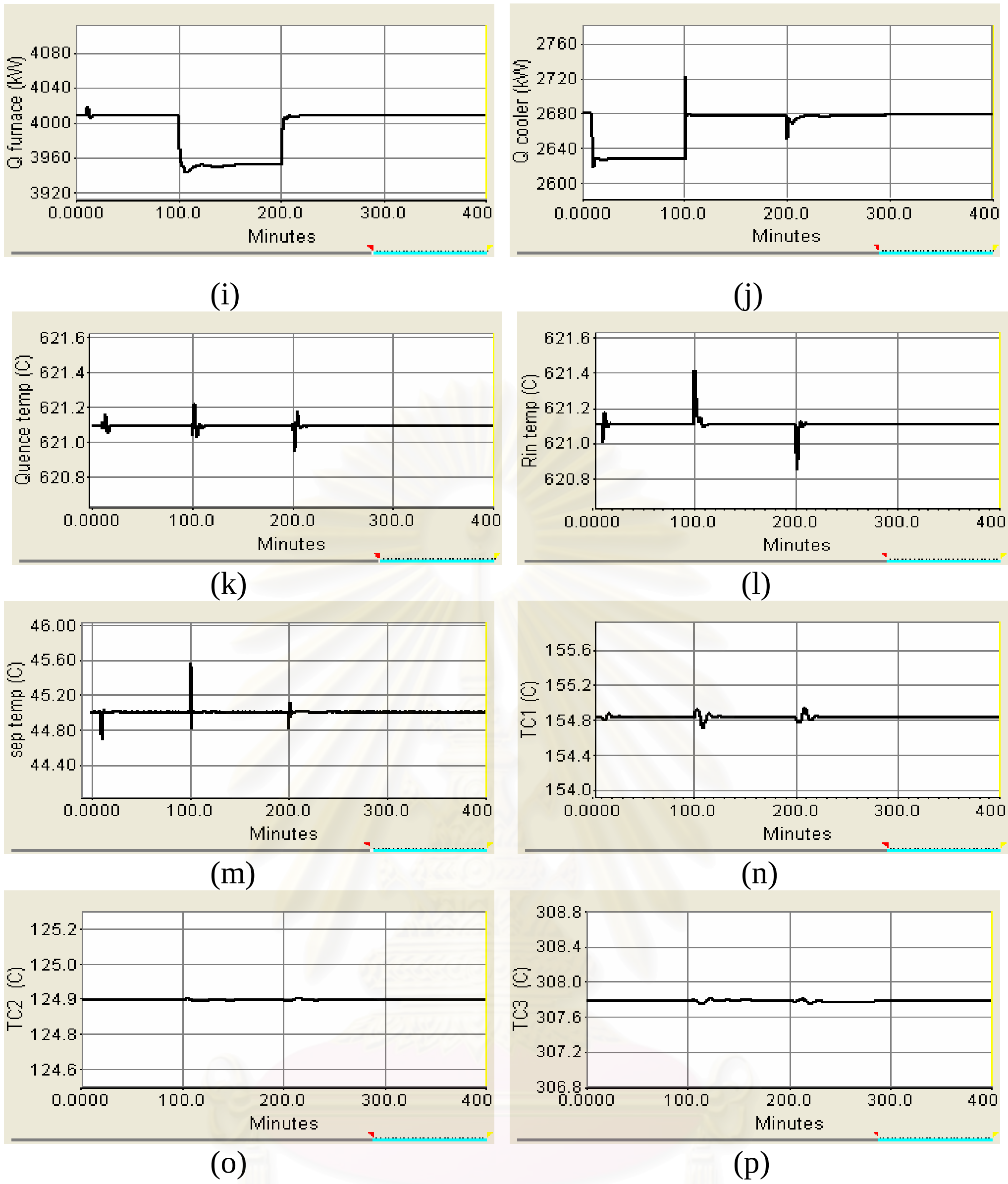


Figure 6.25 (continue)

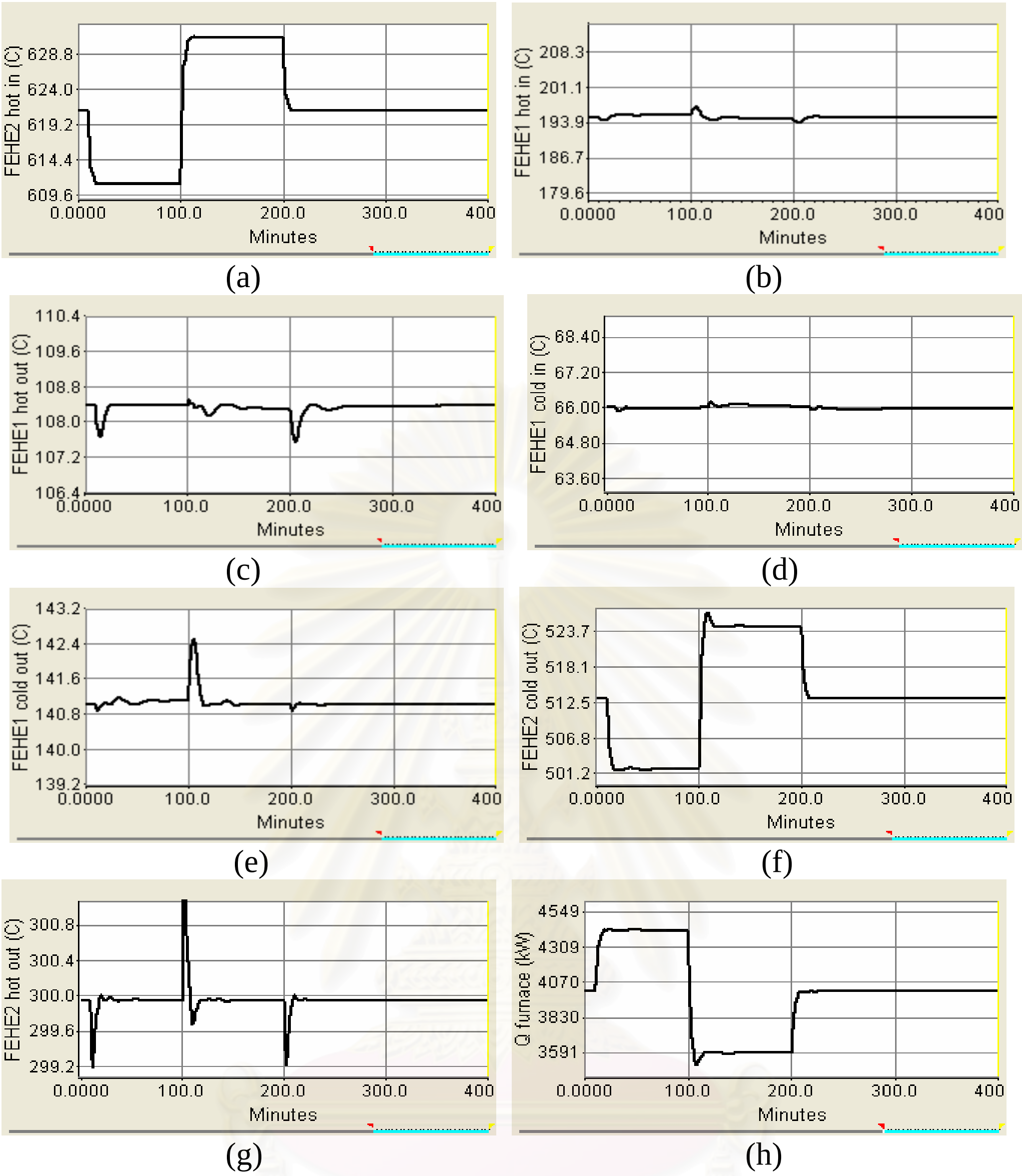
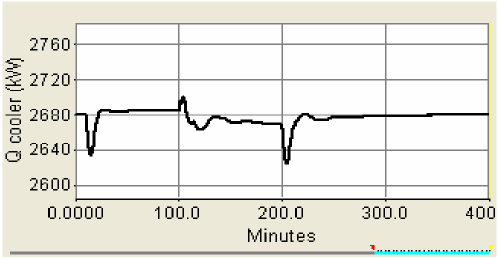


Figure 6.26 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot outlet temperature, (g) FEHE2 cold outlet temperature, (h) furnace duty, (i) cooler duty, (j) quench temperature, (k) reactor inlet temperature, (l) separator temperature, (m) stabilizer column tray temperature, (n) product column tray temperature, (o) recycle column tray temperature



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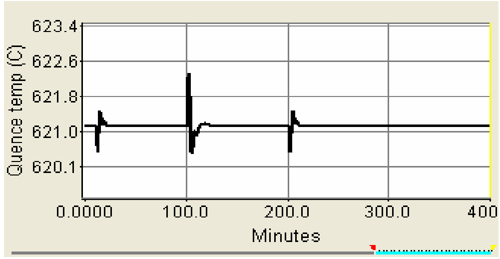

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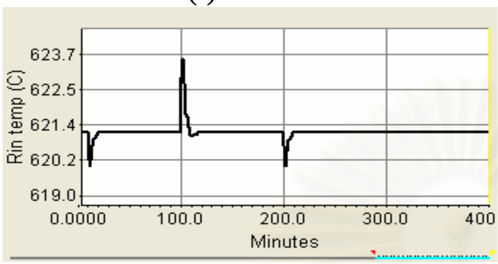

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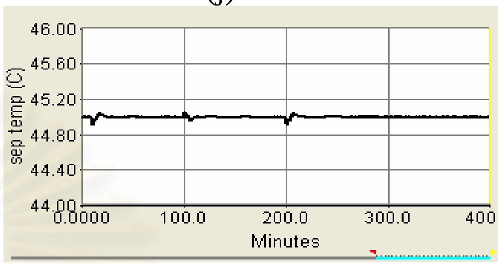

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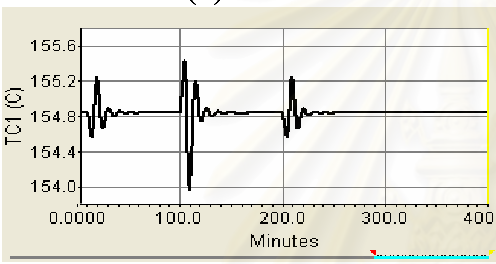

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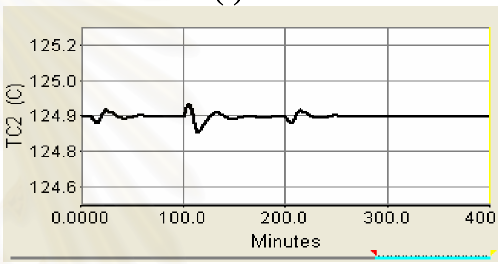

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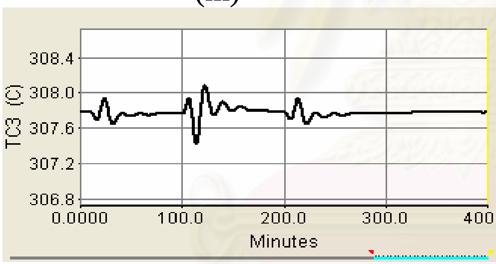


(o)

Figure 6.26 (continue)

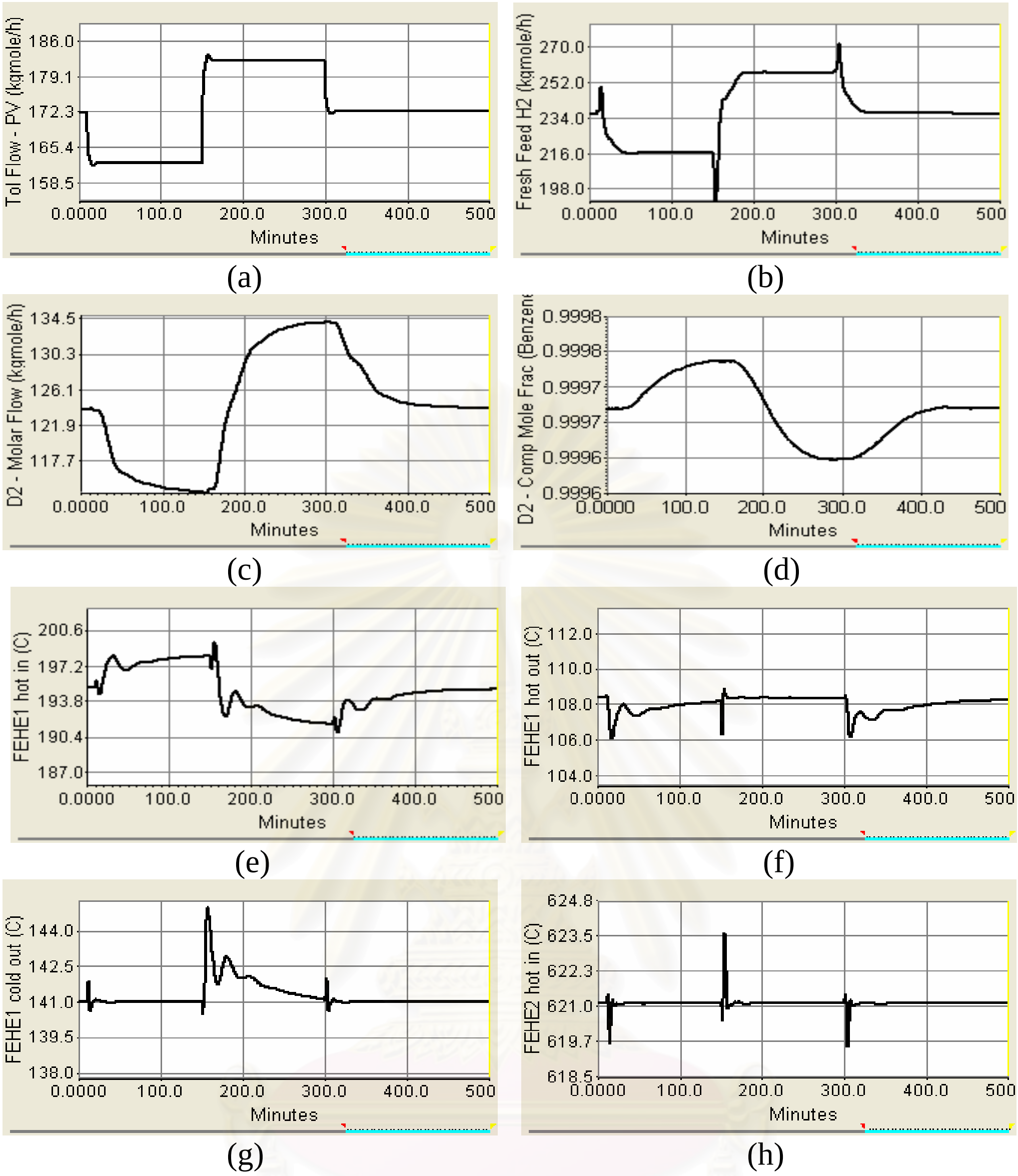
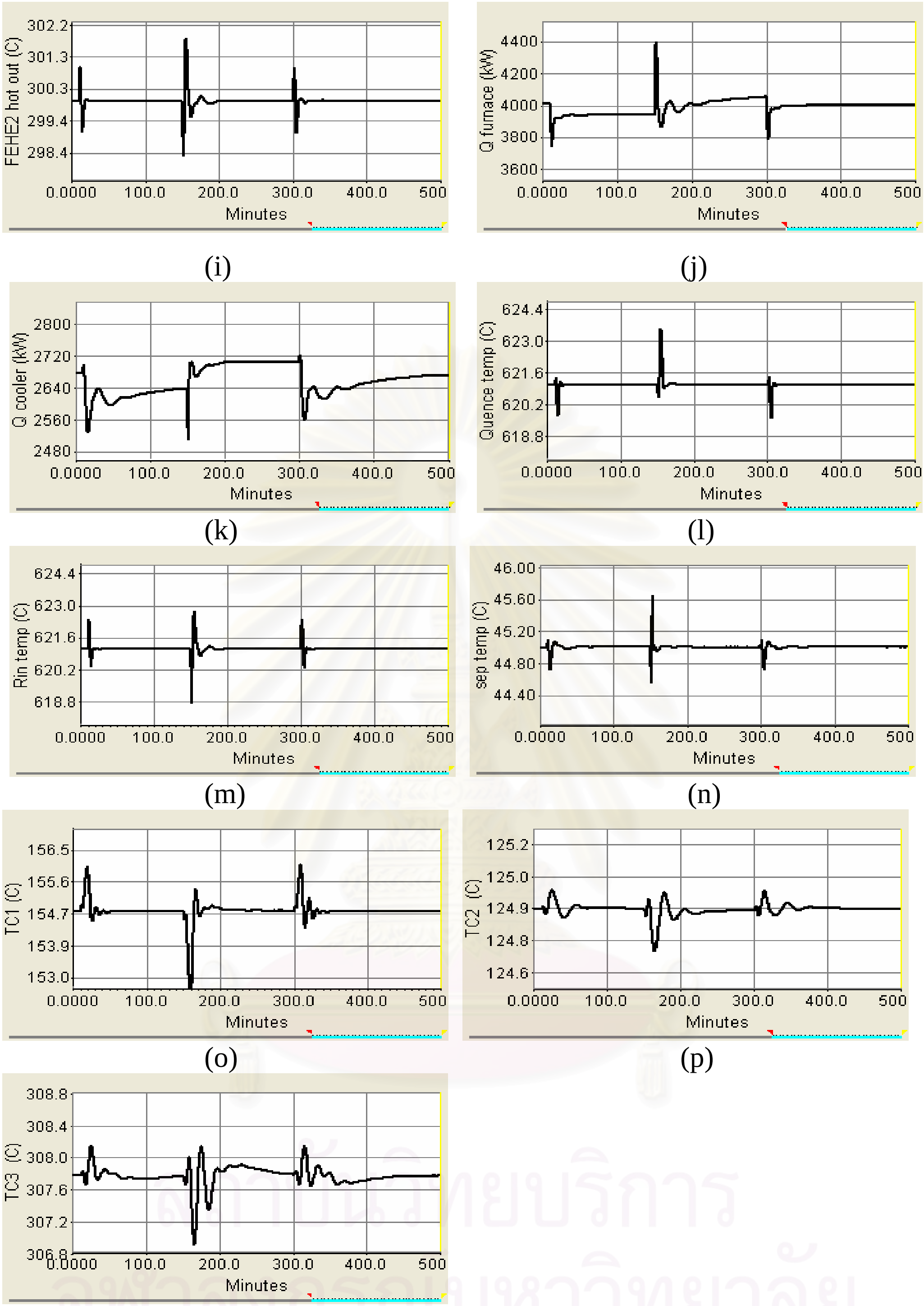


Figure 6.27 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates: CS2, where: (a) total toluene feed flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) furnace duty, (k) cooler duty, (l) quench temperature, (m) reactor inlet temperature, (n) separator temperature, (o) stabilizer column tray temperature, (p) product column tray temperature, (q) recycle column tray temperature



(i) (j) (k) (l) (m) (n) (o) (p) (t)

Figure 6.27 (continue)

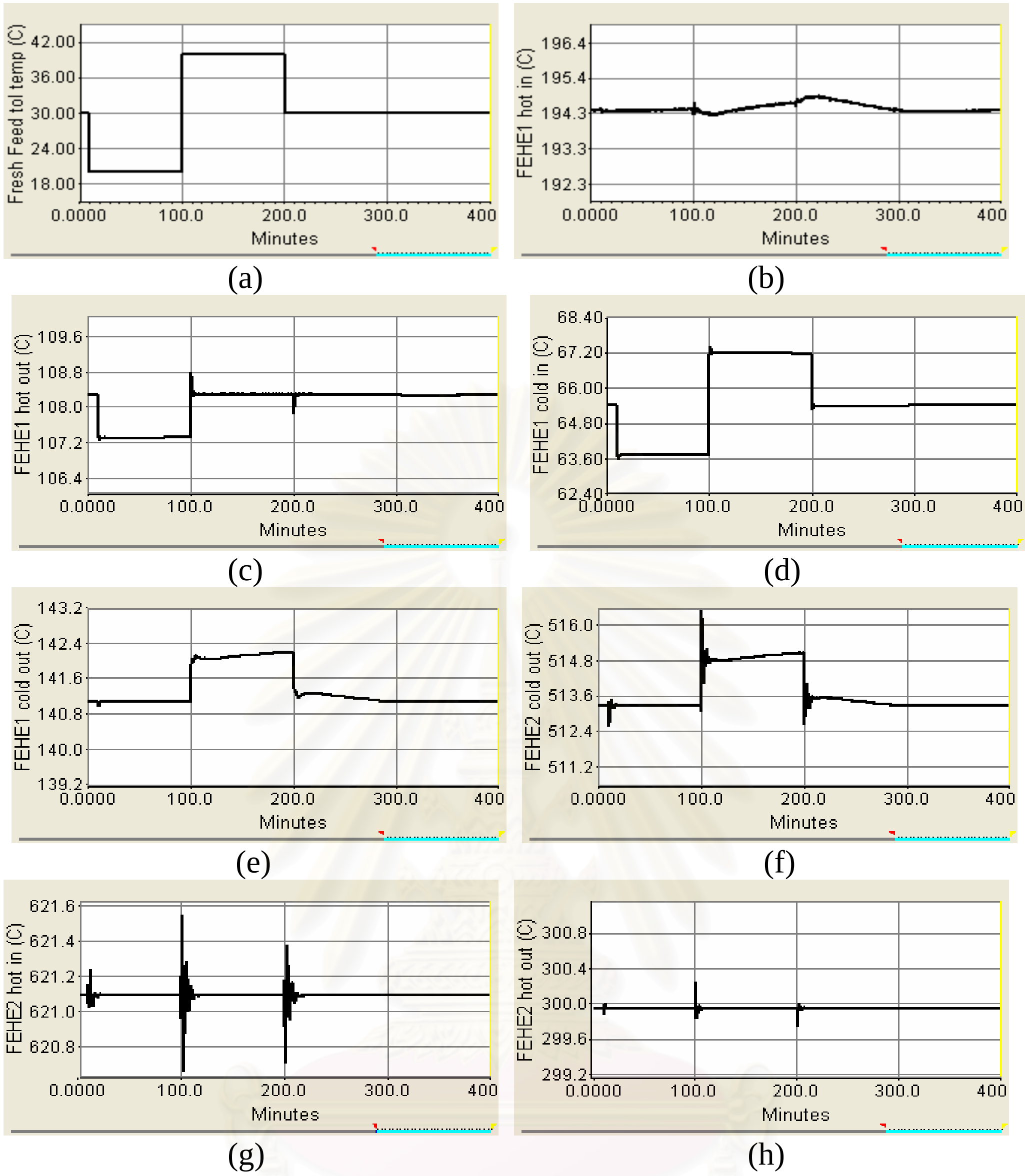
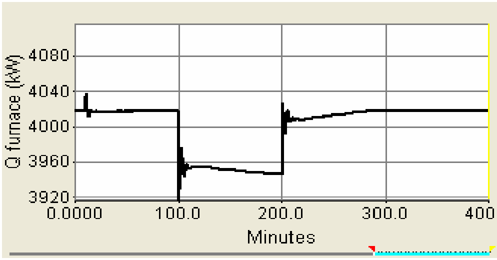


Figure 6.128 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature, (i) furnace duty, (j) cooler duty, (k) quench temperature, (l) reactor inlet temperature, (m) separator temperature, (n) stabilizer column tray temperature, (o) product column tray 12 temperature, (p) product column tray 17 temperature, (q) recycle column tray temperature



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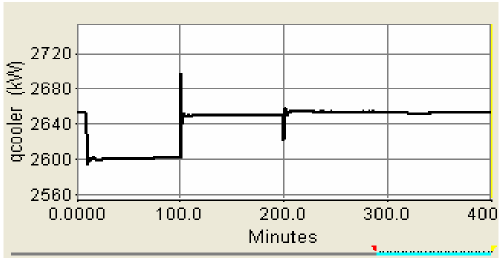

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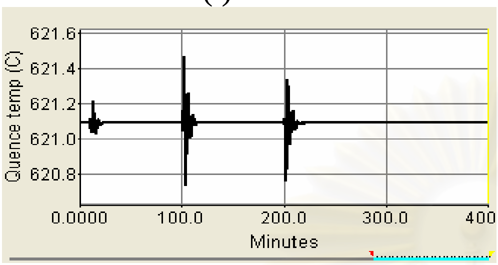

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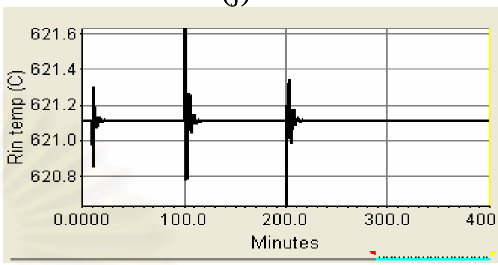

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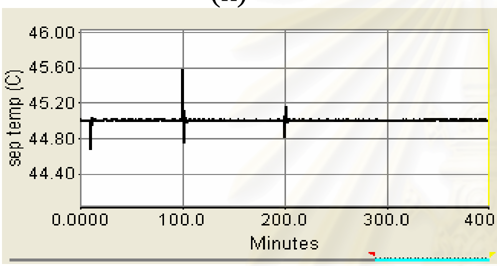

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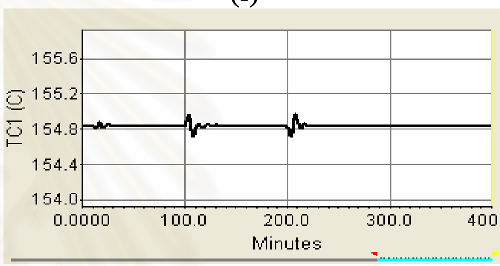

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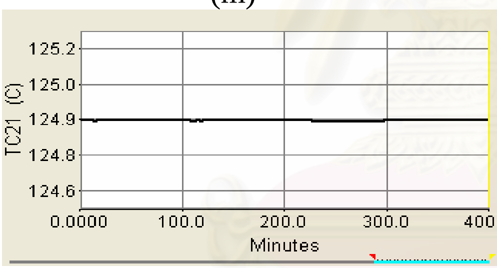

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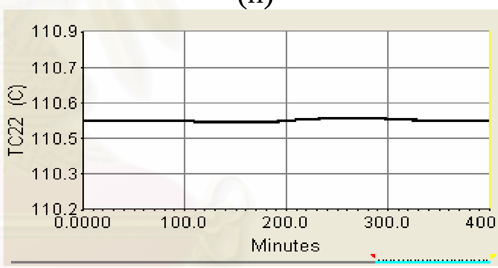

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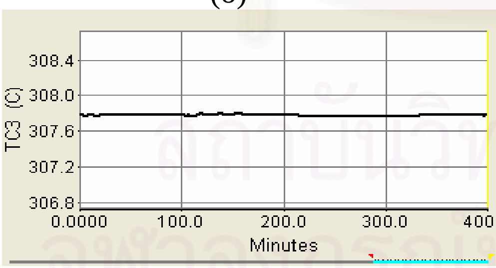


(q)

Figure 6.28 (continue)

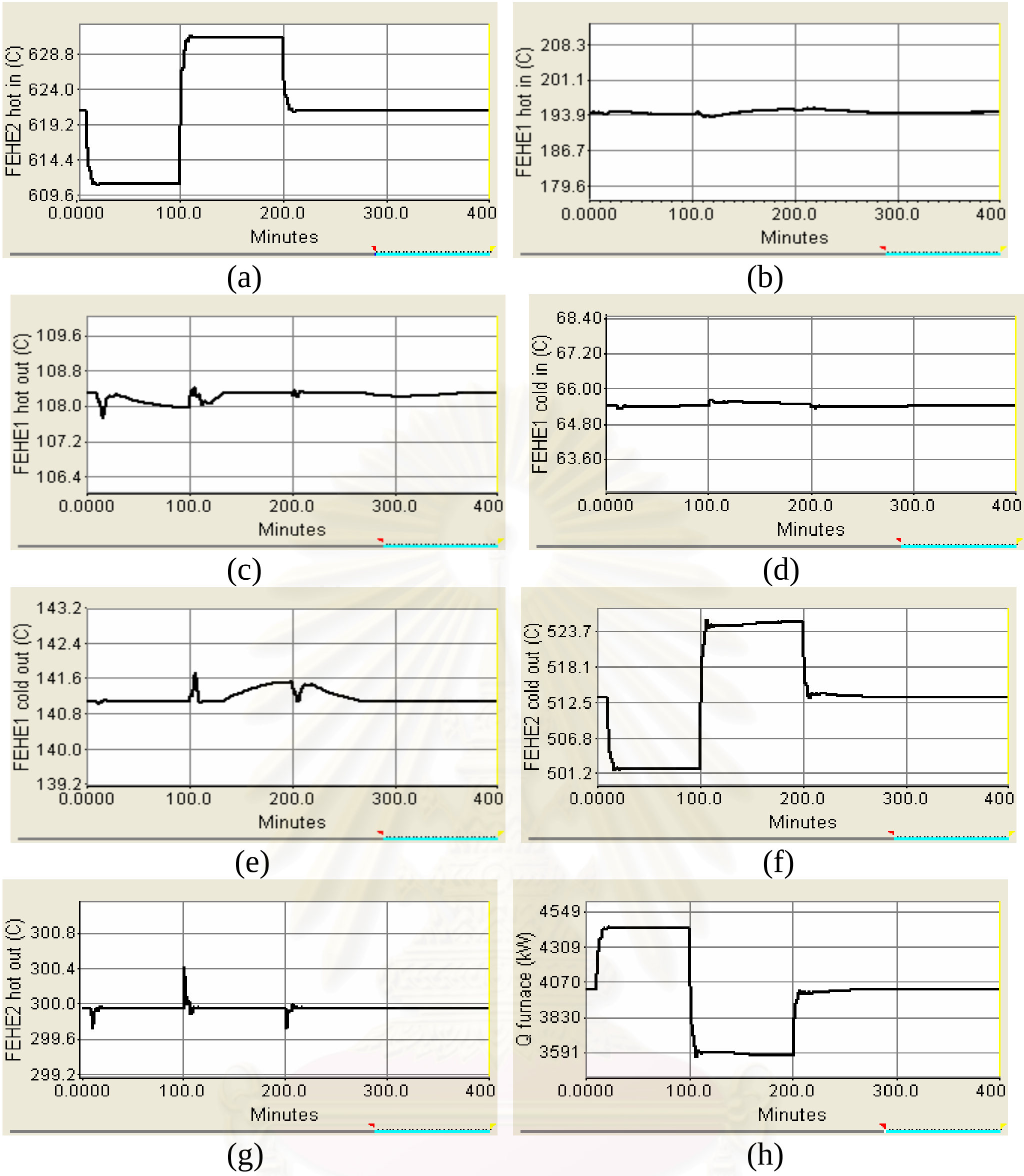


Figure 6.29 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot outlet temperature, (g) FEHE2 cold outlet temperature, (h) furnace duty, (i) cooler duty, (j) quench temperature, (k) reactor inlet temperature, (l) separator temperature, (m) stabilizer column tray temperature, (n) product column tray 12 temperature, (o) product column tray 17 temperature, (p) recycle column tray temperature

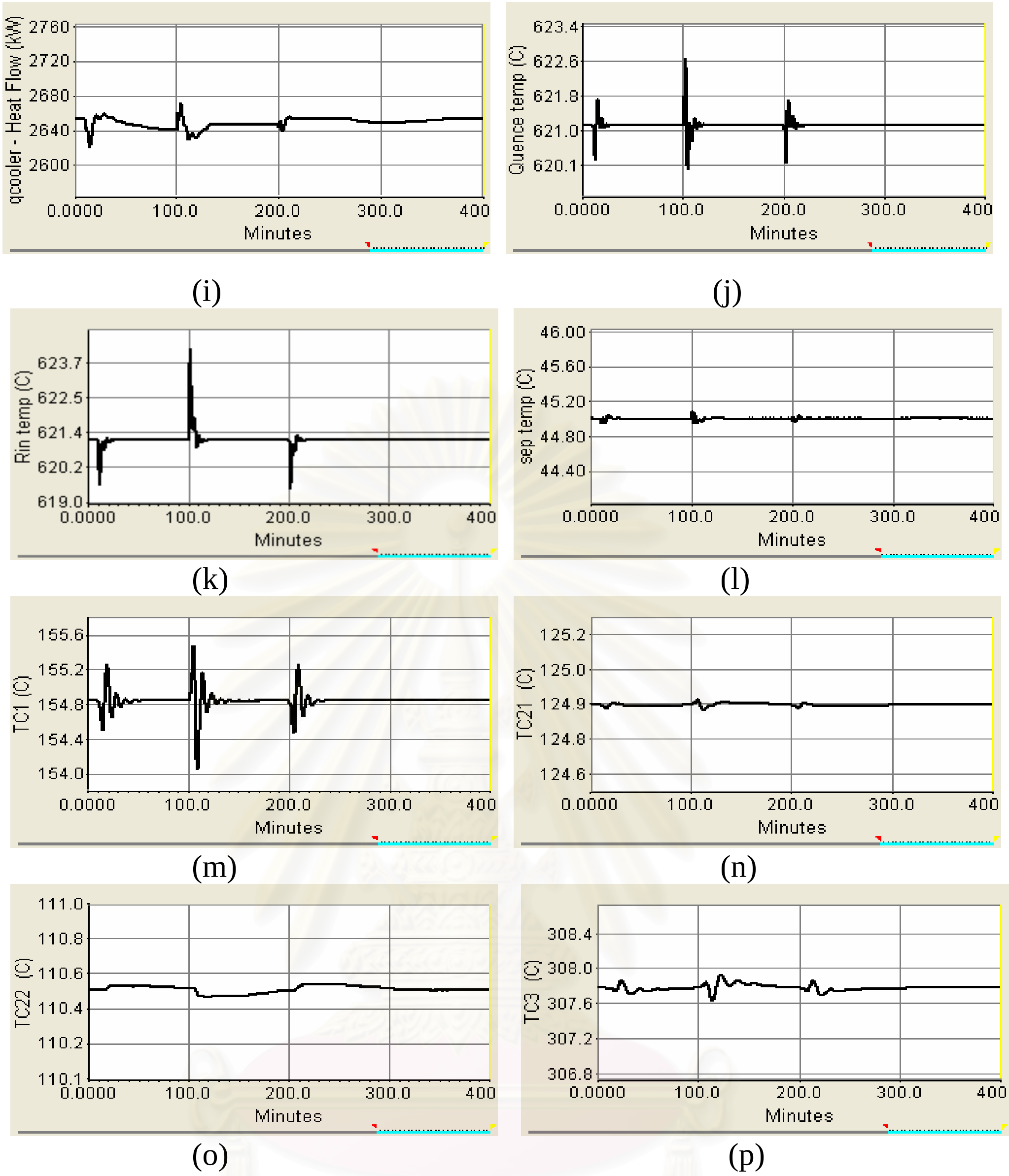


Figure 6.29 (continue)

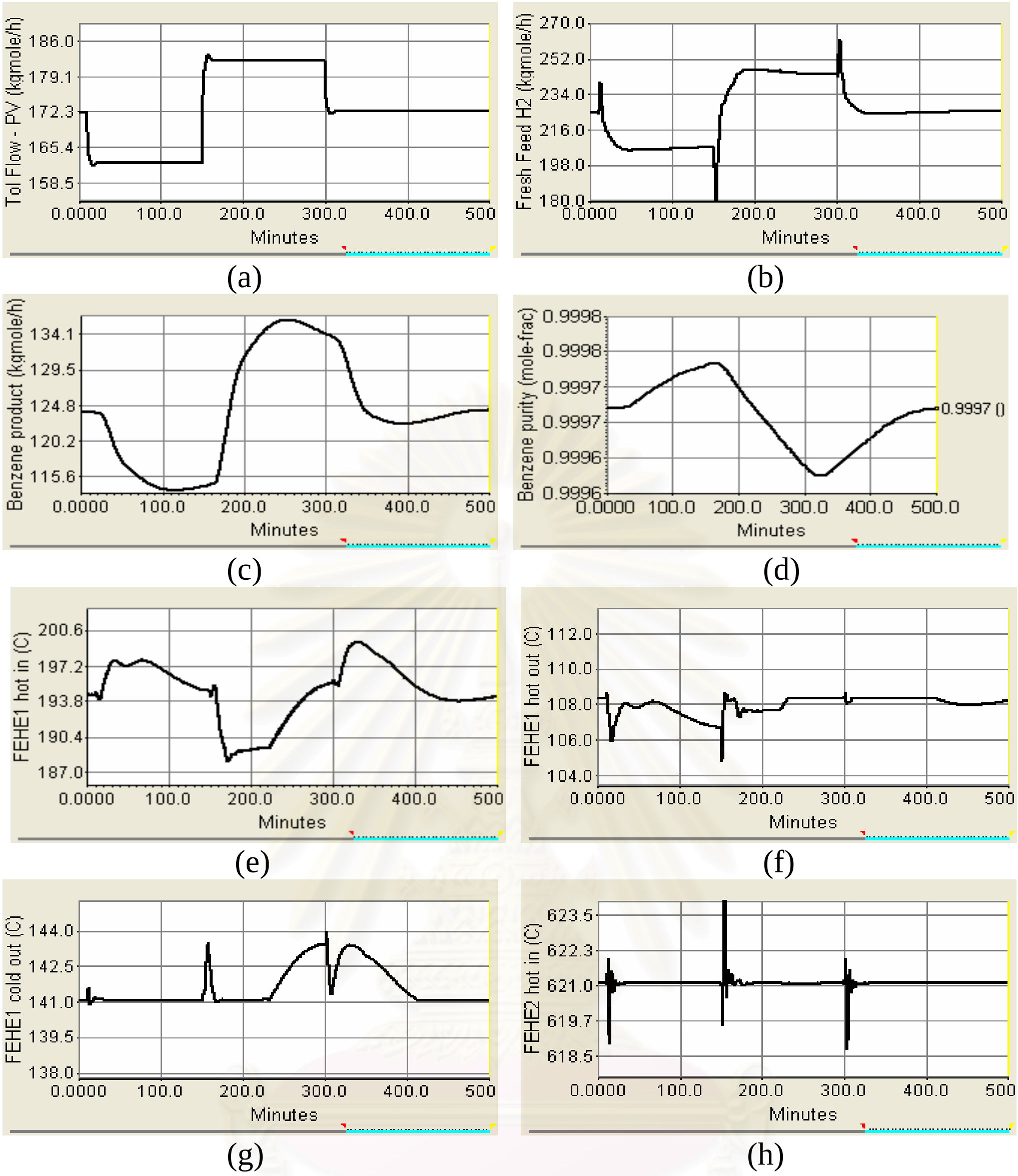
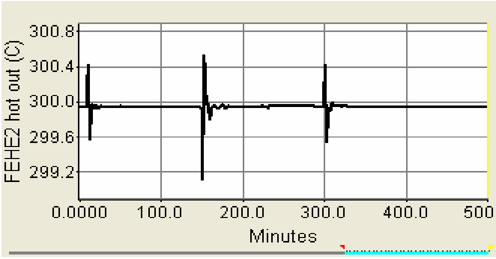


Figure 6.30 Dynamic Responses of the HDA Process Alternative 5 with Minimum Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates: CS3, where: (a) total toluene feed flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) furnace duty, (k) cooler duty, (l) quench temperature, (m) reactor inlet temperature, (n) separator temperature, (o) stabilizer column tray temperature, (p) product column tray 12 temperature, (q) product column tray 17 temperature, (r) recycle column tray temperature



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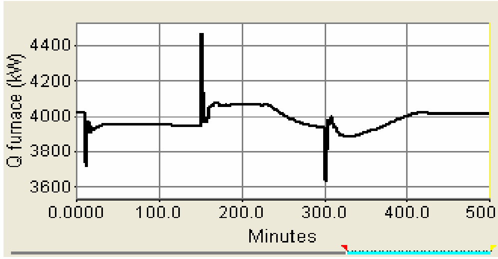

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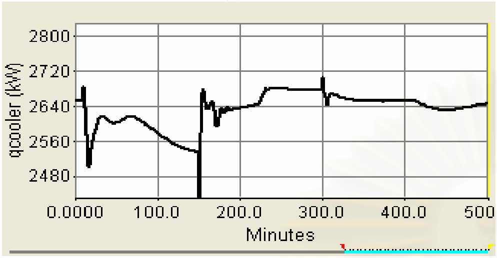

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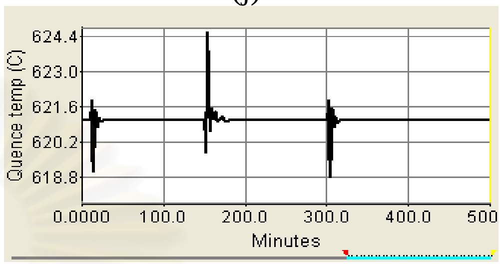

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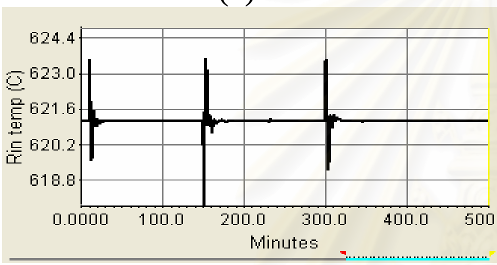

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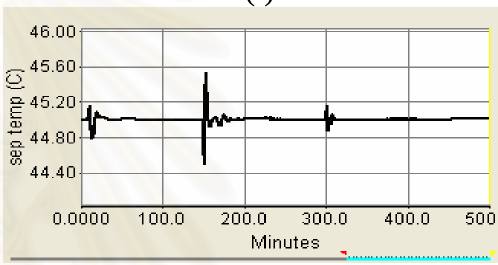

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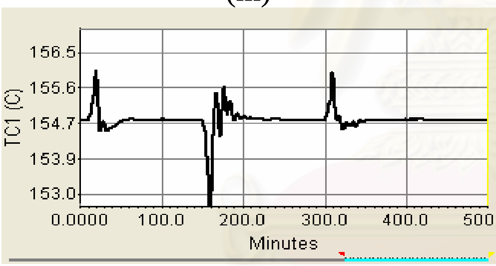

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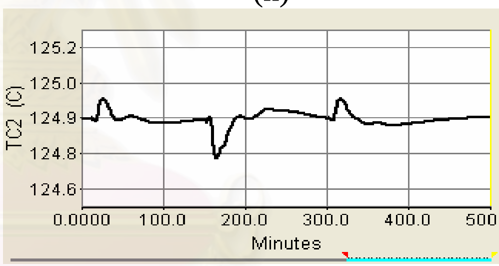

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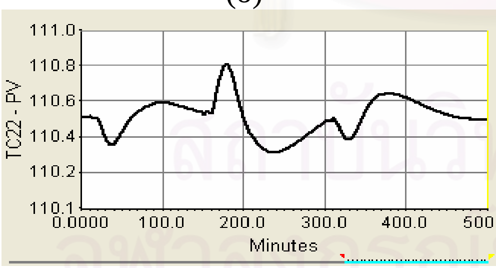

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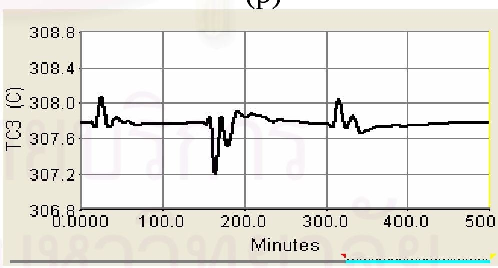


(r)

Figure 6.30 (continue)

6.5 Dynamic Simulation Results for HDA Process Alternative 6 with four Auxiliary Utility Units and Minimum Auxiliary Utility Units

In order to illustrate the dynamic behavior of the control structure in HDA process alternative 6 with four auxiliary utility units and minimum auxiliary utility units, several disturbance loads are made. The dynamic results are explained in this part.

6.5.1 Dynamic Simulation Results for HDA Process Alternative 6 with four Auxiliary Utility Units: CS1, CS2 and CS3

Three disturbance loads are used to evaluate the dynamic performance of the new control structure 1, 2 and 3 for HDA process alternative 6 with four auxiliary utility units.

6.5.1.1 Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream)

Figures 6.31, 6.34 and 6.37 show the dynamic responses of the control systems of HDA process alternative 6 with four auxiliary utility units to a change in the heat load disturbance of cold stream (reactor feed stream). In order to make this disturbance, first the fresh toluene feed temperature is decreased from 30 to 20^{o}C at time equals 10 minutes, and the temperature is increased from 20 to 40^{o}C at time equals 100 minutes, then its temperature is returned to its nominal value of 30^{o}C at time equals 200 minutes (Figures 6.31.a, 6.34.a and 6.37.a).

The three new control structures (CS1, CS2 and CS3) give the same result for shifting the thermal disturbance load in cold steam (reactor feed stream) to furnace or cooler as follows.

Whenever the cold inlet temperature of FEHE1 is decreased and then both the cold and hot outlet temperatures of FEHE1 decrease suddenly. The hot outlet temperature decreasing is a desired condition, hence the LSS switches the control action to control the cold outlet temperature of FEHE1 at its set point value of 144.5^{o}C (Figures 6.31.e, 6.34.e and 6.37.e). As a result, the hot outlet temperature of FEHE1 rapidly drops to a new steady state value (Figures 6.31.c, 6.34.c and 6.37.c), and the cooler duty decreases as DMER achieve (Figures 6.31.j, 6.34.j and 6.37.j). Whenever the cold inlet temperature of FEHE1 increases, both the cold and hot outlet temperatures of FEHE1 increase. In order to achieve maximum energy recovery, the increasing cold outlet temperature is a desired condition. So the LSS switches the control action to control the hot outlet temperature of FEHE1 at its set point value of 108.3 ^{o}C (Figures 6.31.c, 6.34.c and 6.37.c). As a result, the cold outlet temperature of FEHE1 temperature quickly increases a new steady state value (Figures 6.31.e, 6.34.e and 6.37.e), and the furnace duty decreases as DMER achieve (Figures 6.31.i, 6.34.i and 6.37.i).

The hot outlet temperature of FEHE2 and FEHE3 is slightly well controlled to prevent the thermal disturbance load propagation to the heat exchanger R1 (reboiler of stabilizer) and R3 (reboiler of recycle column).

As can be seen, this disturbance load has a little bit effect to the tray temperatures in the stabilizer and product columns except the average tray temperature in recycle column however the three new control structures can control the tray temperature in recycle column slightly well (Figures 6.31.n, 6.34.n and 6.37.n). The dynamic response of CS2 is more oscillation than CS1 and CS3.

The reactor inlet temperature, the quench temperature, the separator temperature are slightly well controlled but the dynamic response of CS2 is smoother than CS1 and CS3.

6.4.1.2 Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream)

Figures 6.32, 6.35 and 6.38 shows the dynamic responses of the control systems of HDA process alternative 6 with three auxiliary utility units to a change in the heat load disturbance of hot stream (the hot reactor product). This disturbance is made as follows: first the set point of FEHE3-hot-inlet temperature controller (i.e. TCX1 in Figures 6.7 6.8 and 6.9) is decreased from 621.1 to 611.1oC at time equals 10 minutes, and the temperature is increased from 611.1 to 631.1oC at time equals 100 minutes, then its temperature is returned to its nominal value of 621.1oC at time equals 200 minutes (Figures 6.32.a, 6.35.a and 6.38.a). As can be seen, this temperature response is very fast (Figures 6.32.a, 6.35.a and 6.38.a), the new steady state is reached quickly (Figures 6.32.a, 6.35.a and 6.38.a).

Since the hot outlet temperature of FEHE2 (i.e. the hot temperature at the entrance of reboiler at stabilizer column, R1) is controlled (Figures 6.32.g, 6.35.g and 6.38.g) at its set point value of 300 ^{o}C to prevent the propagation of the thermal disturbance, both the positive and negative disturbance loads of the hot stream are shifted to the furnace utility. Therefore, whenever the negative disturbance load comes with the hot stream (i.e. the hot inlet temperature of FEHE2 decreases Figures 6.32.a, 6.35.a and 6.38.a), this disturbance load is shifted to the furnace utility. The furnace duty will be increased in this case (Figures 6.32.h, 6.35.h and 6.38.h). Consider the case when the hot inlet temperature of FEHE2 increases (Figures 6.32.a, 6.35.a and 6.38.a), this is a desired condition to shift the disturbance load to the cold stream. Therefore, the furnace duty decreases to a new steady state value (Figures 6.32.h, 6.35.h and 6.38.h).

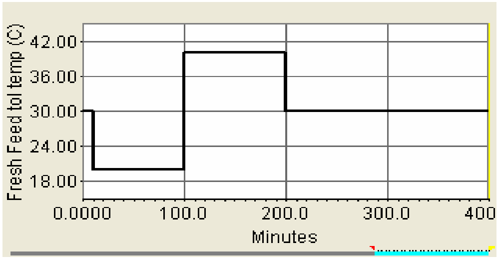
Again, the reactor inlet temperature, the quench temperature, the separator temperature and tray temperature in three columns are slightly well controlled.

6.4.1.3 Change in the Total Toluene Feed Flowrates

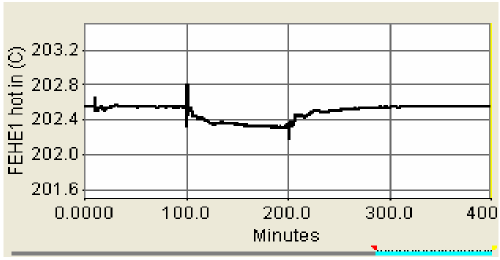
Figures 6.33, 6.36 and 6.39 shows the dynamic responses of the control systems of HDA process alternative 5 with three auxiliary utility units to a change in the total toluene flowrates. This disturbance is made by decreasing toluene flowrates from 172.3 to 162.3 kgmole/h at time equals 10 minutes, and the flowrates is increased from 162.3 to 182.3 kgmole/h at time equals 100 minutes, then its flowrates is returned to its nominal value of 172.3 kgmole/h at time equals 200 (Figures 6.33.a , 6.36.a and 6.39.a).

The dynamic result can be seen that the drop in total toluene feed flowrates reduces the reaction rate, so the benzene product flowrates drops (Figures 6.33.c, 6.36.c and 6.39.c).and the benzene product quality increases (Figures 6.33.d, 6.36.d and 6.39.d).Consider the case when the total toluene feed flowrates increase (Figures 6.33.a, 6.36.a and 6.39.a), and the benzene product flowrates increase because of the reaction rate enlargement. The benzene product quality will increase in this case (Figures 6.33.d, 6.36.d and 6.39.d). The deviation of benzene product purity from nominal value in CS1, CS2 and CS3 are same when total toluene feed flowrates change.

As can be seen, this disturbance has height effect to trey temperatures in the stabilizer, product and recycle columns however all control structure can control the tray temperatures in the three columns slightly well.



(a)



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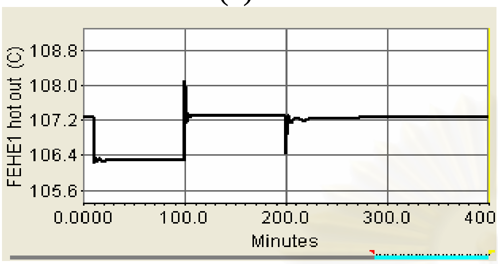

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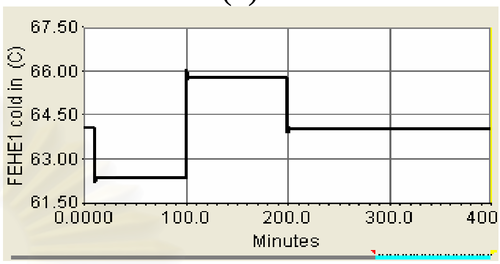

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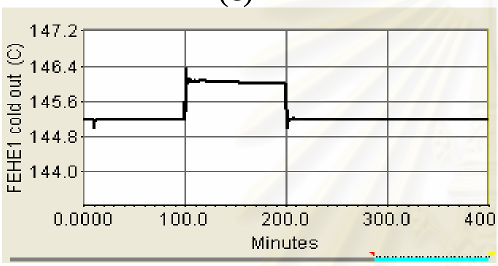

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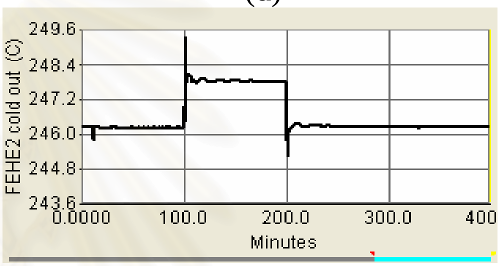

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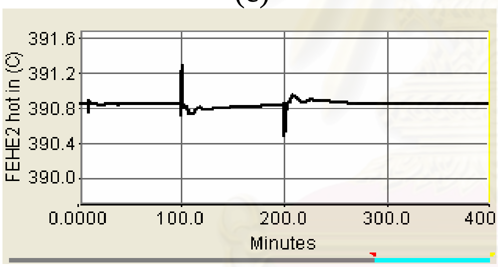
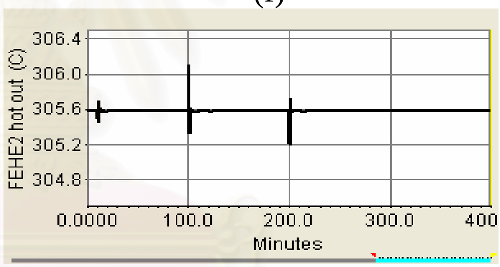
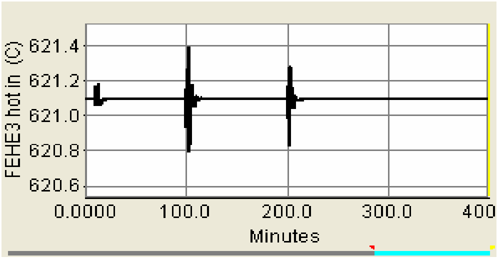


(g)



(h)

Figure 6.31 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE3 hot inlet temperature, (j) FEHE3 hot outlet temperature(k) FEHE3 cold outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray temperature, (s) recycle column tray temperature



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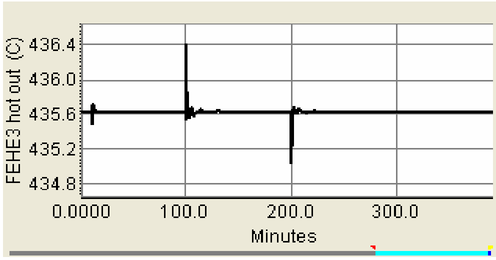

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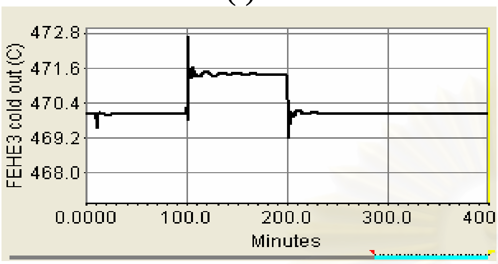

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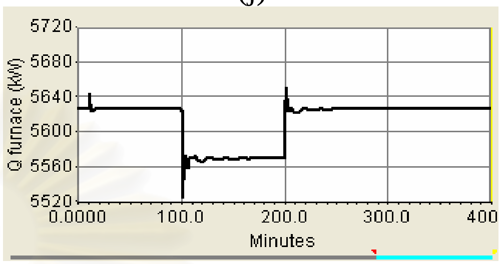

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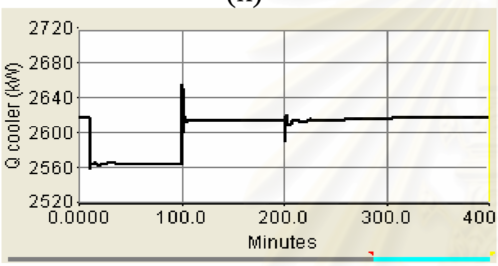

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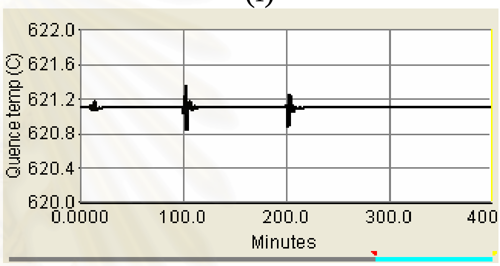

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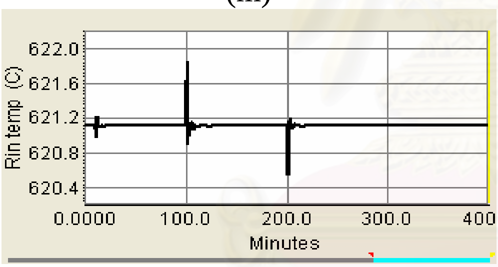

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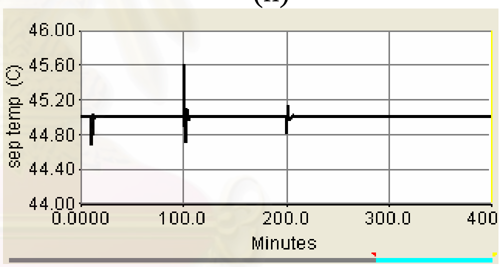

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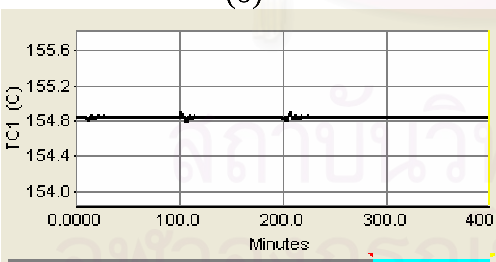

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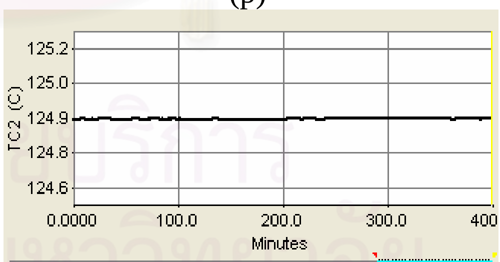

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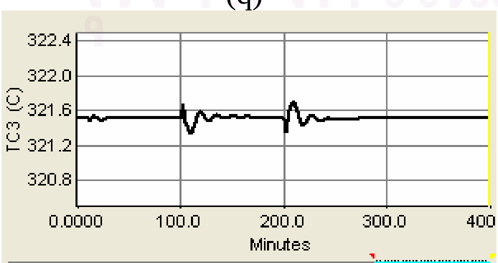

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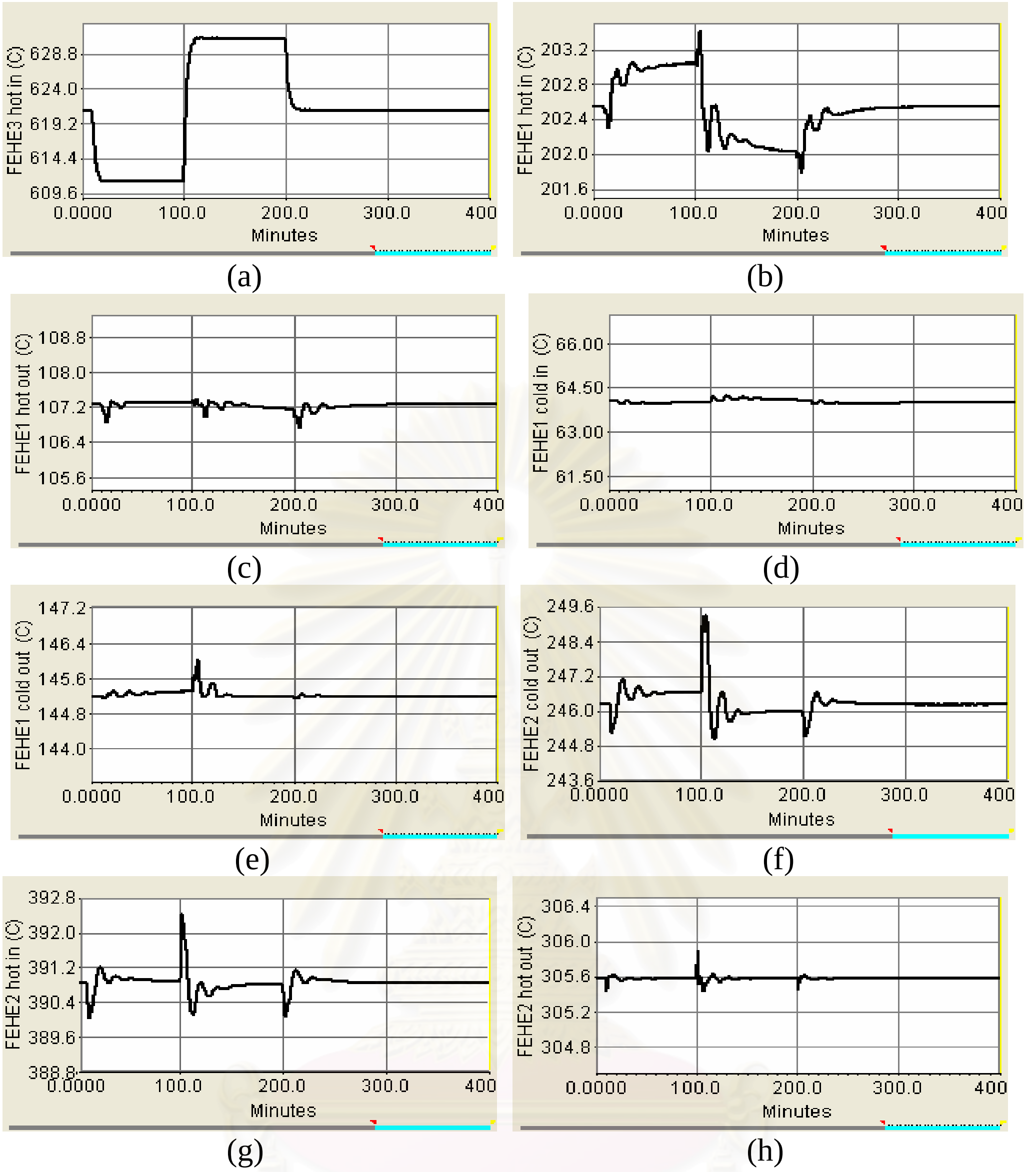
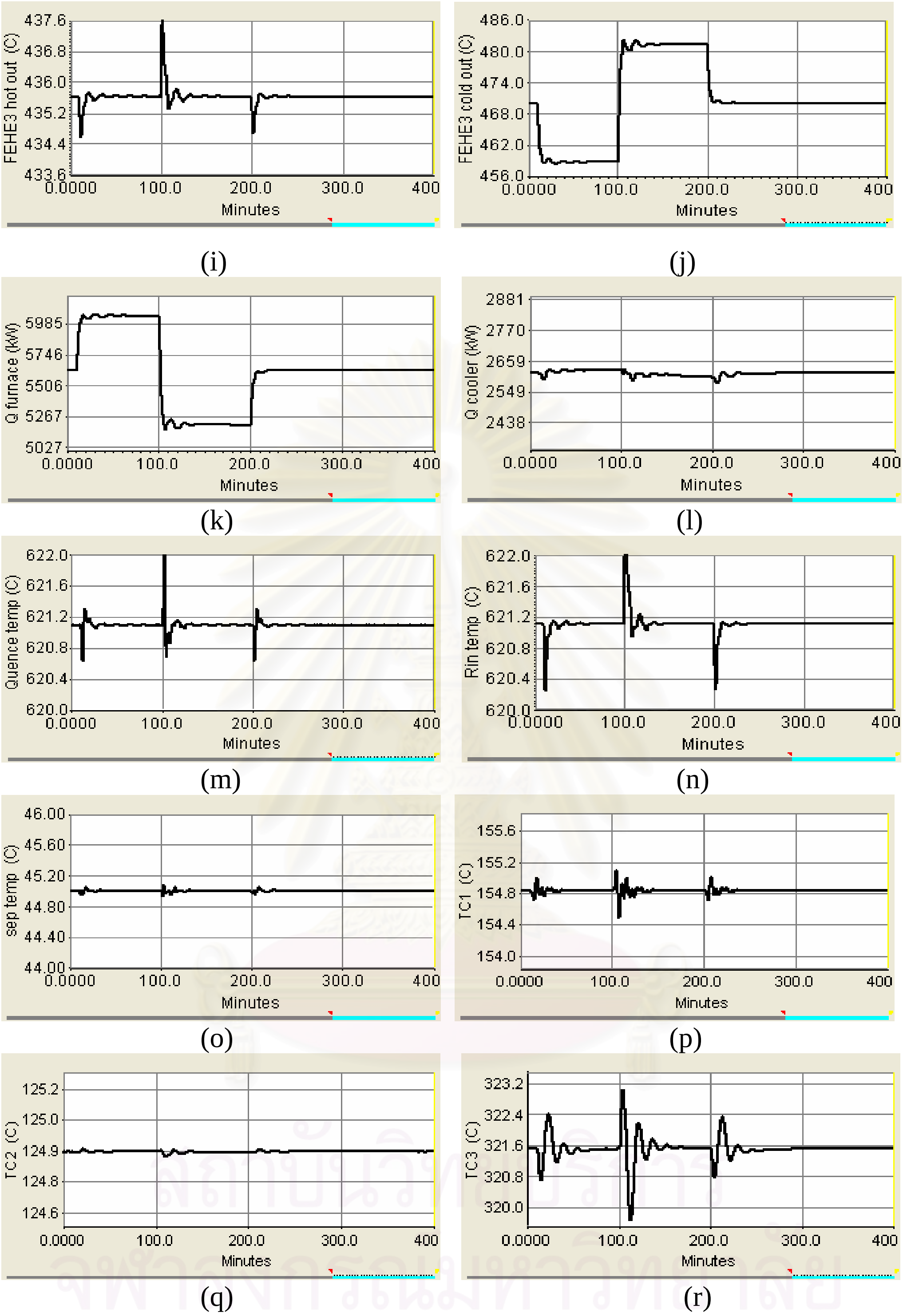


Figure 6.32 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature,(i) FEHE3 hot outlet temperature, (j) FEHE3 cold outlet temperature, (k) furnace duty, (l) cooler duty, (m) quench temperature, (n) reactor inlet temperature, (o) separator temperature, (p) stabilizer column tray temperature, (q) product column tray temperature, (r) recycle column tray temperature



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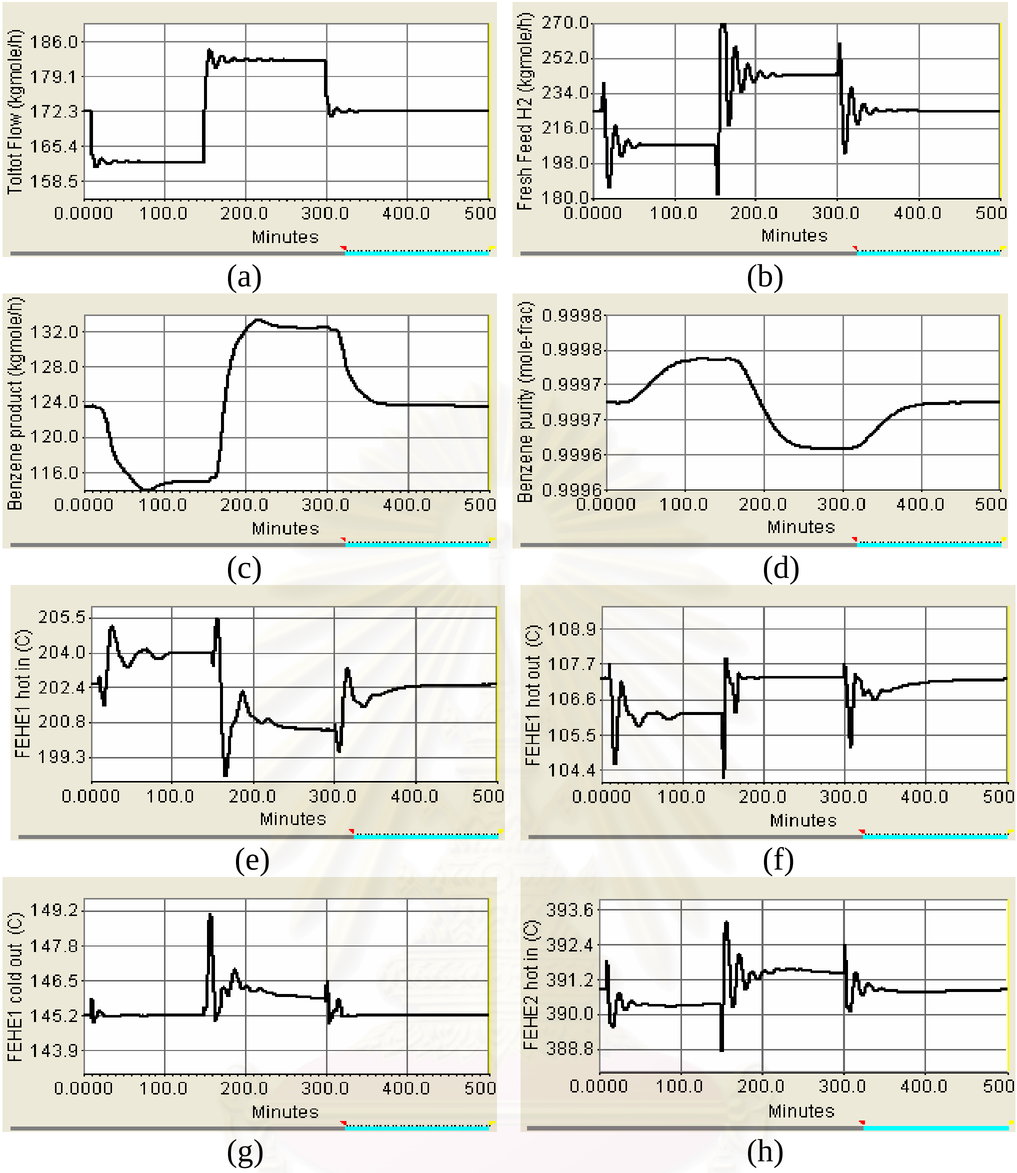
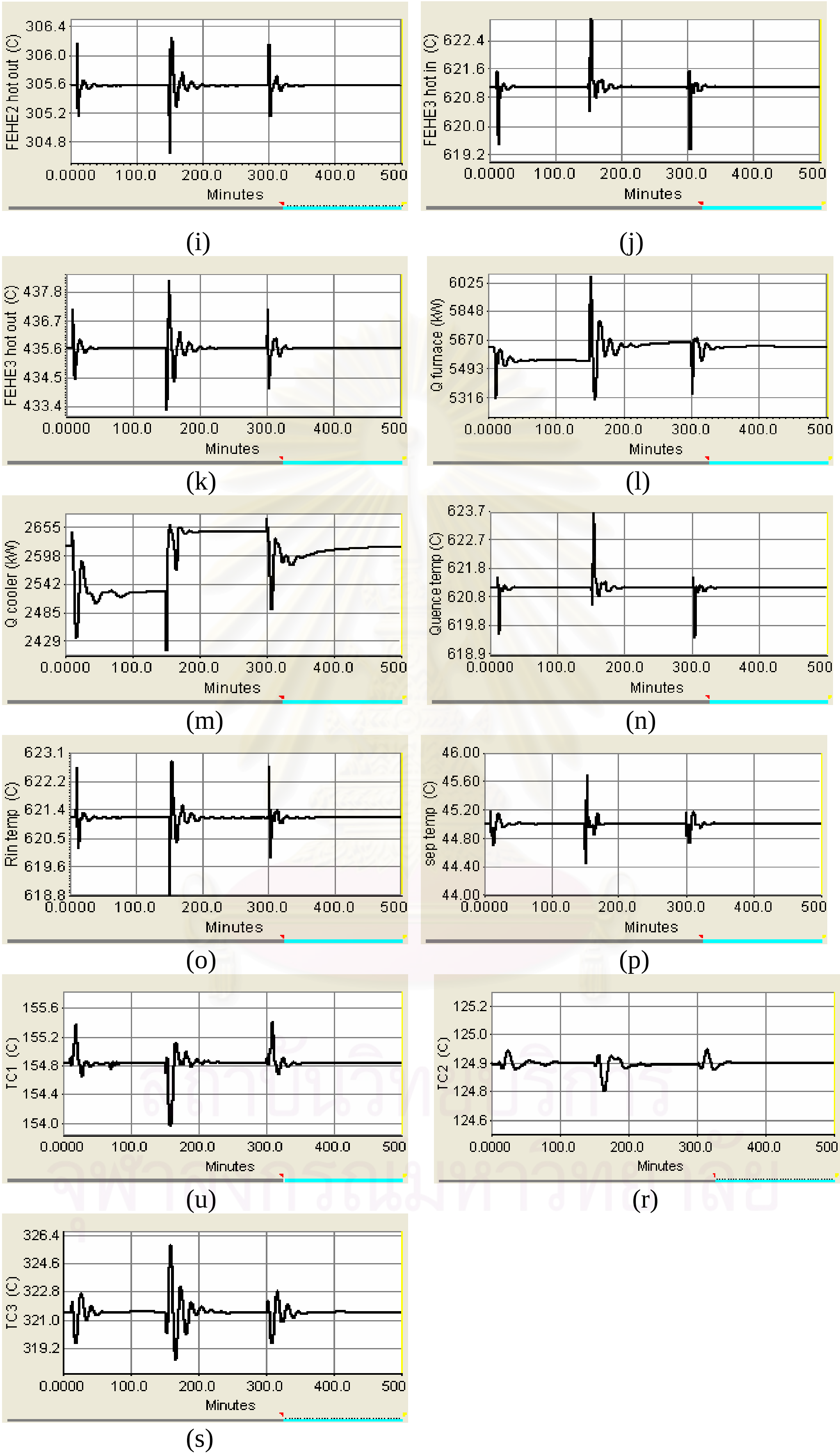


Figure 6.33 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates (Reactor Feed Stream):CS1, where: (a) recycle toluene flowrates, (b) fresh feed hydrogen flowratres, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) FEHE3 hot inlet temperature, (k) FEHE3 hot outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray temperature, (s) recycle column tray temperature



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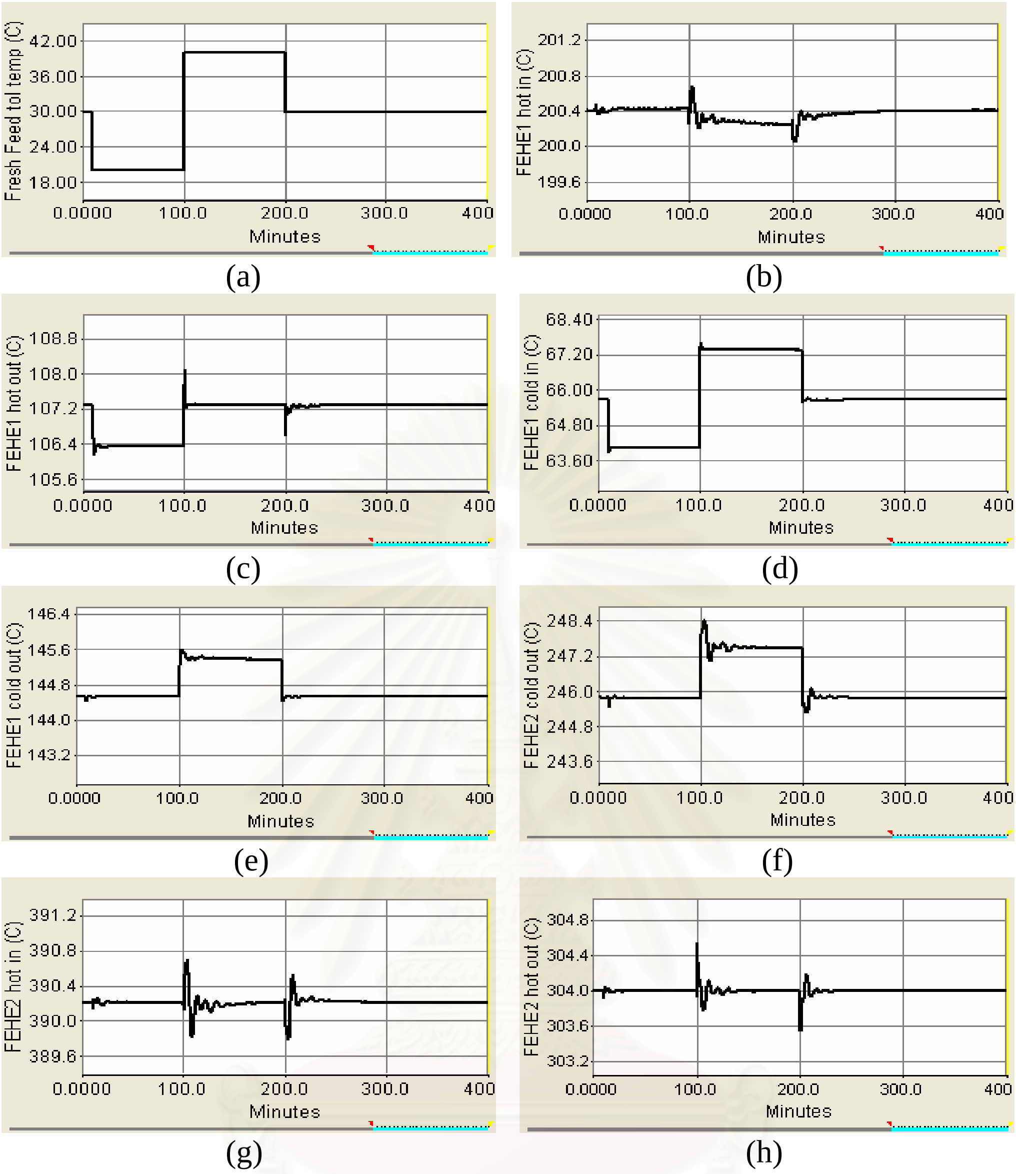
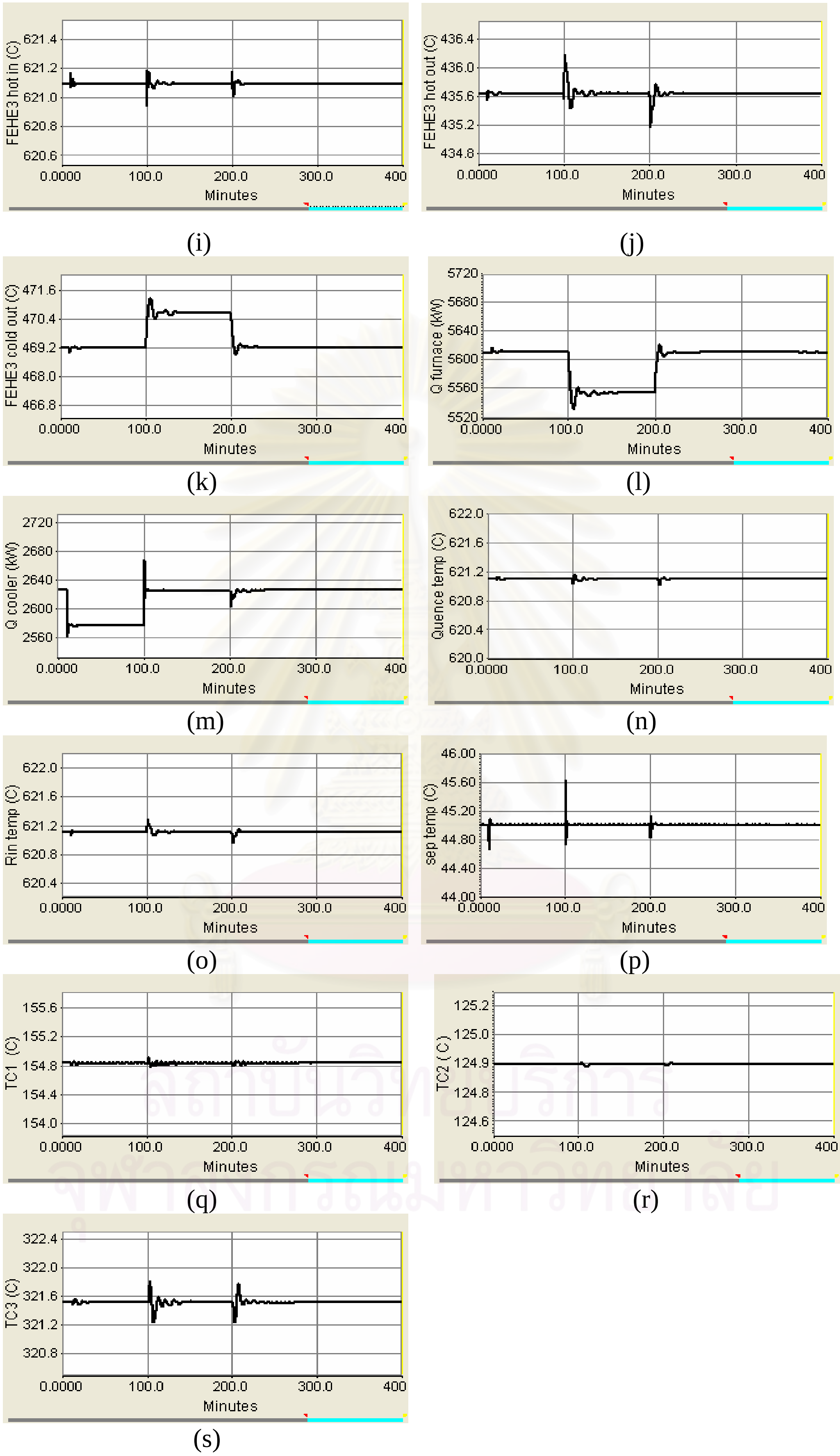


Figure 6.34 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE3 hot inlet temperature, (j) FEHE3 hot outlet temperature(k) FEHE3 cold outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray temperature, (s) recycle column tray temperature



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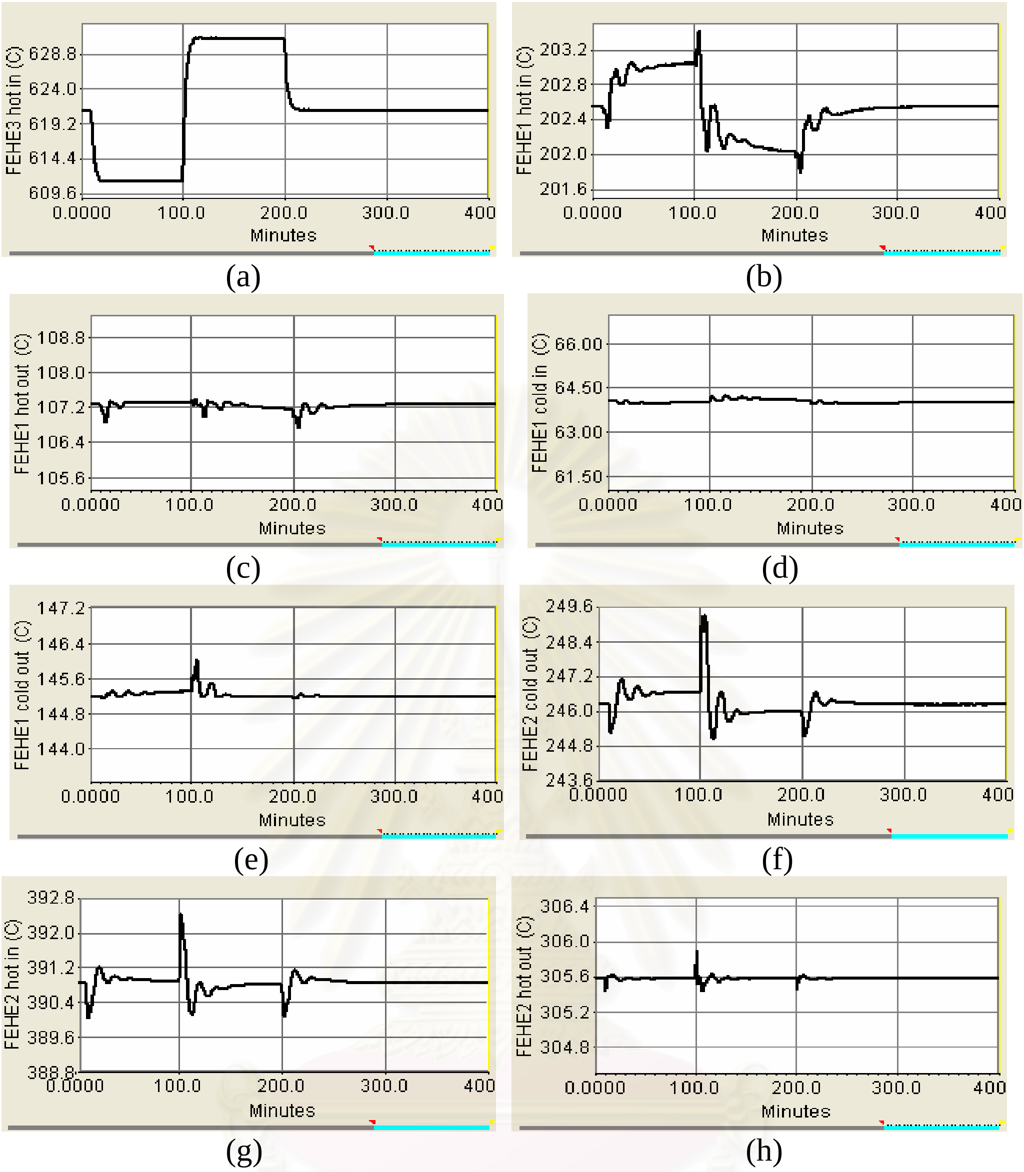
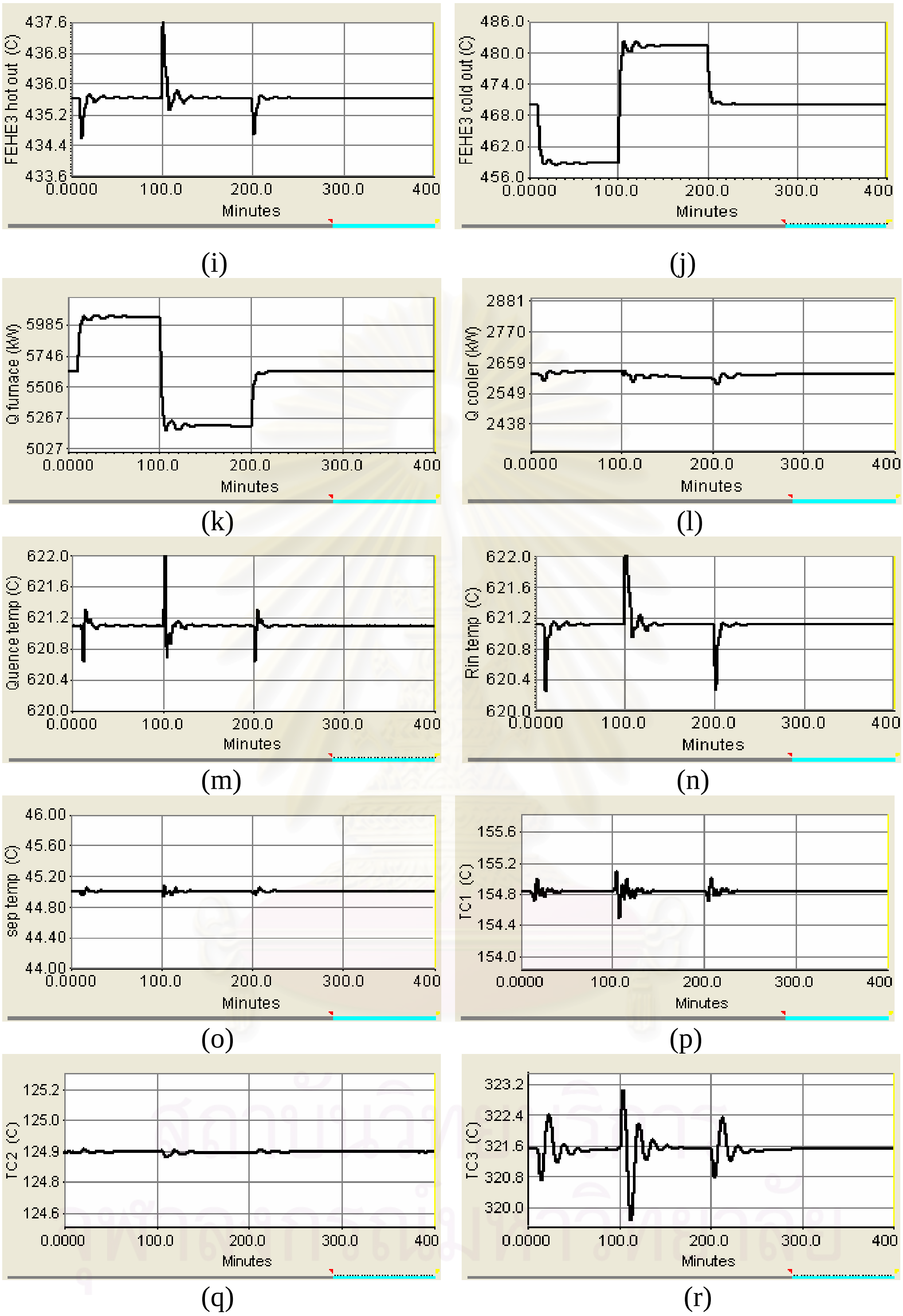


Figure 6.35 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature,(i) FEHE3 hot outlet temperature, (j) FEHE3 cold outlet temperature, (k) furnace duty, (l) cooler duty, (m) quench temperature, (n) reactor inlet temperature, (o) separator temperature, (p) stabilizer column tray temperature, (q) product column tray temperature, (r) recycle column tray temperature



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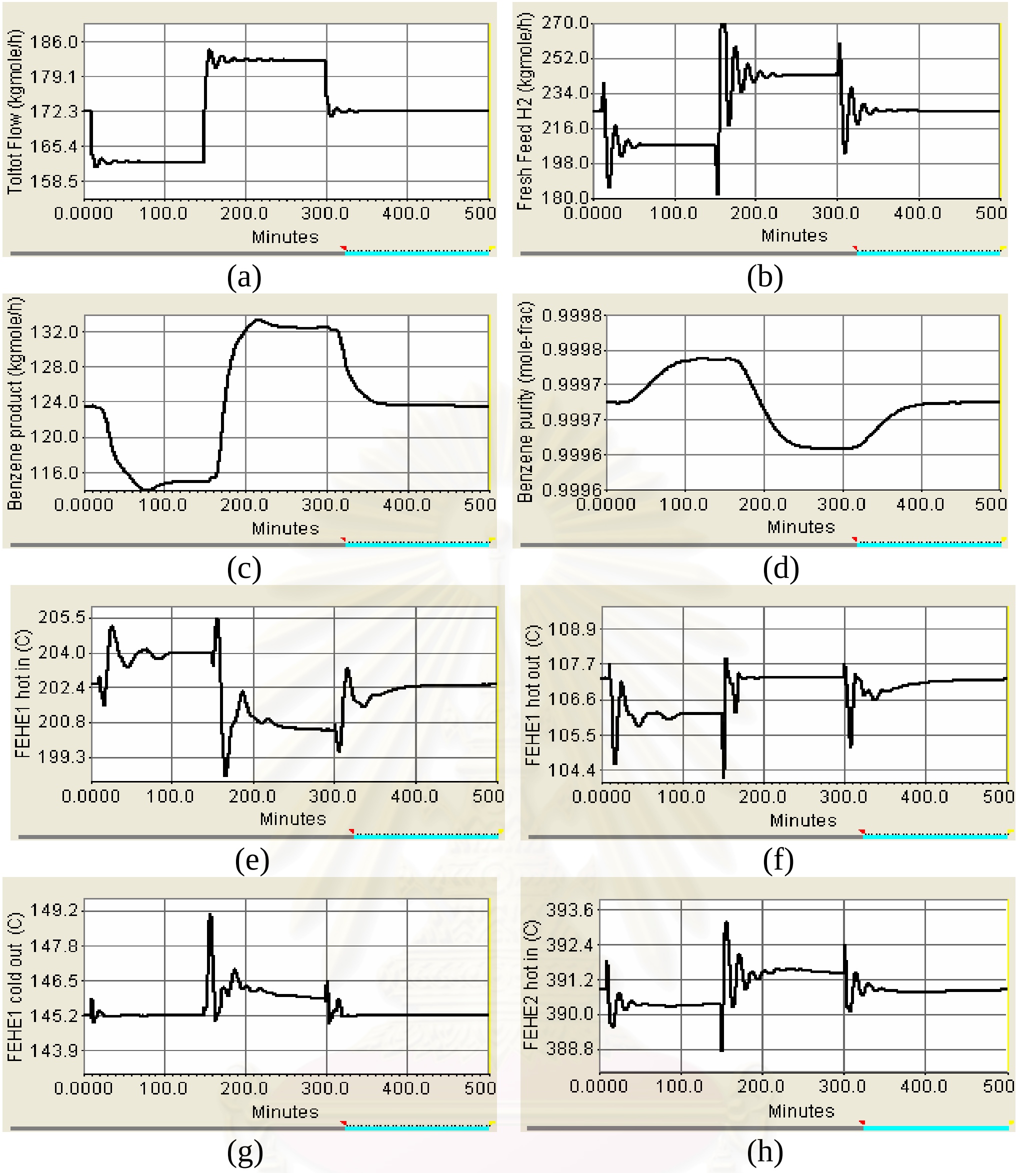
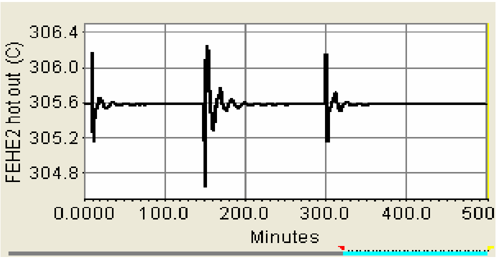


Figure 6.36 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates (Reactor Feed Stream):CS1, where: (a) recycle toluene flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) FEHE3 hot inlet temperature, (k) FEHE3 hot outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray temperature, (s) recycle column tray temperature



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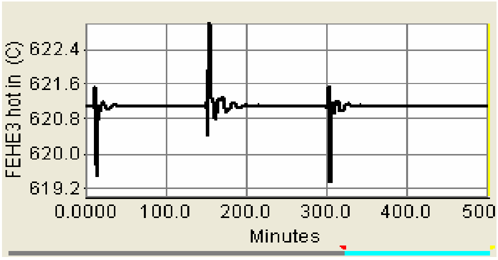

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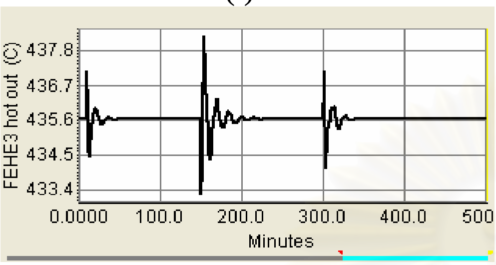

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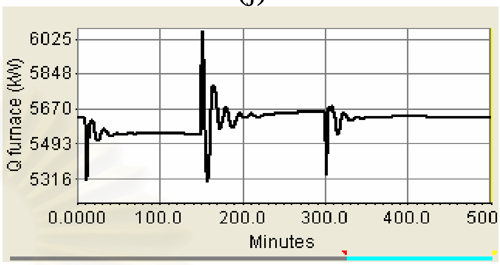

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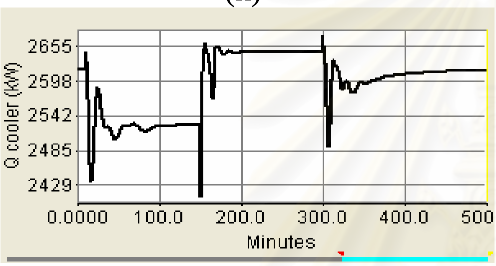

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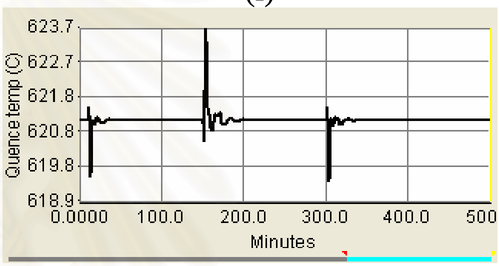

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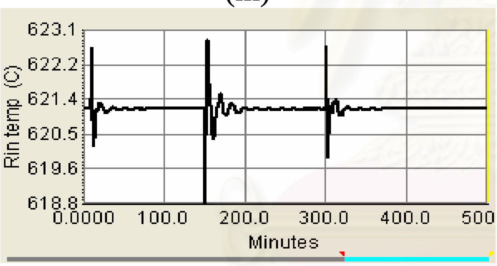

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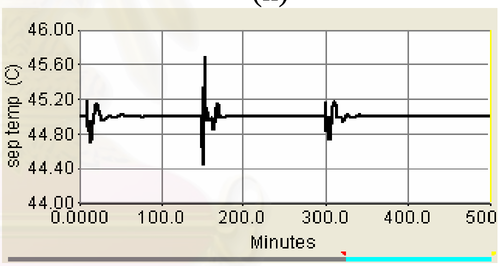

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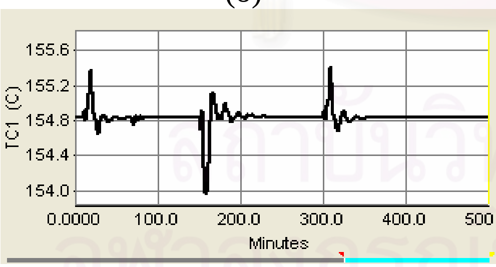

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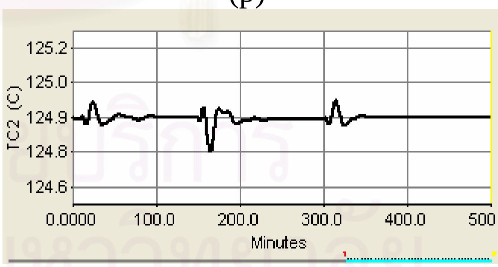
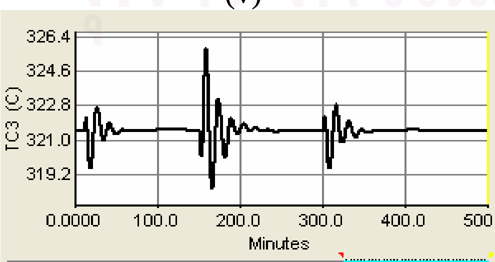


(r)



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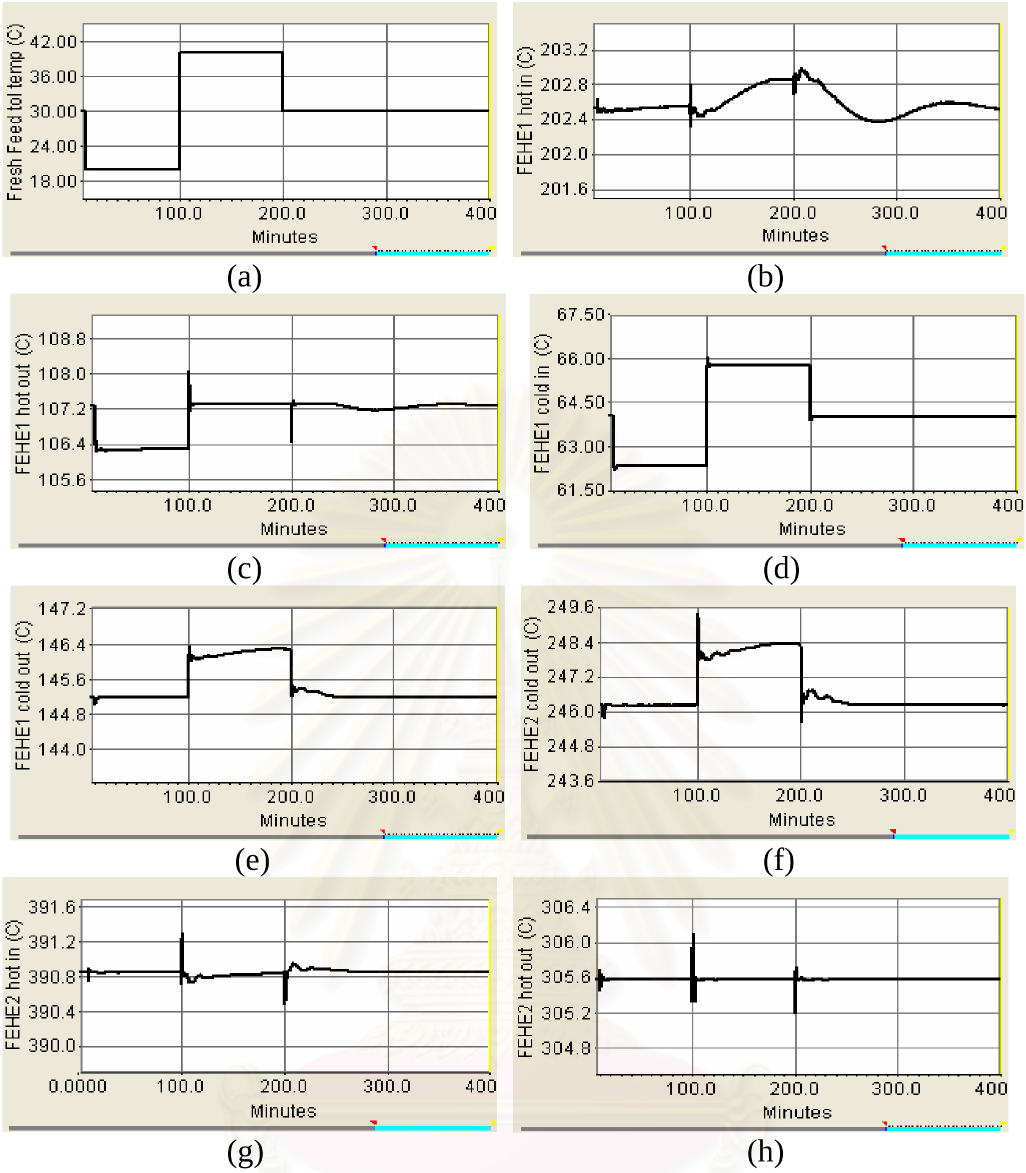
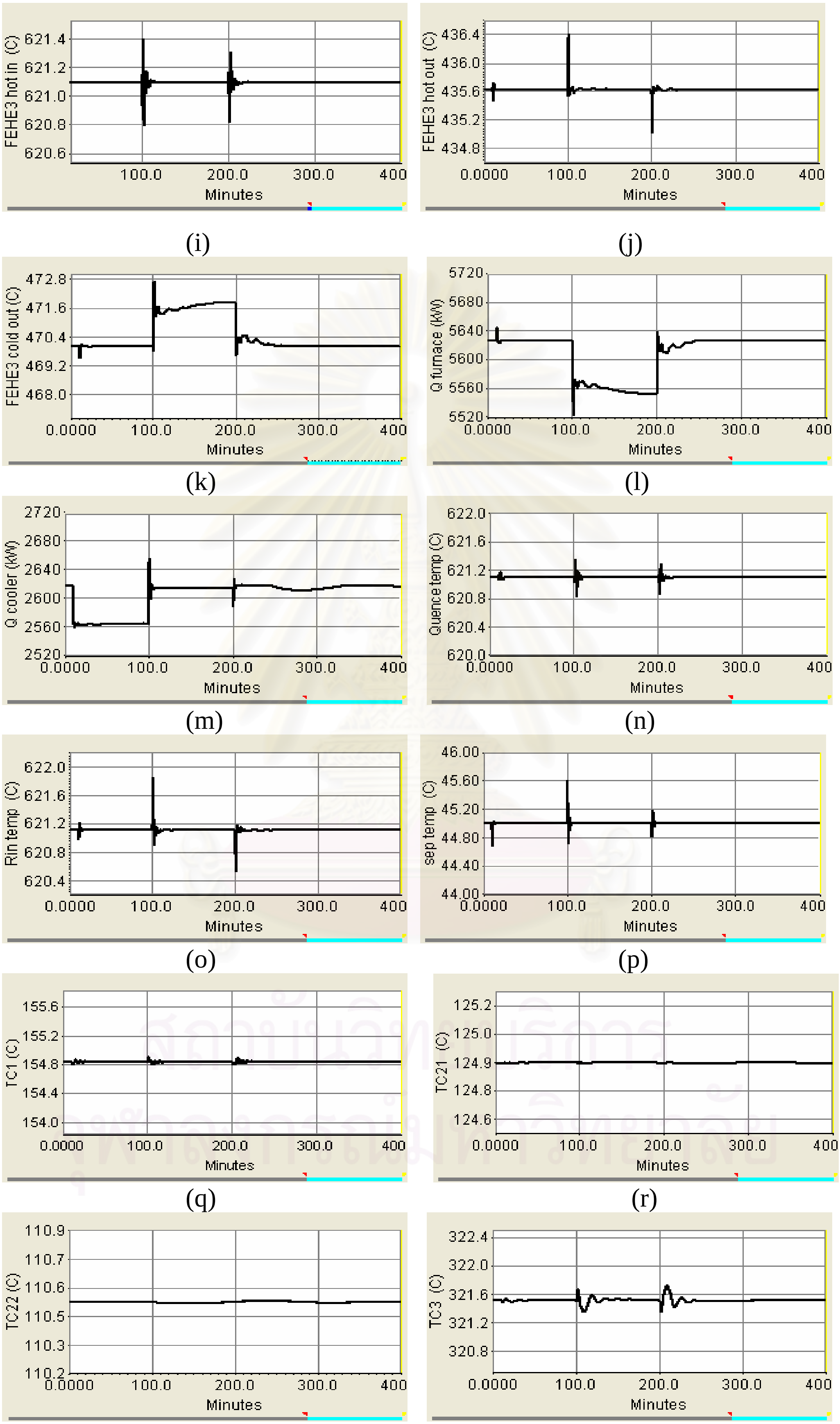


Figure 6.37 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE3 hot inlet temperature, (j) FEHE3 hot outlet temperature(k) FEHE3 cold outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray 12 temperature, (s) product column tray 17 temperature, (t) recycle column tray temperature



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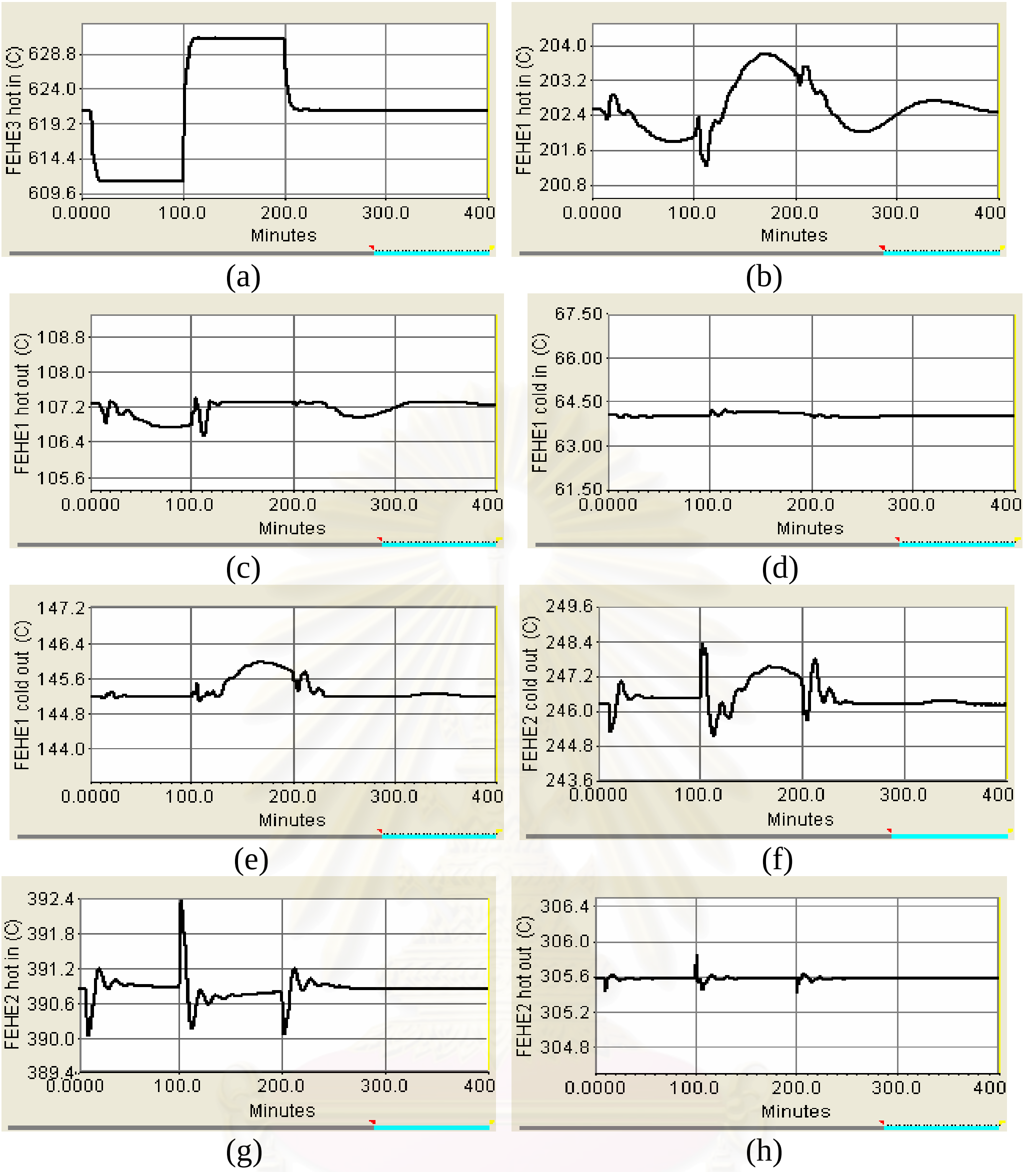
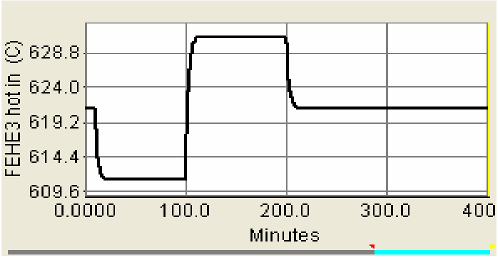
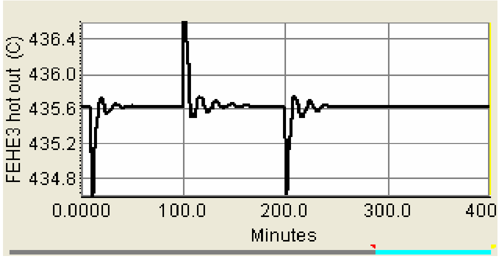
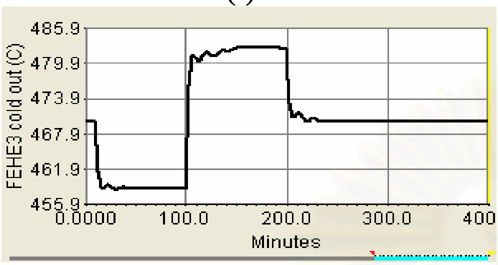


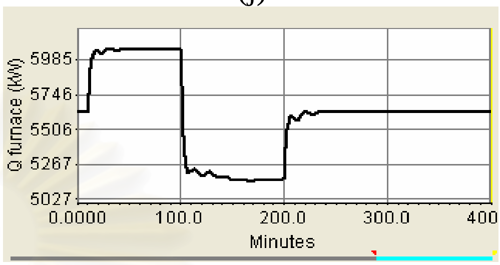
Figure 6.38 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature,(i) FEHE3 hot outlet temperature, (j) FEHE3 cold outlet temperature, (k) furnace duty, (l) cooler duty, (m) quench temperature, (n) reactor inlet temperature, (o) separator temperature, (p) stabilizer column tray temperature, (q) product column tray 12 temperature, (r) product column tray 17 temperature, (s) recycle column tray temperature





(i) (j)





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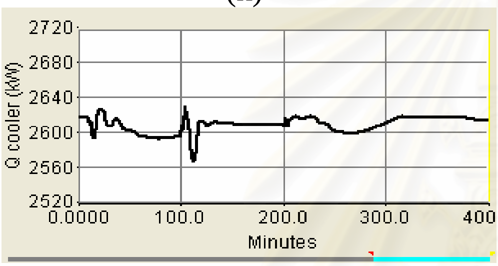

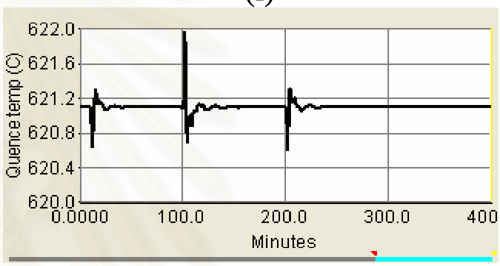

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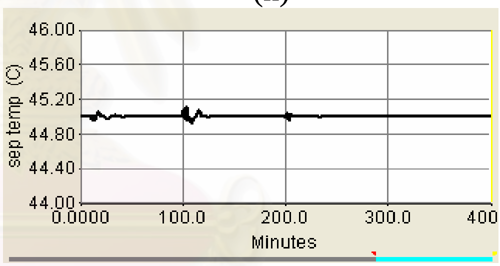
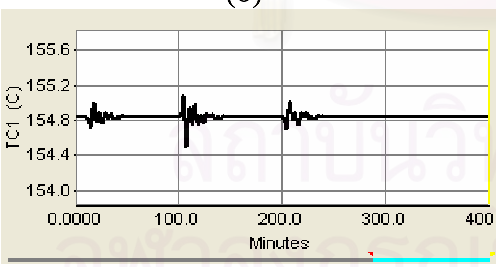
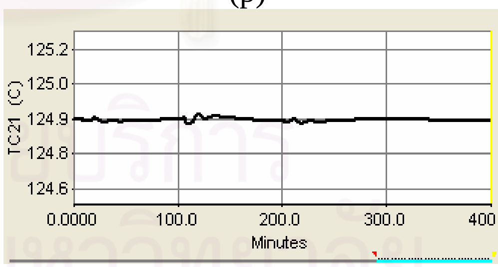


(o) (p)





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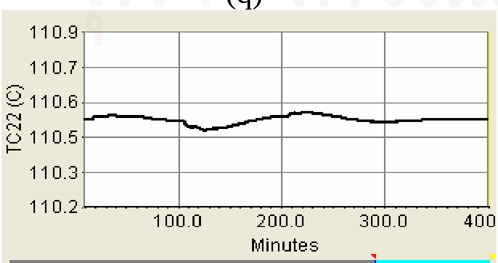
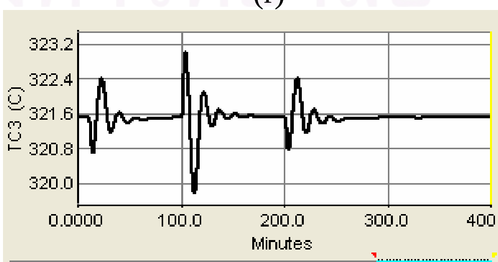
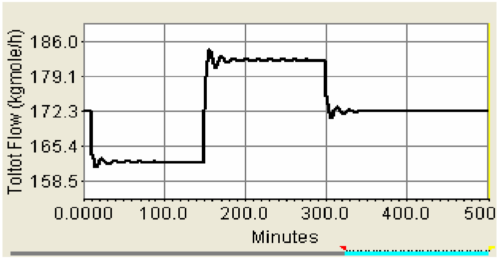


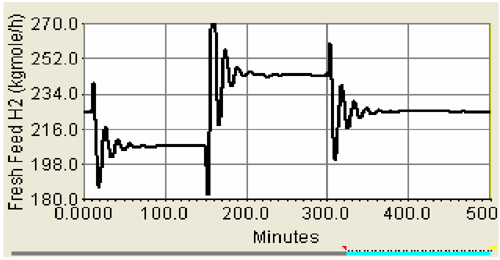


(s) (t)

Figure 6.38 (continue)



(a)



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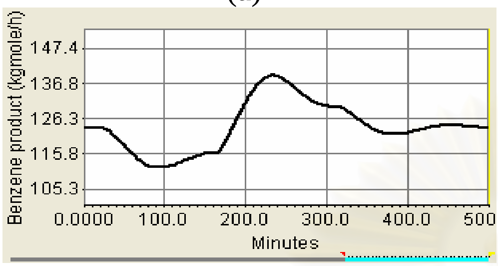

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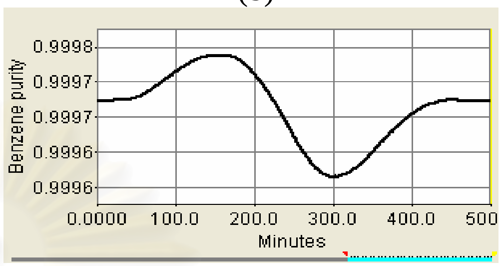

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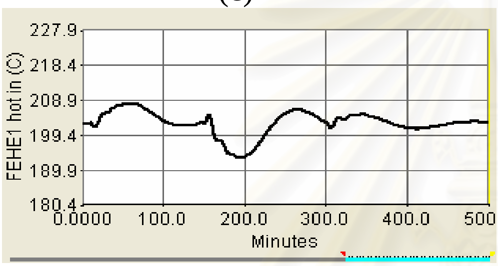

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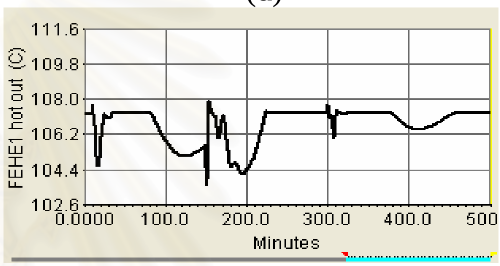

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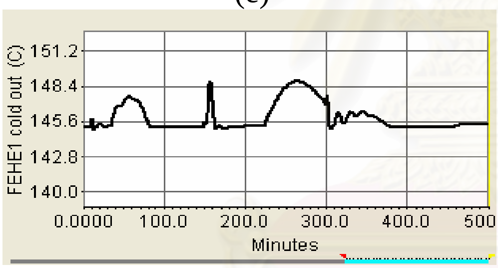
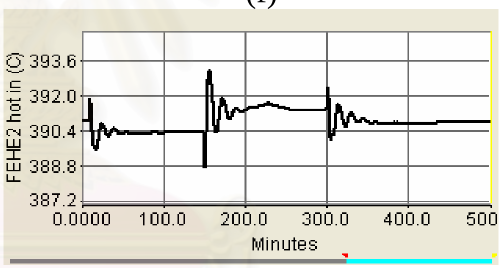
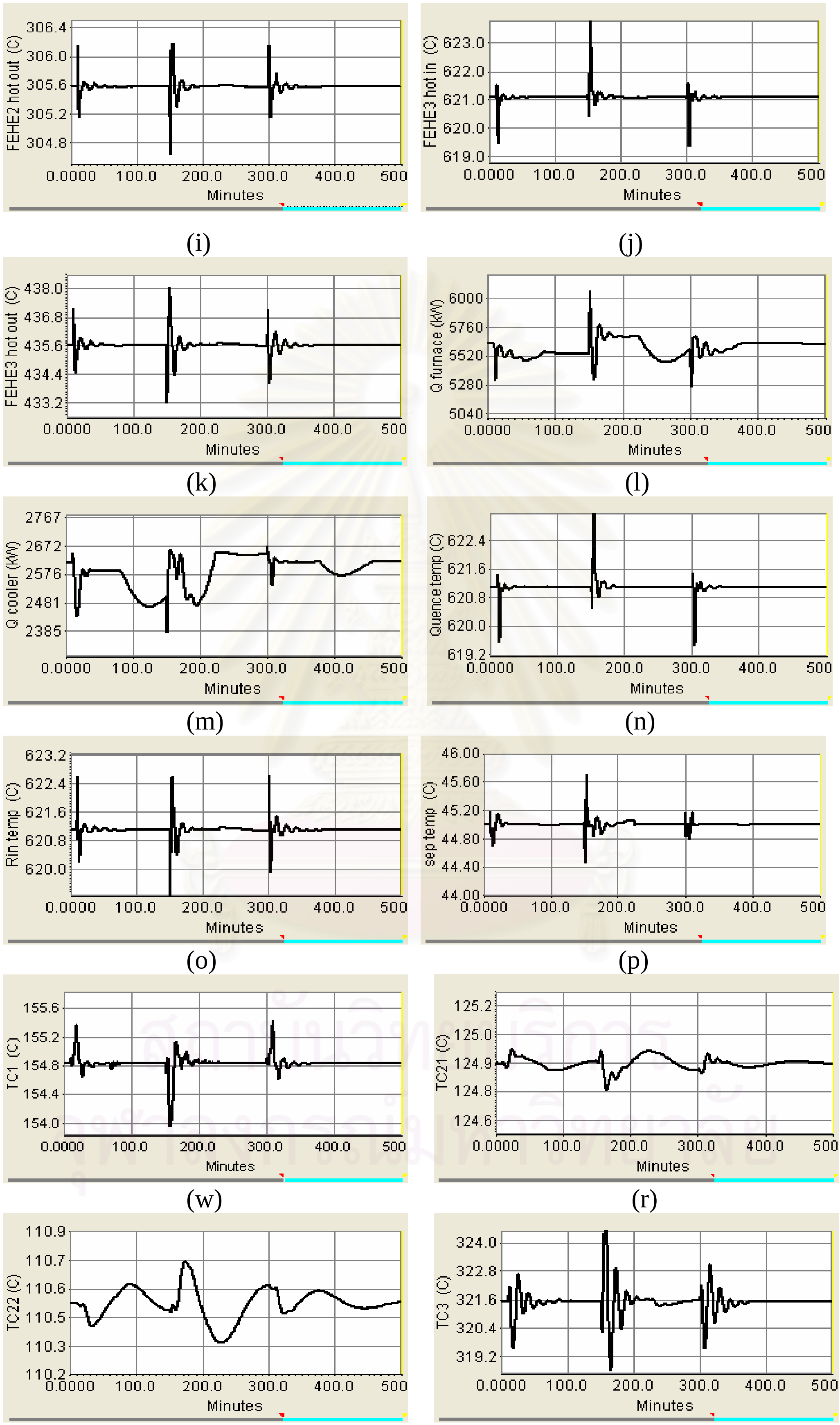


(g)



(h)

Figure 6.39 Dynamic Responses of the HDA Process Alternative 6 with four Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates (Reactor Feed Stream):CS3, where: (a) recycle toluene flowrates, (b) fresh feed hydrogen flowratres, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) FEHE3 hot inlet temperature, (k) FEHE3 hot outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray 12 temperature, (s) product column tray 17 temperature, (t) recycle column tray temperature



(i) (j)

(k) (l)

(m) (n)

(o) (p)

(w) (r)

(s) (t)

Figure 6.39 (continue)

6.5.2 Dynamic Simulation Results for HDA Process Alternative 6 with Minimum Auxiliary Utility Units

Three disturbance loads are also used to evaluate the dynamic performance of the previous control structure (CS1, CS2 and CS3) for HDA process alternative 6 with minimum auxiliary utility units.

6.5.2.1 Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream)

Figures 6.40, 6.43 and 6.46 show the dynamic responses of the control systems of HDA process alternative 6 with minimum auxiliary utility units to a change in the heat load disturbance of cold stream (reactor feed stream). In order to make this disturbance, first the fresh toluene feed temperature is decreased from 30 to 20^{o}C at time equals 10 minutes, and the temperature is increased from 20 to 40^{o}C at time equals 100 minutes, then its temperature is returned to its nominal value of 30^{o}C at time equals 200 minutes (Figures 6. 40.a, 6.43.a and 6.46.a).

As can be seen, the dynamic responses for HDA process alternative 6 with minimum auxiliary utility units and with four auxiliary utility units are same in CS1, CS2 and CS3.

6.5.2.2 Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream)

Figures 6.41, 6.44 and 6.47 shows the dynamic responses of the control systems of HDA process alternative 6 with three auxiliary utility units to a change in the heat load disturbance of hot stream (the hot reactor product). This disturbance is made as follows: first the set point of FEHE3-hot-inlet temperature controller (i.e. TCX1 in Figures 6.10 6.11 and 6.12) is decreased from 621.1 to 611.1oC at time equals 10 minutes, and the temperature is increased from 611.1 to 631.1oC at time equals 100 minutes, then its temperature is returned to its nominal value of 621.1oC at time equals 200 minutes (Figures 6.41.a, 6.44.a and 6.47.a). As can be seen, this temperature response is very fast (Figures 6.41.a, 6.44.a and 6.47.a), the new steady state is reached quickly (Figures 6.41.a, 6.44.a and 6.47.a).

As can be seen, the dynamic responses for HDA process alternative 6 with minimum auxiliary utility units and with three auxiliary utility units are same in CS1, CS2 and CS3.

6.5.2.3 Change in the Total Toluene Feed Flowrates

Figures 6.42, 6.45 and 6.48 shows the dynamic responses of the control systems of HDA process alternative 6 with three auxiliary utility units to a change in the total toluene flowrates. This disturbance is made by decreasing toluene flowrates from 172.3 to 162.3 kgmole/h at time equals 10 minutes, and the flowrates is increased from 162.3 to 182.3 kgmole/h at time equals 100 minutes, then its flowrates is returned to its nominal value of 172.3 kgmole/h at time equals 200 (Figures 6.42.a , 6.45.a and 6.48.a).

As can be seen, the dynamic responses for HDA process alternative 6 with minimum auxiliary utility units and with three auxiliary utility units are same in CS1, CS2 and CS3.

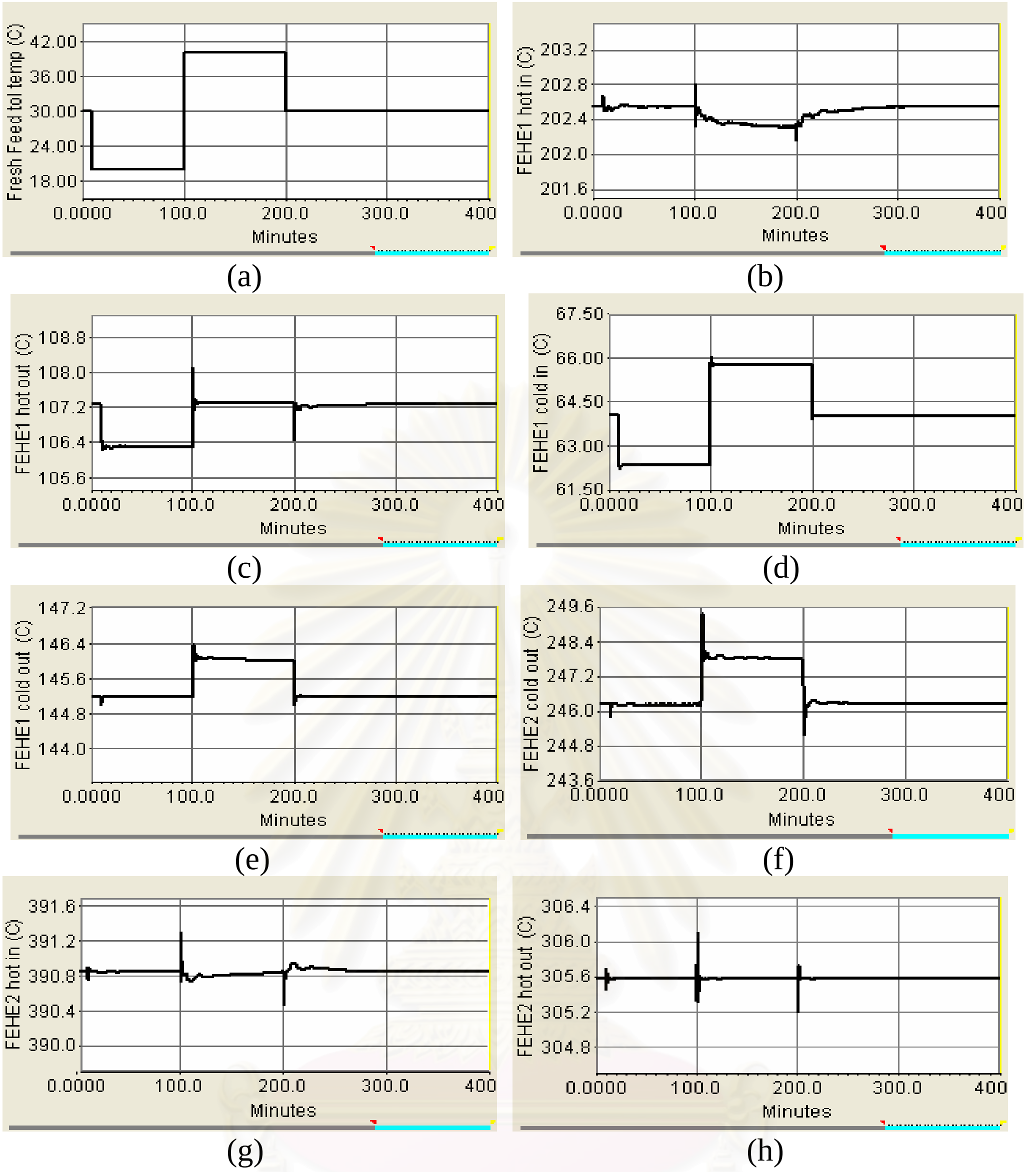
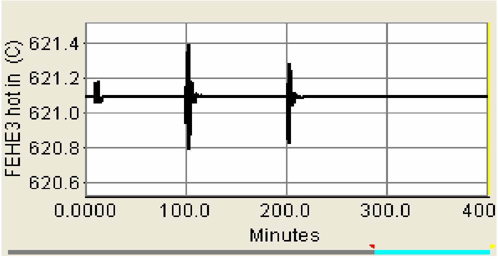
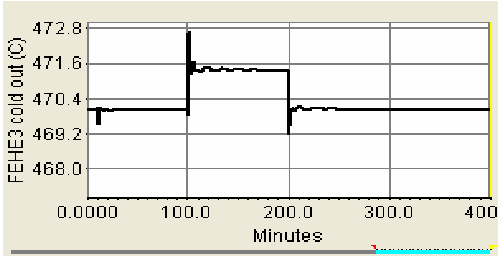


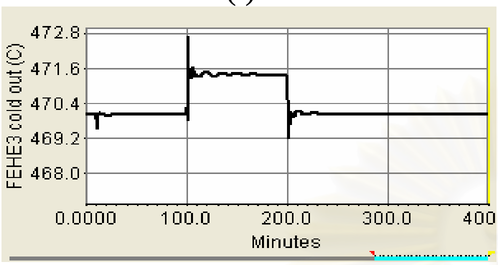
Figure 6.40 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE3 hot inlet temperature, (j) FEHE3 hot outlet temperature(k) FEHE3 cold outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray temperature, (s) recycle column tray temperature



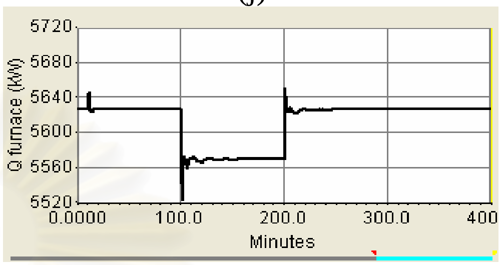
(i)



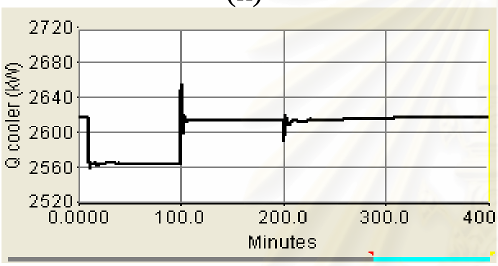
(j)



(k)



(l)



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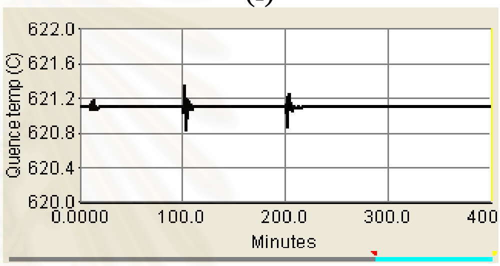

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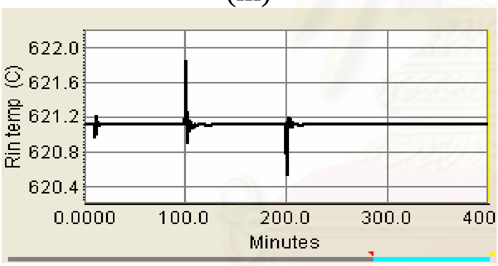
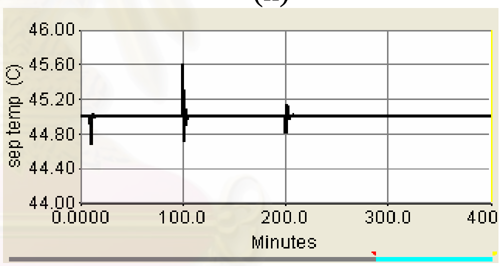
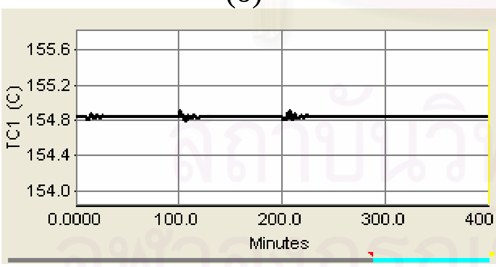


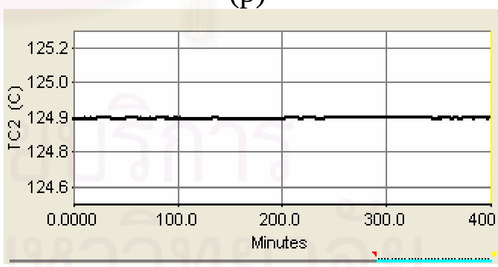
(o)



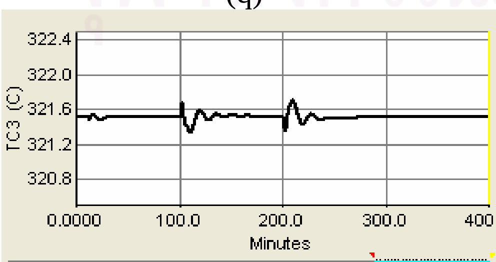
(p)



(q)



(r)



(s)

Figure 6.40 (continue)

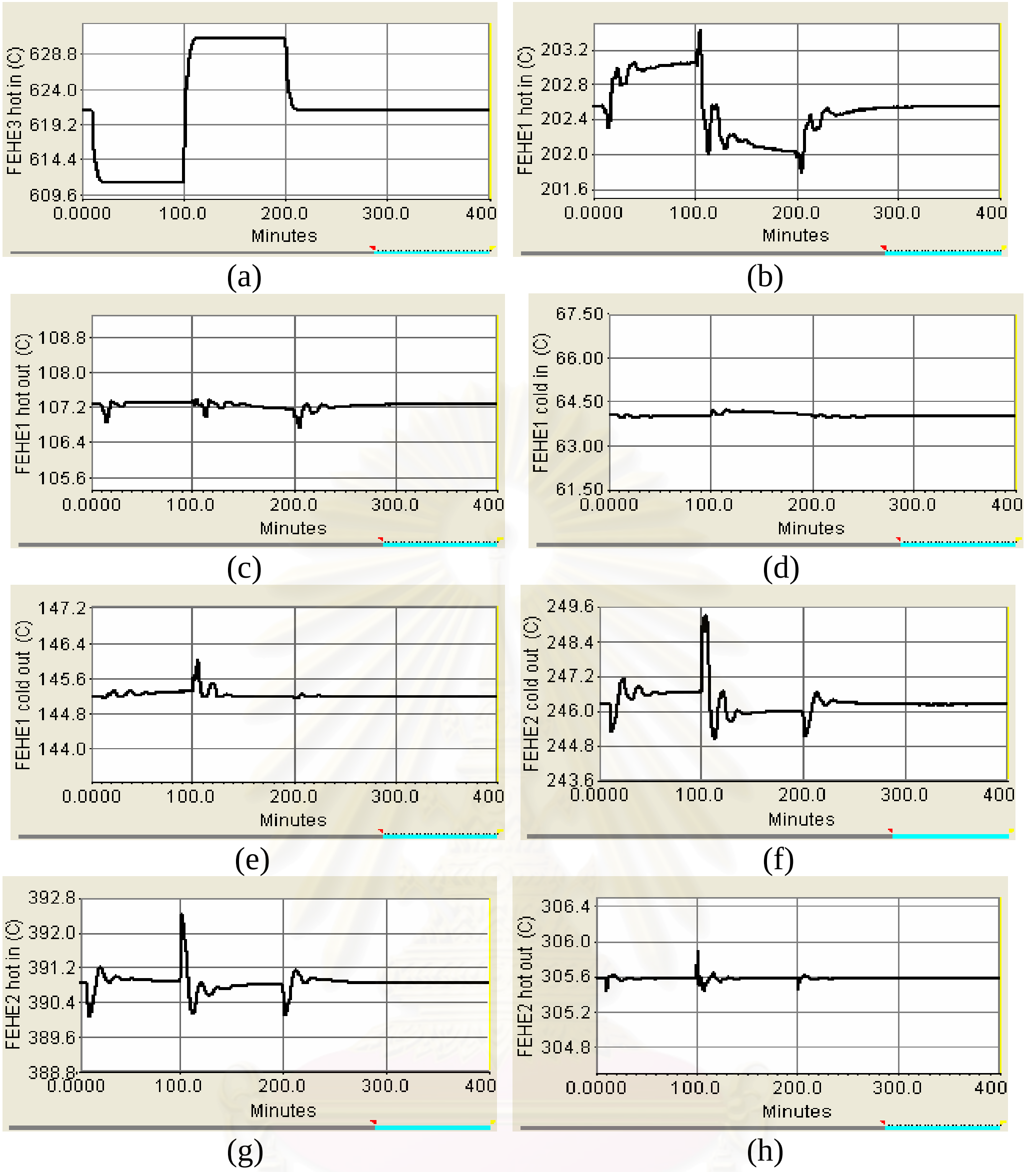


Figure 6.41 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature,(i) FEHE3 hot outlet temperature, (j) FEHE3 cold outlet temperature, (k) furnace duty, (l) cooler duty, (m) quench temperature, (n) reactor inlet temperature, (o) separator temperature, (p) stabilizer column tray temperature, (q) product column tray temperature, (r) recycle column tray temperature

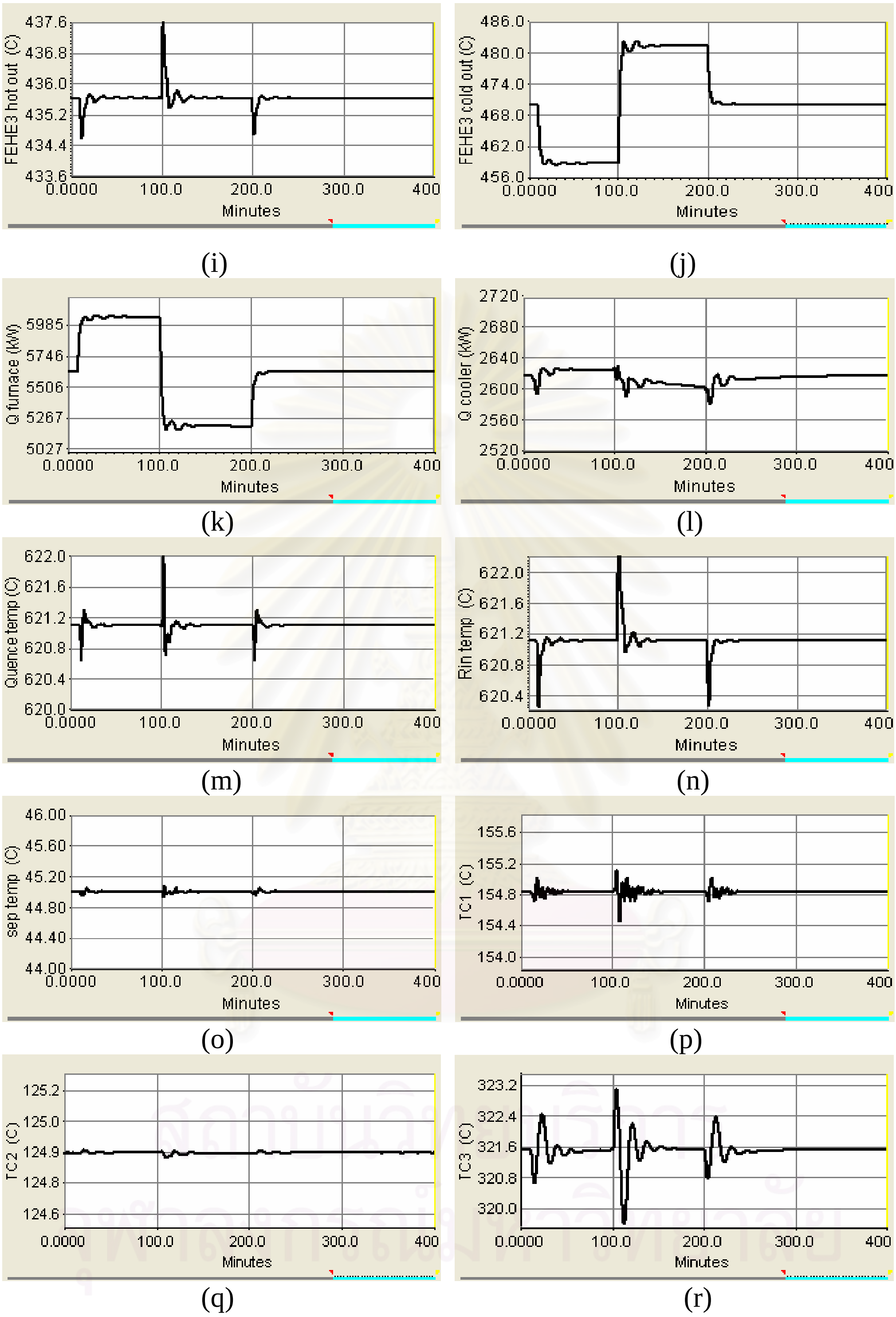


Figure 5.41 (continue)

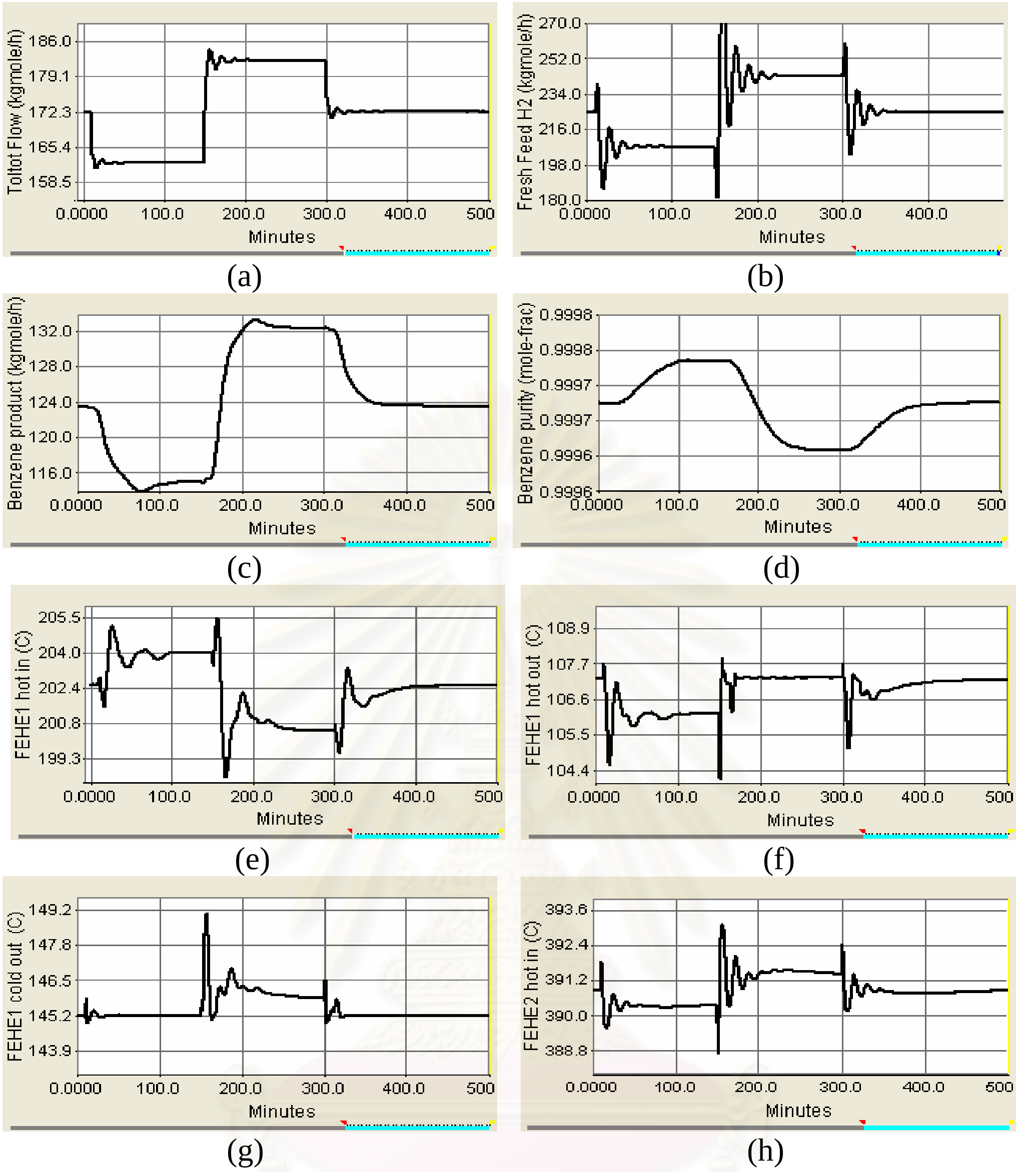
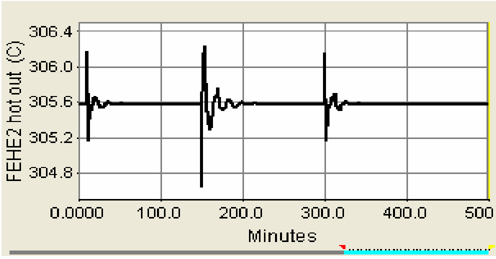


Figure 6.42 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates (Reactor Feed Stream):CS1, where: (a) recycle toluene flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) FEHE3 hot inlet temperature, (k) FEHE3 hot outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray temperature, (s) recycle column tray temperature



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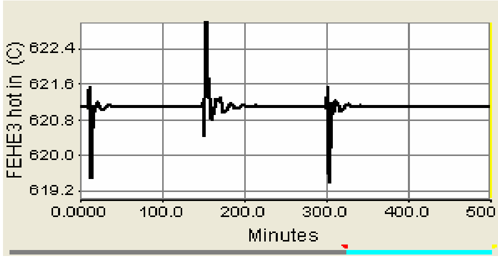

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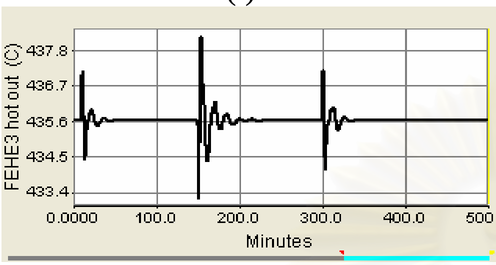

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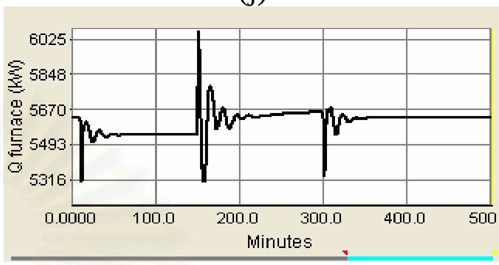

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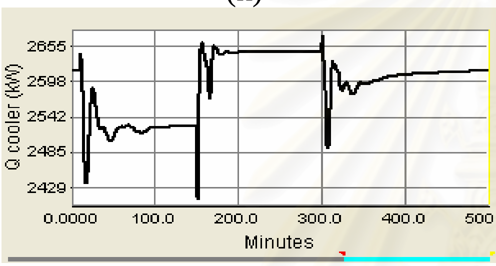

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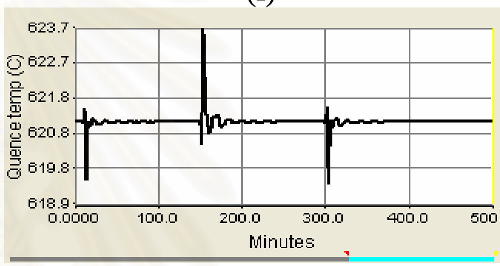

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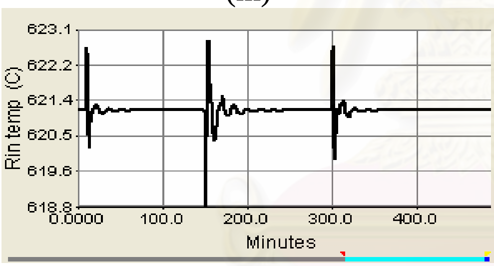

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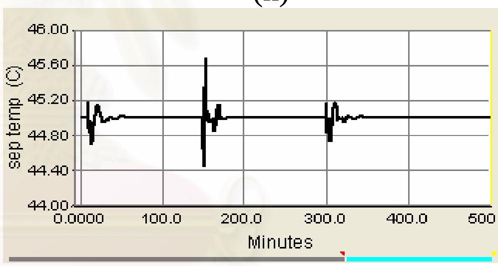

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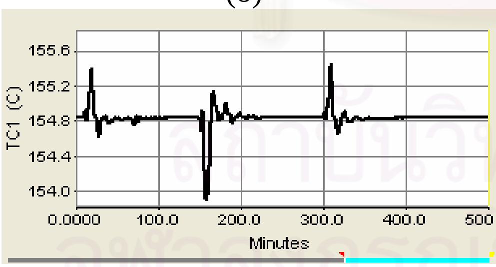

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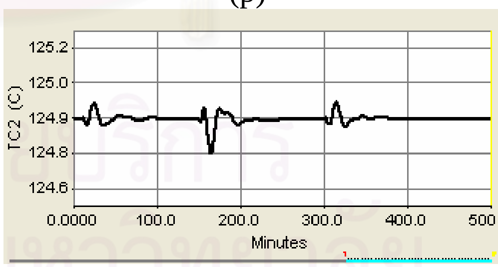

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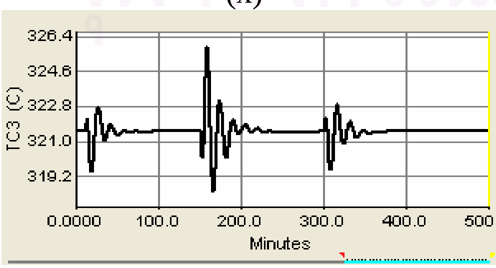


(s)

Figure 6.42 (continue)

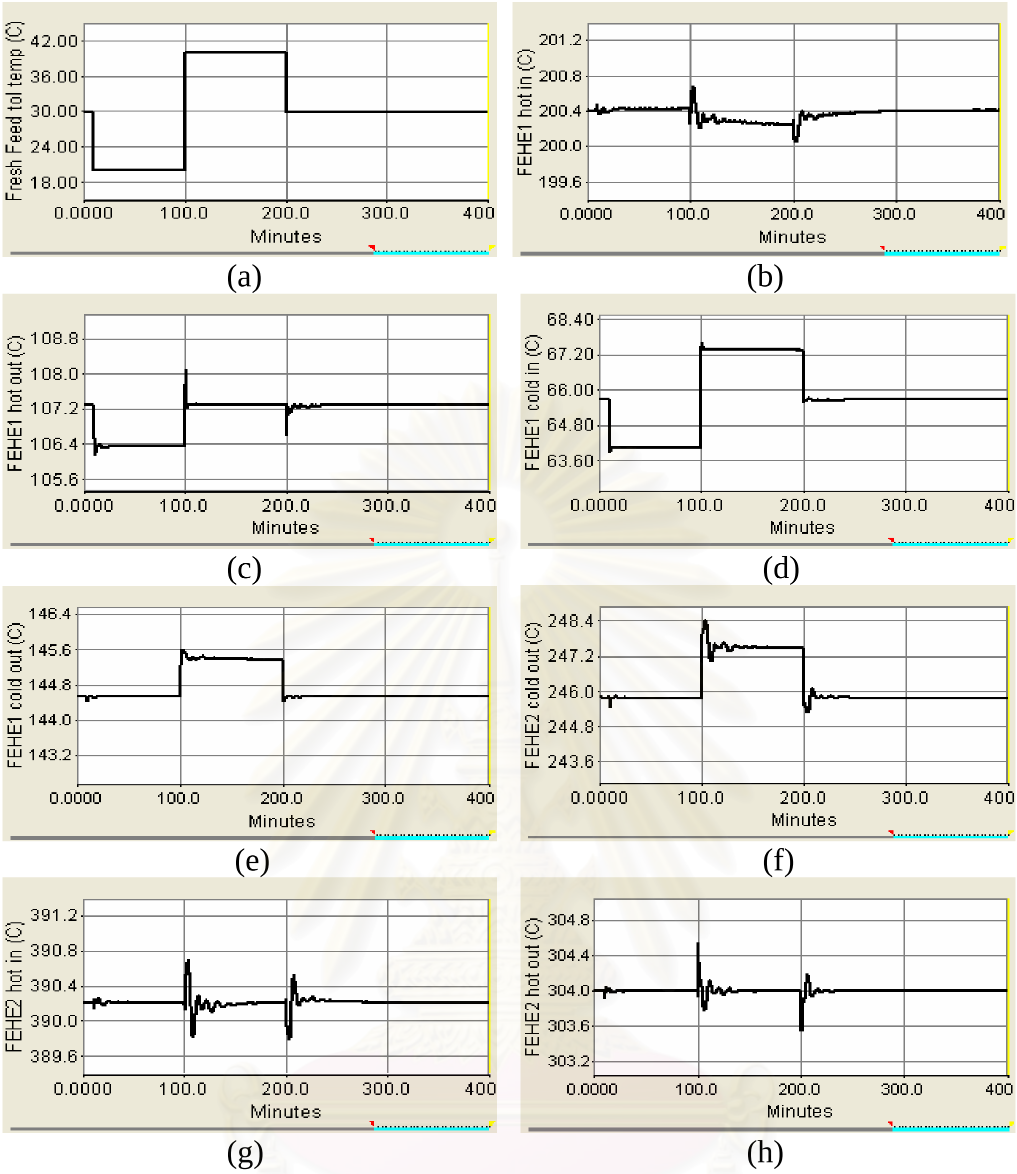
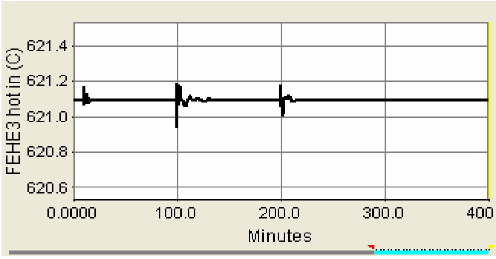
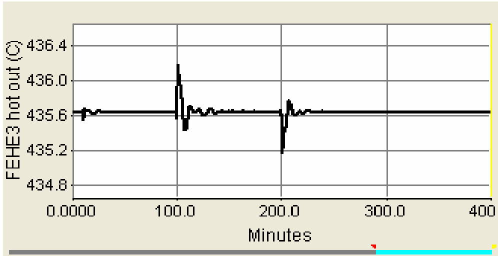


Figure 6.43 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE3 hot inlet temperature, (j) FEHE3 hot outlet temperature(k) FEHE3 cold outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray temperature, (s) recycle column tray temperature



(i)



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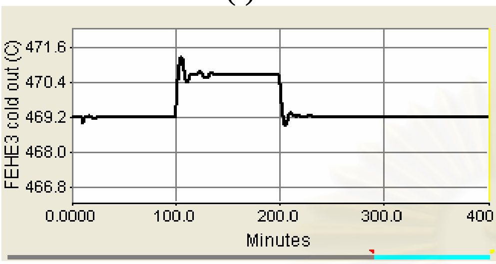
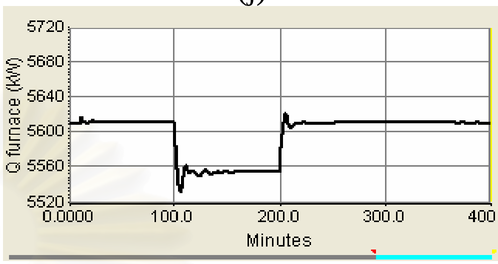
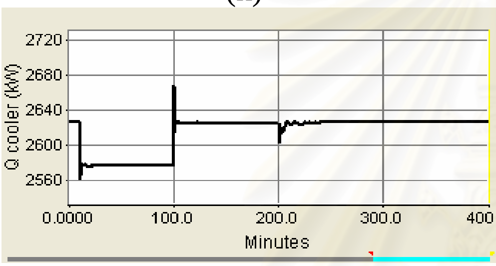


(k)



(l)



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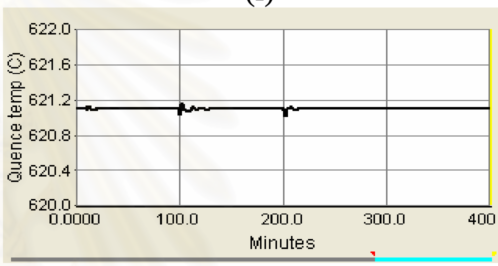

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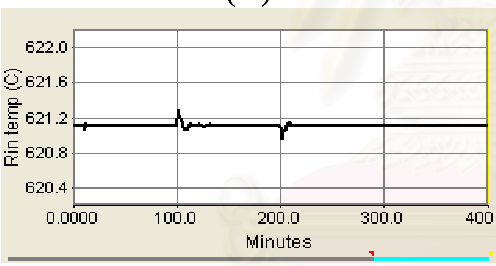
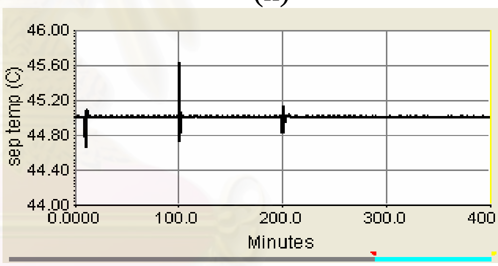


(o)



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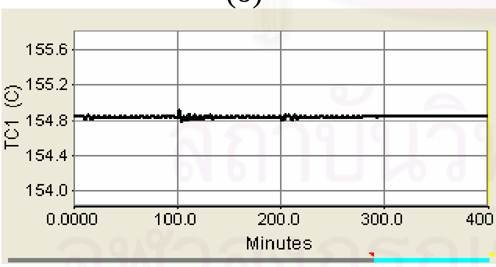

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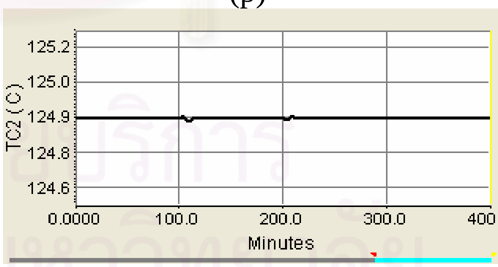
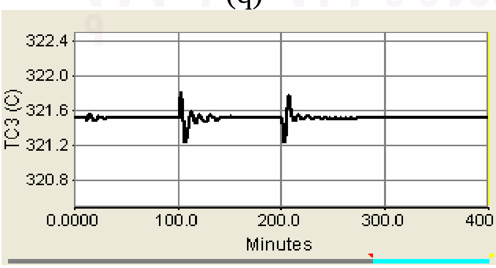


(r)



(s)

Figure 6.43 (continue)

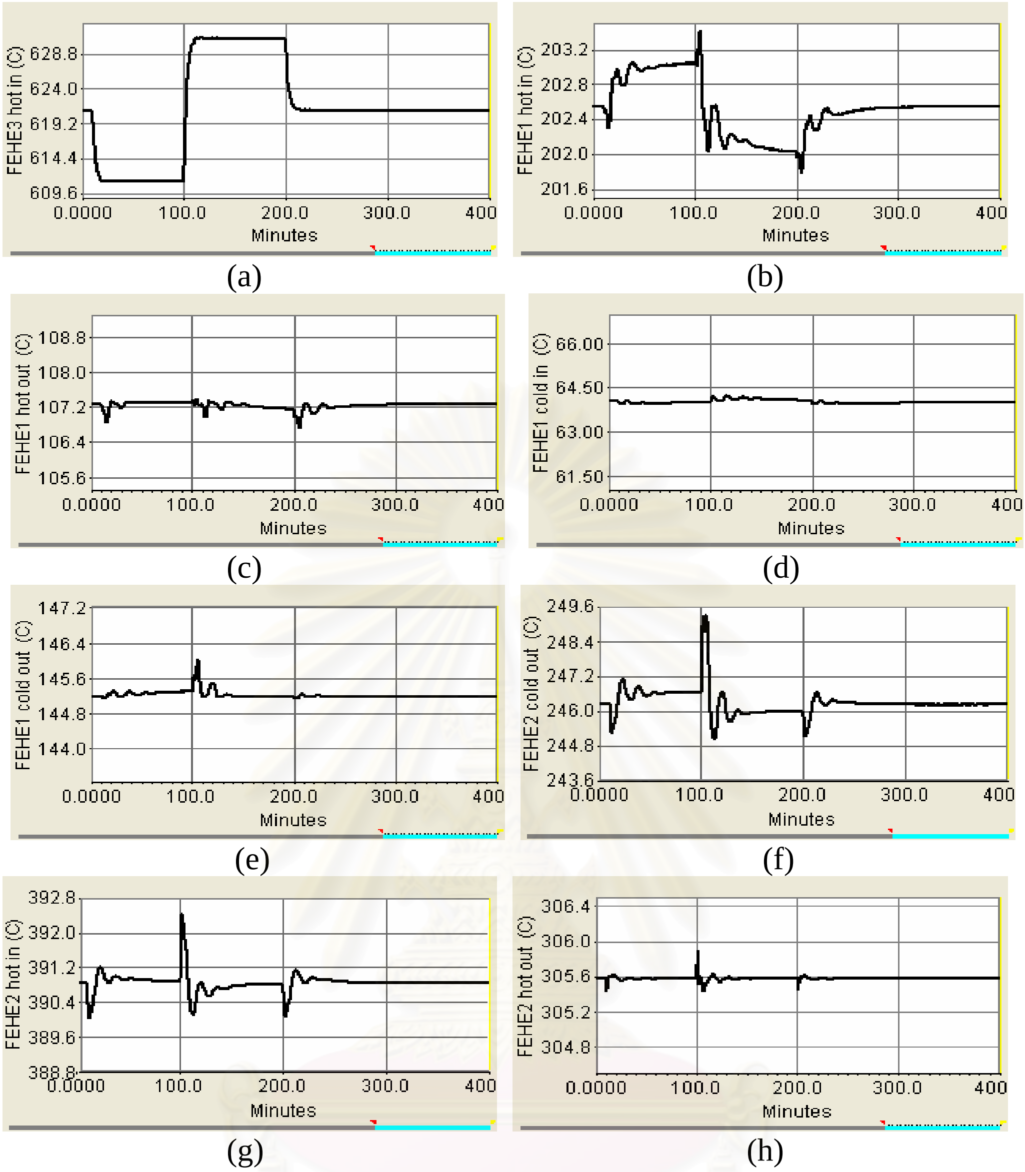
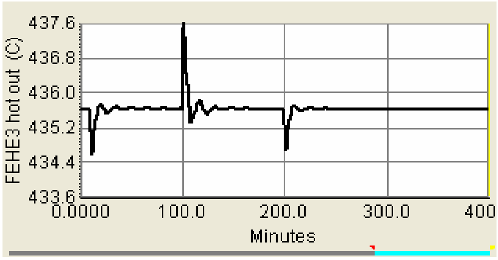


Figure 6.44 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature,(i) FEHE3 hot outlet temperature, (j) FEHE3 cold outlet temperature, (k) furnace duty, (l) cooler duty, (m) quench temperature, (n) reactor inlet temperature, (o) separator temperature, (p) stabilizer column tray temperature, (q) product column tray temperature, (r) recycle column tray temperature



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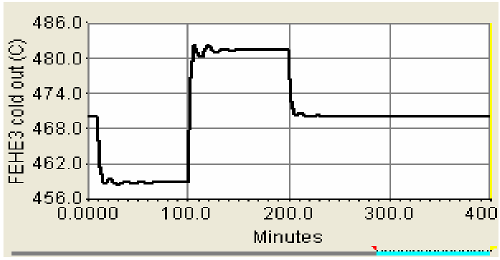

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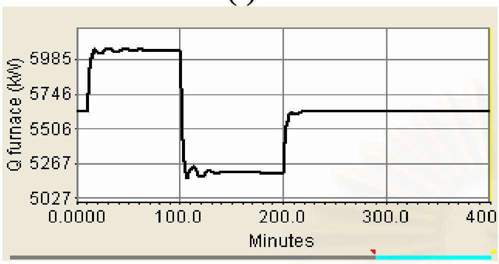

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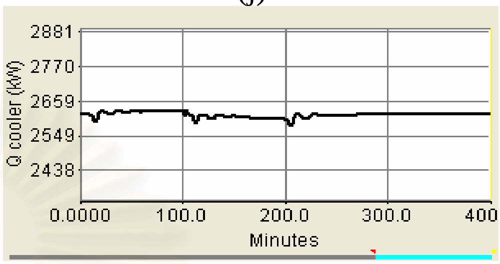

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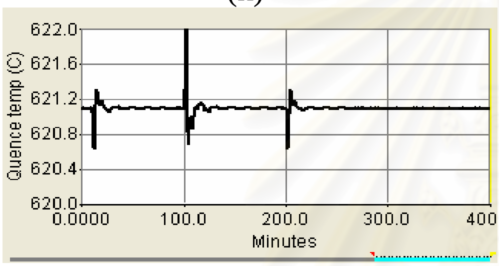

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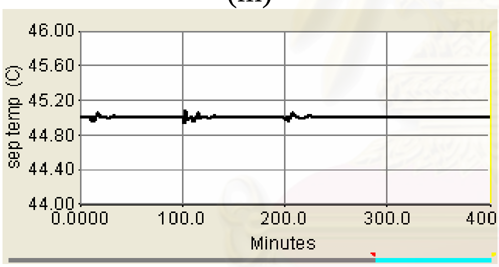

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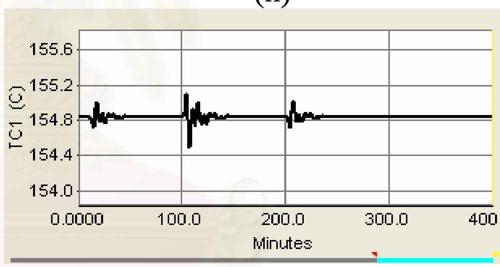

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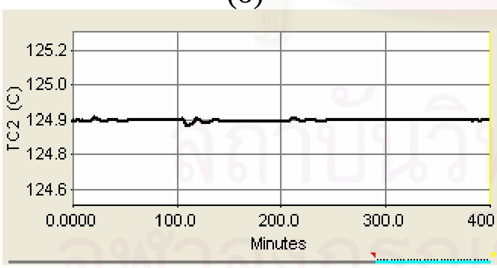

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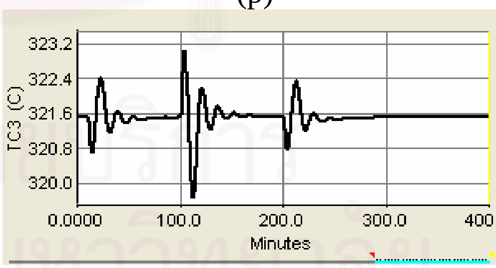

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Figure 6.44 (continue)

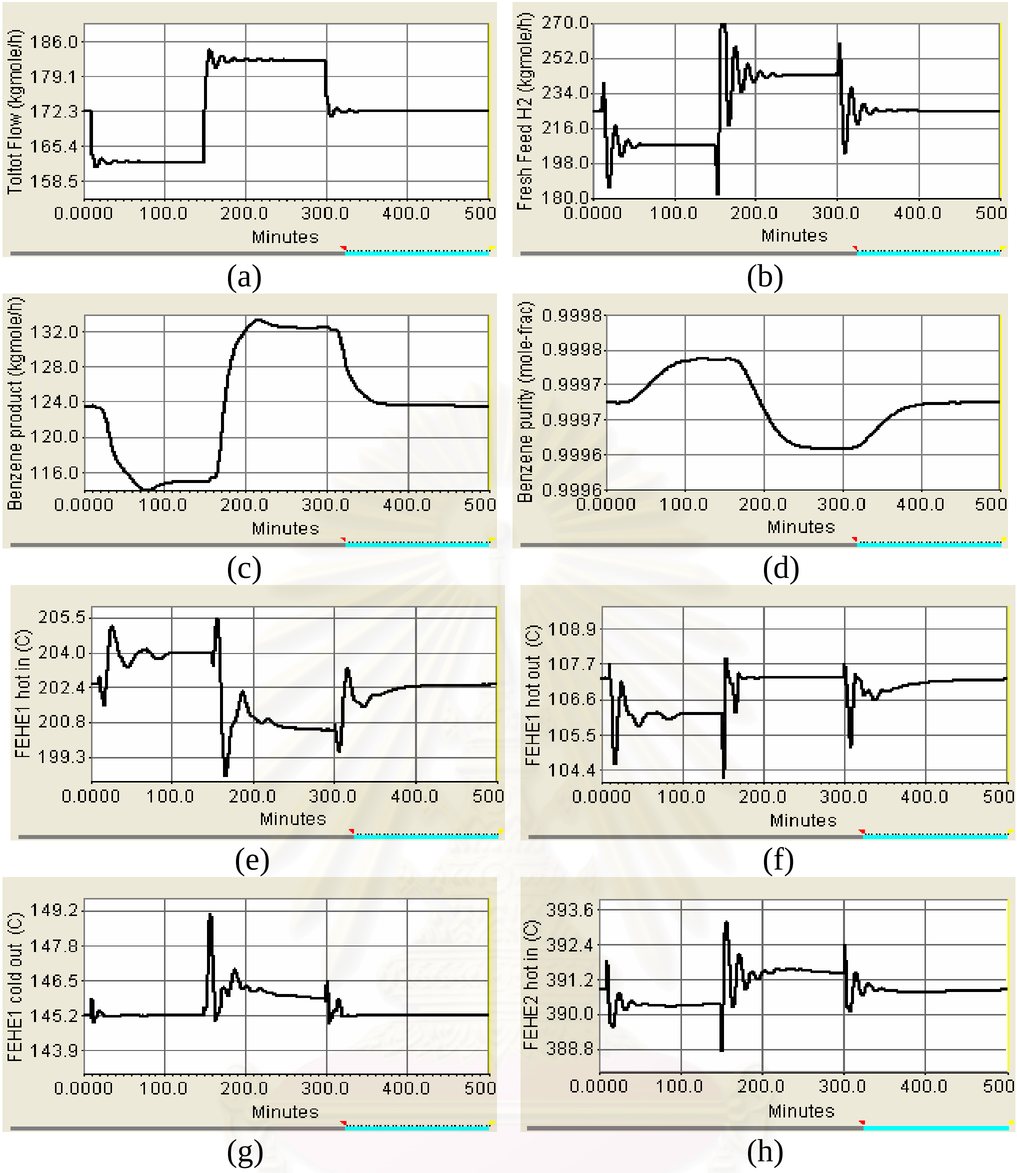
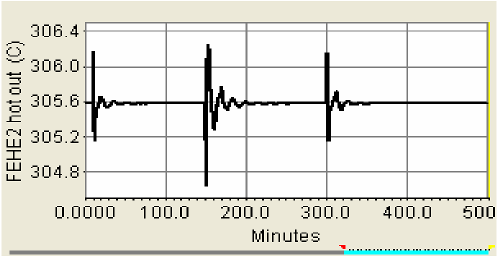


Figure 6.45 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates (Reactor Feed Stream):CS2, where: (a) recycle toluene flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) FEHE3 hot inlet temperature, (k) FEHE3 hot outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray temperature, (s) recycle column tray temperature



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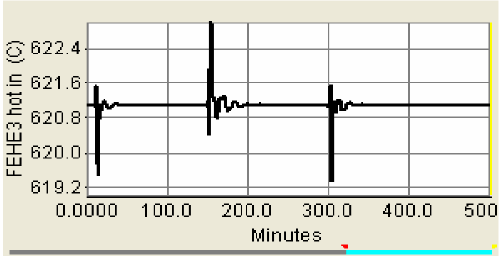

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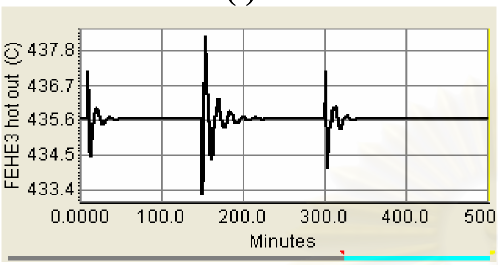

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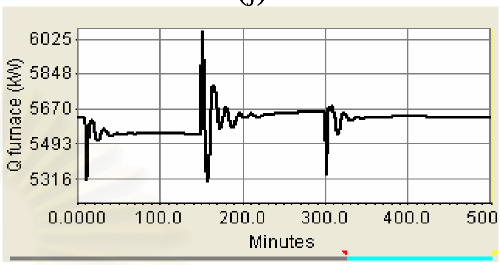

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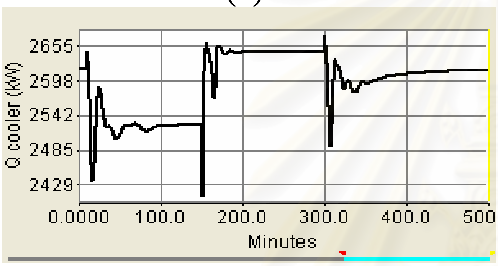

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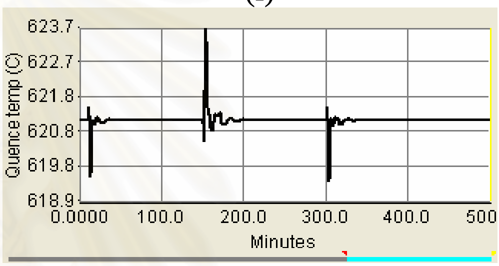

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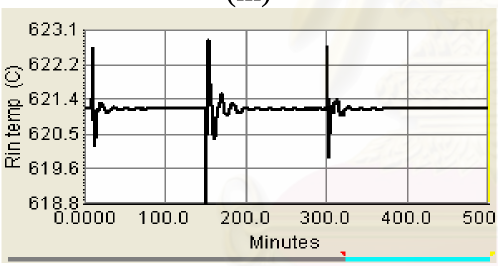

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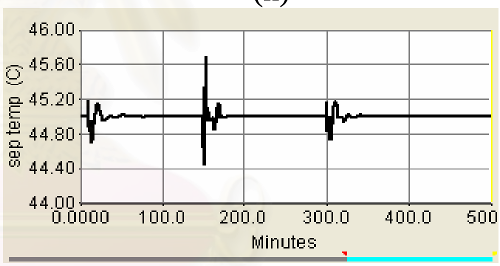

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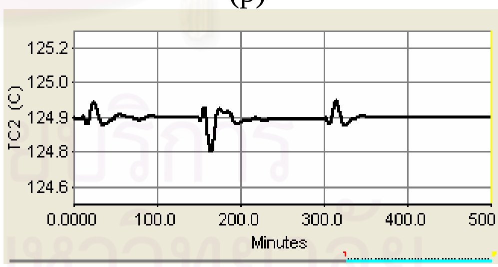

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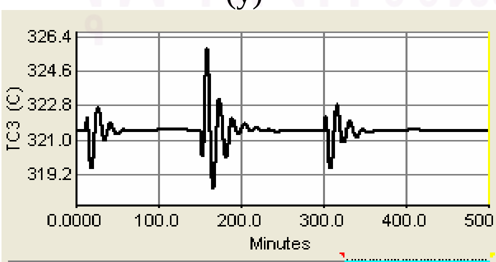


(s)

Figure 6.45 (continue)

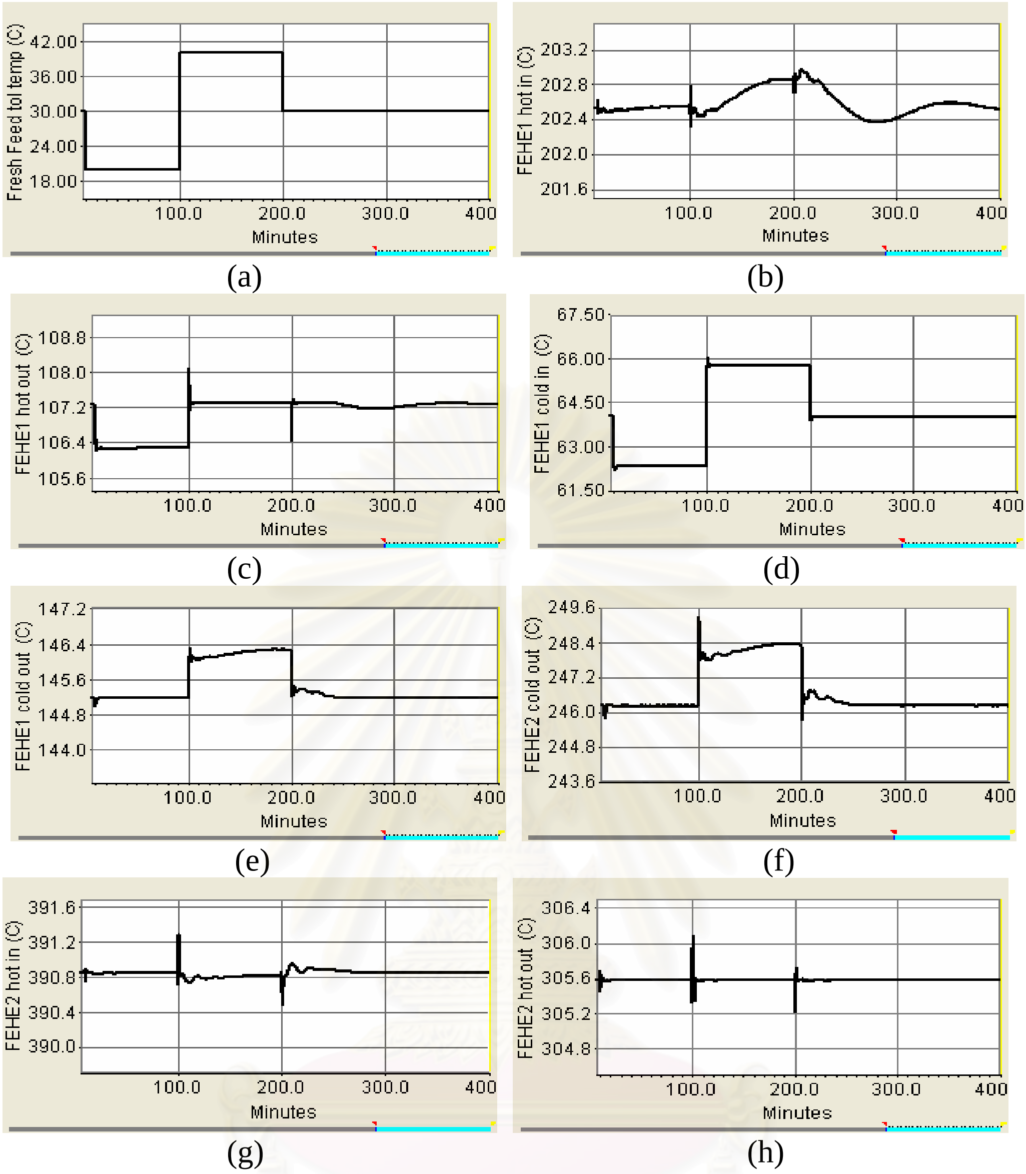
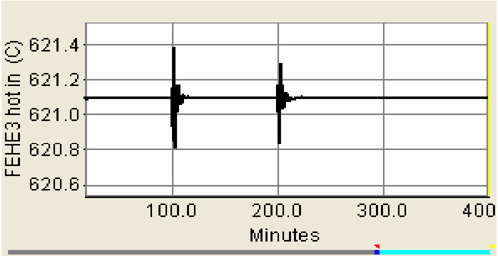


Figure 6.46 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) fresh feed toluene temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE3 hot inlet temperature, (j) FEHE3 hot outlet temperature(k) FEHE3 cold outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray 12 temperature, (s) product column tray 17 temperature, (t) recycle column tray temperature



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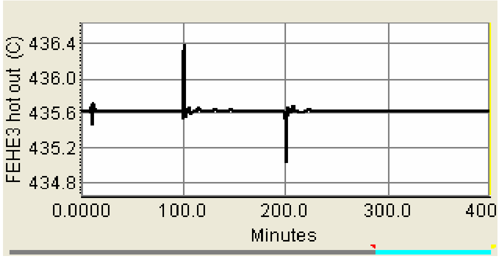

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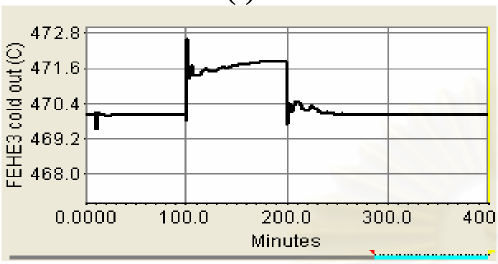

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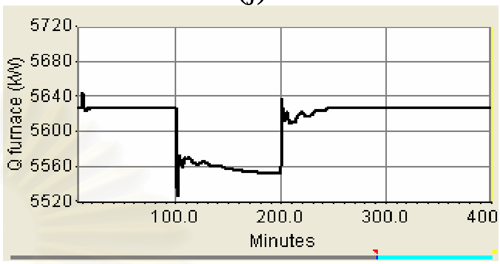

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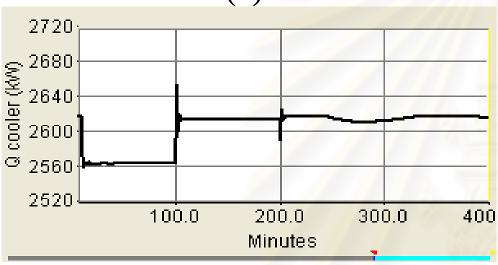

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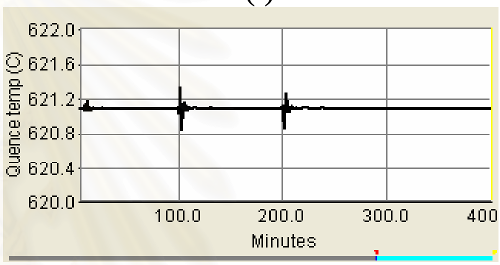

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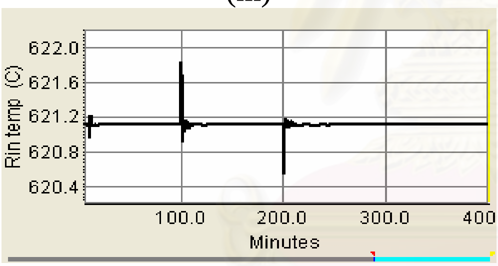

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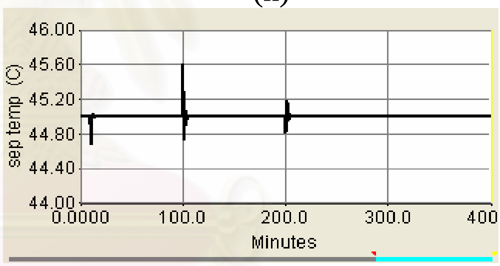

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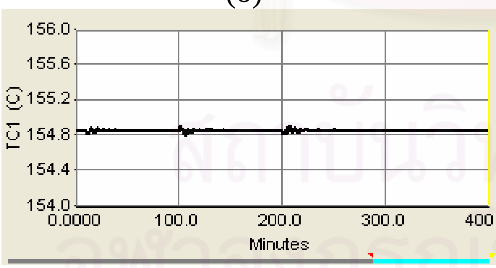

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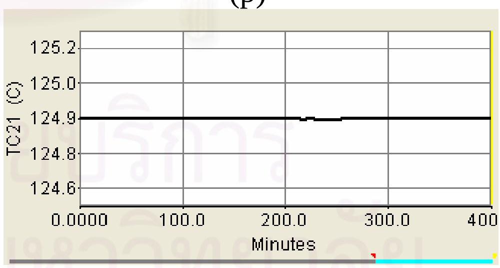

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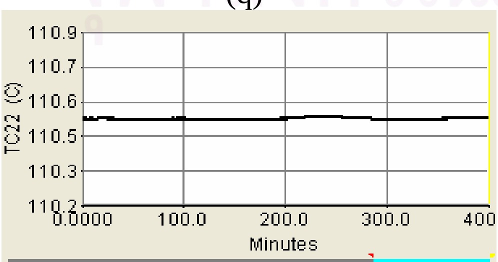

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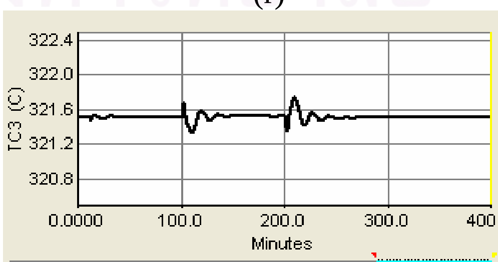

(t)

Figure 6.46 (continue)

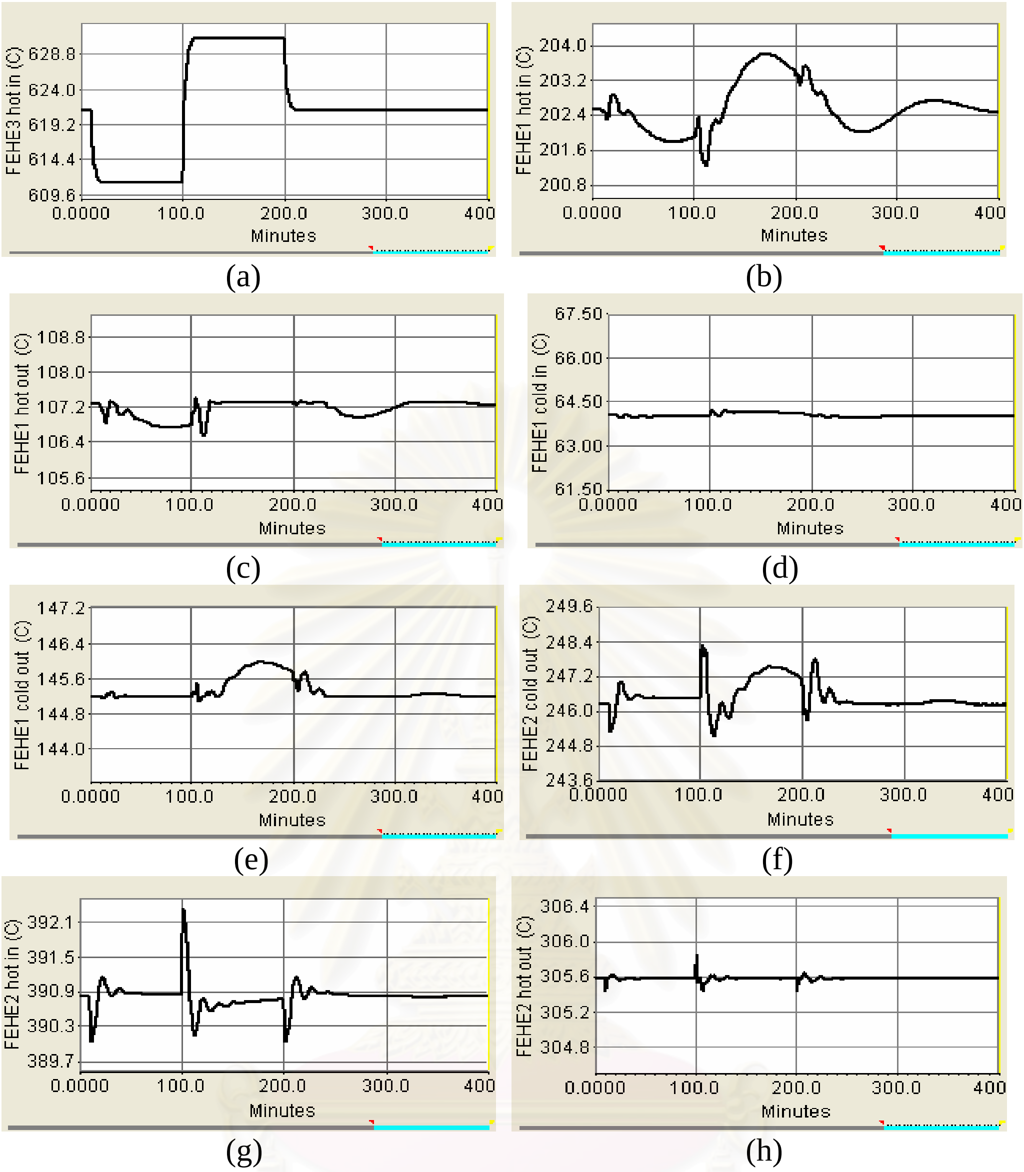

Figure 6.47 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE3 hot inlet temperature, (b) FEHE1 hot inlet temperature, (c) FEHE1 hot outlet temperature, (d) FEHE1 cold inlet temperature, (e) FEHE1 cold outlet temperature, (f) FEHE2 hot inlet temperature, (g) FEHE2 hot outlet temperature(h) FEHE2 cold outlet temperature,(i) FEHE3 hot outlet temperature, (j) FEHE3 cold outlet temperature, (k) furnace duty, (l) cooler duty, (m) quench temperature, (n) reactor inlet temperature, (o) separator temperature, (p) stabilizer column tray temperature, (q) product column tray 12 temperature, (r) product column tray 17 temperature, (s) recycle column tray temperature

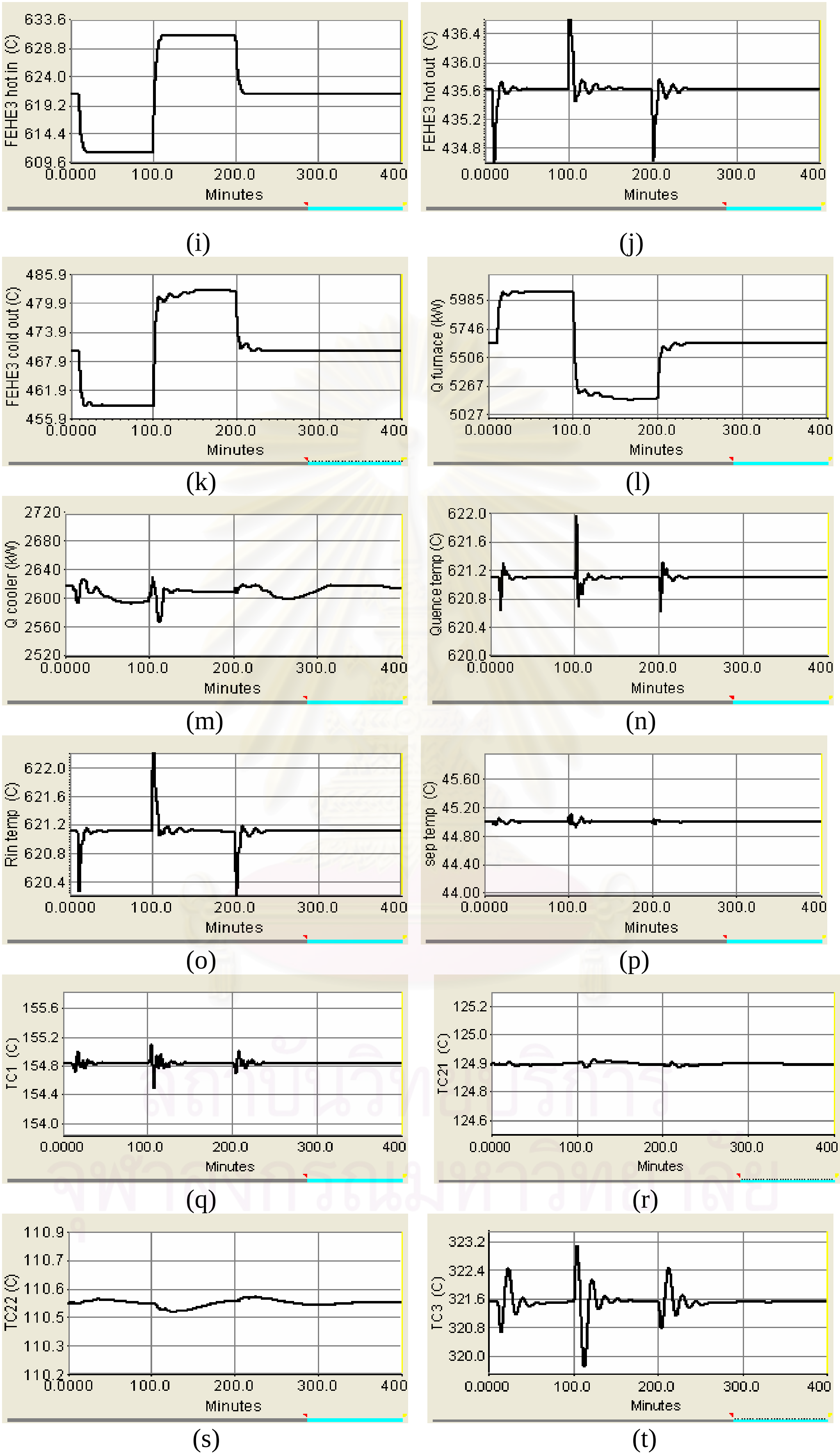

Figure 6.47 (continue)

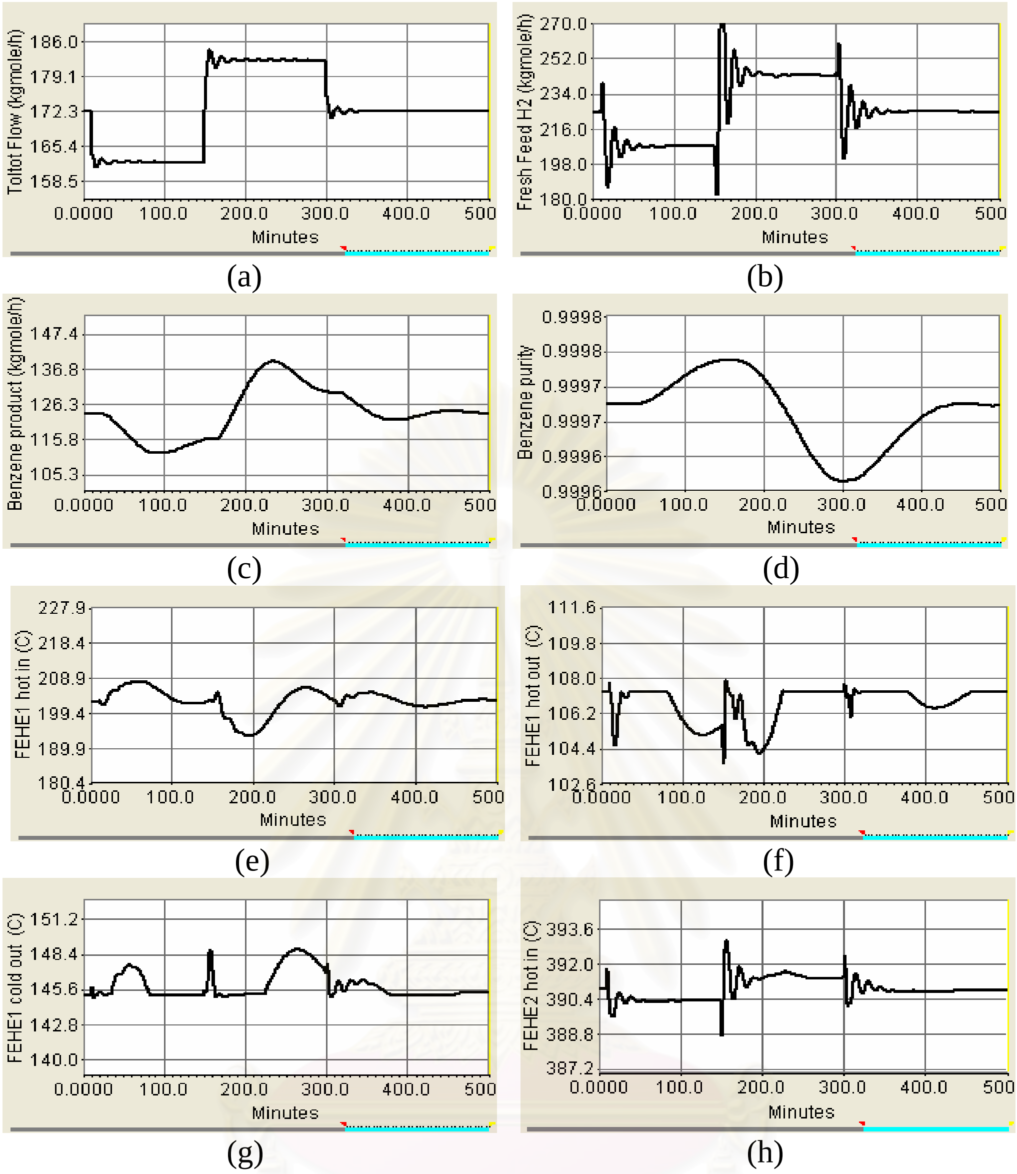

Figure 6.48 Dynamic Responses of the HDA Process Alternative 6 with Minimum Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates (Reactor Feed Stream):CS3, where: (a) recycle toluene flowrates, (b) fresh feed hydrogen flowrates, (c) benzene product flowrates, (d) benzene purity in the product stream, (e) FEHE1 hot inlet temperature, (f) FEHE1 hot outlet temperature, (g) FEHE1 cold outlet temperature, (h) FEHE2 hot inlet temperature, (i) FEHE2 hot outlet temperature, (j) FEHE3 hot inlet temperature, (k) FEHE3 hot outlet temperature, (l) furnace duty, (m) cooler duty, (n) quench temperature, (o) reactor inlet temperature, (p) separator temperature, (q) stabilizer column tray temperature, (r) product column tray 12 temperature, (s) product column tray 17 temperature, (t) recycle column tray temperature

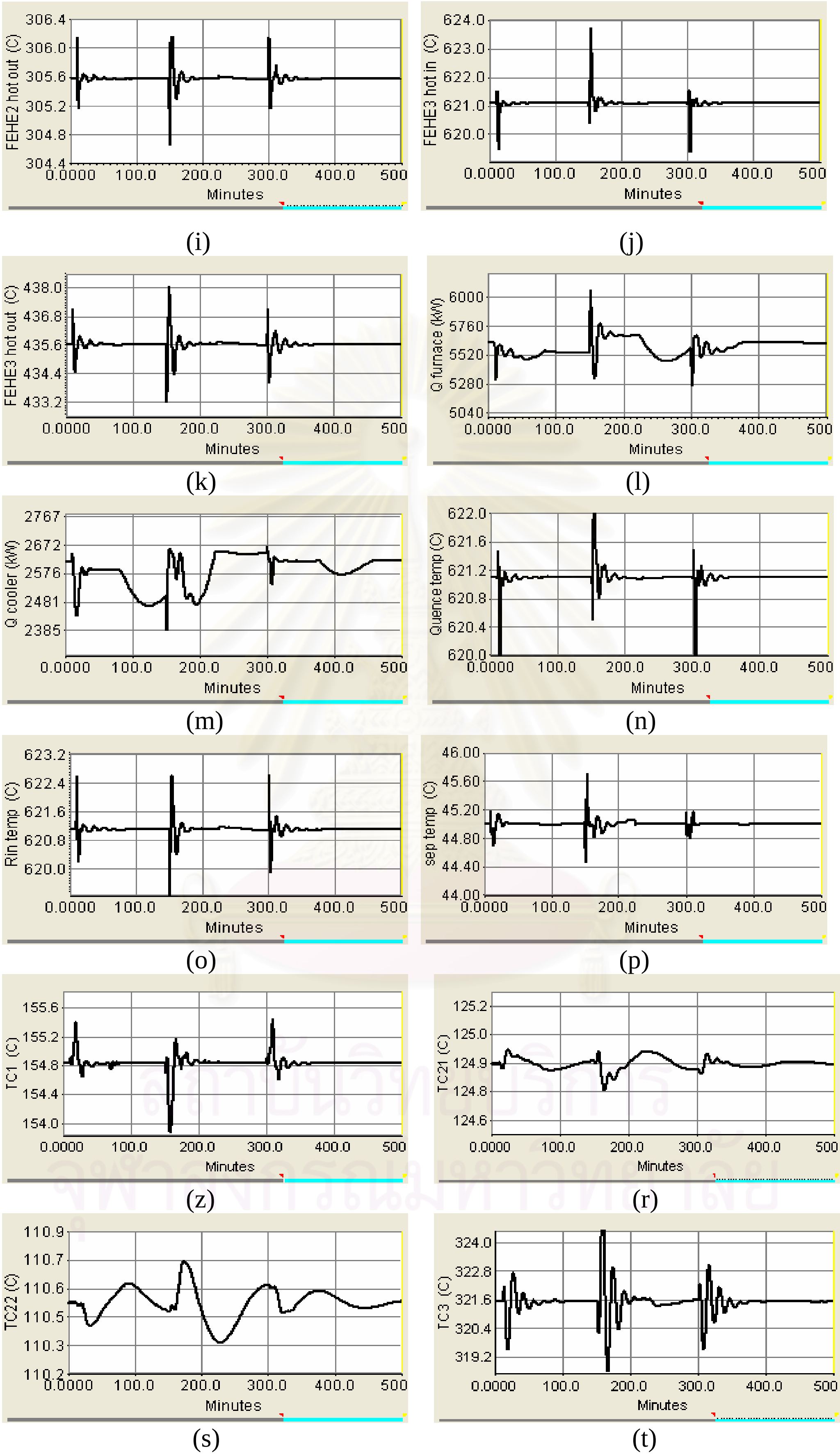

Figure 6.48 (continue)

6.6 Evaluation of the Dynamic Performance

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

$$IAE = \int |\varepsilon(t)|\, dt$$

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

In this work, IAE method is used to evaluate the dynamic performance of the designed control systems. Figures 6.13 – 6.15 show the total IAE results HDA process alternative 5 and figures 6.16 – 18 show the total IAE results HDA process alternative 6.

As can be seen, The IAE of HDA process alternative 5 with minimum auxiliary utility units for all control structures has a little bit deference form the IAE of HDA process alternative 5 with three auxiliary utility units suggested by luyben. It means the control performance of both is same. So the HDA process alternative 5 is employed only one auxiliary utility unit that it can decrease the equipment cost and the control performance still has effective. As can be seen, CS1 is the best control structure for handle disturbances due to it gives better control performances.

Again, The IAE of HDA process alternative 6 with minimum auxiliary utility units has a little bit deference form the IAE of HDA process alternative 6 with four auxiliary utility units suggested by Luyben that it means the control performance of both is same. So the HDA process alternative 6 is employed only one auxiliary utility unit and the control performance still has effective. As can be seen, CS1 is the best control structure for handle disturbances.

Table 6.13 the IAE Values of HDA Process Alternative 5 with three and Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	3 Aux.	Min.Aux.	3 Aux.	Min.Aux.	3 Aux.	Min.Aux.
TC1	2.3	2.5	2.2	2.3	2.3	2.4
TC2	0.2	0.2	0.3	0.2	0.3	0.3
TC3	2.0	2.0	2.3	2.3	1.8	1.8
TCE2h	0.7	0.7	1.8	1.8	0.8	0.8
TCR	3.5	3.5	1.7	1.7	3.7	3.8
TCQ	2.7	2.7	1.1	1.0	2.8	2.8
TCS	1.3	1.3	1.1	1.1	1.2	1.2
Total	12.8	12.9	10.3	10.4	13.0	13.2

Table 6.14 the IAE Values of HDA Process Alternative 5 with three and Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	3 Aux.	Min.Aux.	3 Aux.	Min.Aux.	3 Aux.	Min.Aux.
TC1	13.4	14.1	14.7	15.4	13.4	14.1
TC2	1.3	1.3	2.1	2.1	1.3	1.3
TC3	9.1	9.0	13.9	14.0	8.8	8.8
TCE2h	2.6	2.6	14.7	14.7	2.7	2.7
TCR	19.1	19.1	18.6	18.6	19.5	19.4
TCQ	11.5	11.5	9.2	9.3	11.6	11.6
TCS	1.2	1.2	1.3	1.3	1.1	1.1
Total	58.2	58.9	74.6	75.3	58.4	59.1

Table 6.15 the IAE Values of HDA Process Alternative 5 with three and Minimum Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	3 Aux.	Min.Aux.	3 Aux.	Min.Aux.	3 Aux.	Min.Aux.
TC1	46.3	49.2	41.3	43.7	45.7	48.5
TC2	7.9	8.0	7.1	7.1	9.6	9.7
TC3	24.3	24.4	36.0	36.5	23.3	22.6
TCE2h	5.7	5.7	24.1	24.2	5.9	5.9
TCR	28.7	28.8	22.4	22.5	29.3	29.3
TCQ	27.1	27.2	19.5	19.5	26.9	26.9
TCS	5.5	5.5	5.0	5.0	5.2	5.2
Total	145.5	148.7	155.3	158.6	145.9	148.0

Table 6.16 the IAE Values of HDA Process Alternative 6 with four and Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	4 Aux.	Min.Aux.	4 Aux.	Min.Aux.	4 Aux.	Min.Aux.
TC1	1.09	1.32	2.07	2.1	1.22	1.18
TC2	0.09	0.1	0.16	0.2	0.3	0.31
TC3	5.93	6.16	5.31	5.7	6.29	6.57
TCE2h	0.96	0.97	4.45	4.5	1.07	1.04
TCE3h	2.33	2.35	5.46	5.6	2.57	2.54
TCR	2.07	2.08	1.6	1.6	2.32	2.27
TCQ	1.63	1.62	0.78	0.8	1.74	1.7
TCS	1.45	1.47	1.46	1.4	1.52	1.49
Total	15.54	16.06	21.28	21.8	17.03	17.1

Table 6.17 the IAE Values of HDA Process Alternative 6 with four and Minimum Auxiliary Utility Units to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	4 Aux.	Min.Aux.	4 Aux.	Min.Aux.	4 Aux.	Min.Aux.
TC1	5.24	6.5	7.98	8.6	5.39	5.32
TC2	0.63	0.6	1.25	1.3	1.53	1.54
TC3	50.96	52.9	39.16	42.9	50.62	53.19
TCE2h	2.44	2.4	27.26	28.1	2.43	2.39
TCE3h	19.65	19.5	50.6	51.5	19.75	19.87
TCR	13.3	13.2	15.22	15.5	13.39	13.46
TCQ	6.95	7	6.55	6.6	6.87	6.89
TCS	1.58	1.6	2.34	2.4	1.76	1.77
Total	100.74	103.6	150.37	156.8	101.75	104.4

Table 6.18 the IAE Values of HDA Process Alternative 6 with three and Minimum Auxiliary Utility Units to a Change in the Total Toluene Feed Flowrates

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	4 Aux.	Min.Aux.	4 Aux.	Min.Aux.	4 Aux.	Min.Aux.
TC1	17.8	18.6	23.8	25.7	18.2	18.9
TC2	4.4	4.4	5.9	5.9	10.5	10.5
TC3	117.9	120.1	91.4	99.4	121.5	125.3
TCE2h	9.9	9.7	71.4	73.5	10.9	10.8
TCE3h	43.0	42.2	86.4	88.3	47.4	47.3
TCR	26.6	26.2	25.6	26.1	29.4	29.3
TCQ	22.1	22.1	21.9	22.1	23.0	23.0
TCS	7.7	7.7	9.5	9.6	9.6	9.6
Total	249.4	251.0	335.9	350.6	270.5	274.7

6.7 The Effect of a Selective Controller with Low Selective Switch (LSS) to the Energy Management

The CS1 of HDA process alternative 6, the best control structure for handle disturbances is used to study the effect of selective controller with low selective switch (LSS) to energy management. This control structure employs the selective controller with low selective switch (LSS) at FEHE1 (figure 6.49). In order to study this effect, we will compare between control structure with LSS and control structure without LSS. For control structure without LSS, 2 control structures are designed but the major control loop is same except the control loop at FEHE1. First control structure, the outlet hot temperature is controlled at FEHE1 replacing selective controller (figure 6.50). Another control structure, the outlet cold temperature is controlled at FEHE 1 replacing selective controller (figure 6.51).

In order to study this effect, six disturbance loads are used to illustrate the dynamic behavior both the control structure with LSS and the control structure without LSS. The energy consumption is considered for each disturbance load.

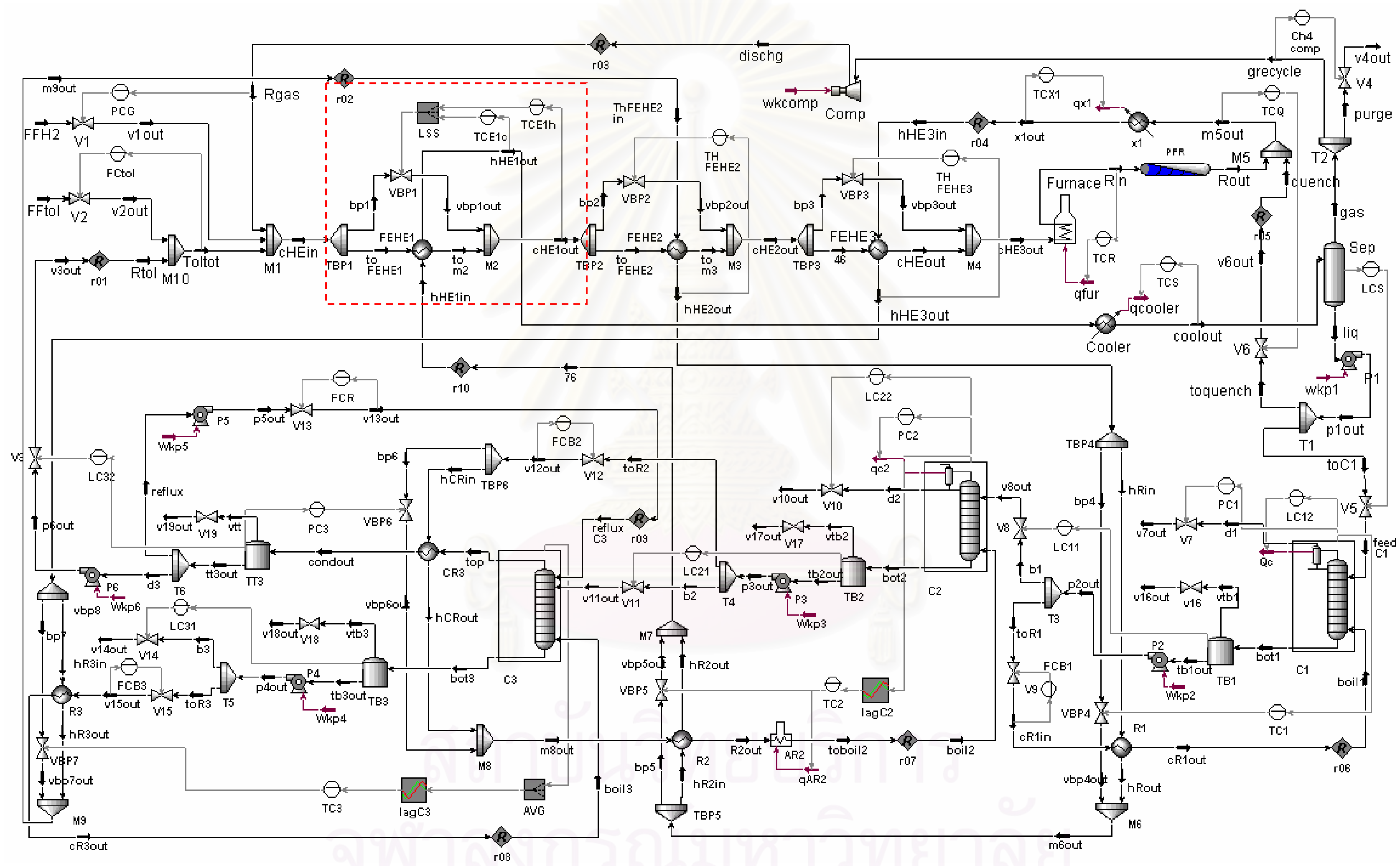

Figure 6.49 Control Structure with Low Selective Switch (LSS) of Selective Controller at FEHE1

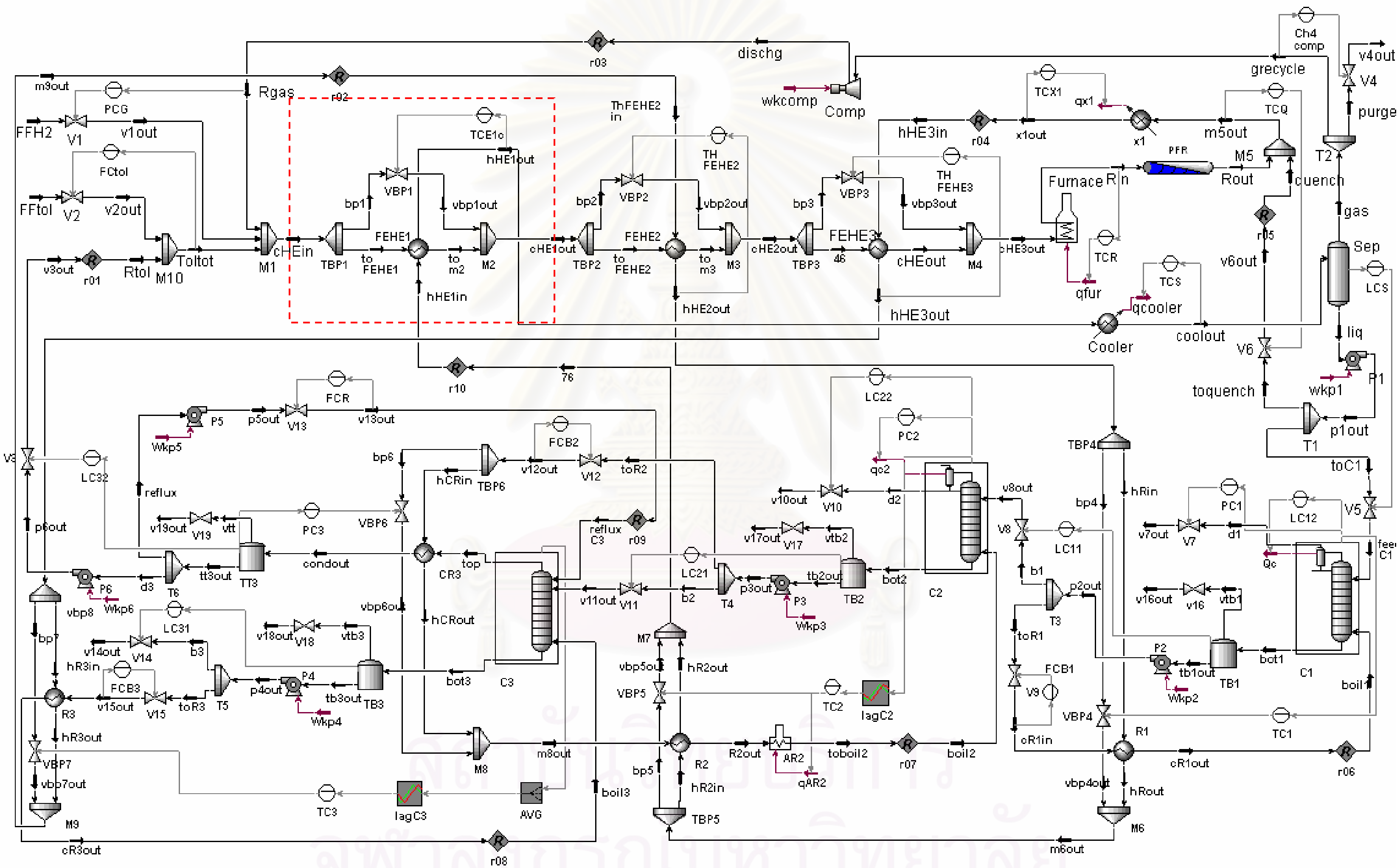

Figure 6.50 Control Structure with outlet hot temperature control at FEHE1

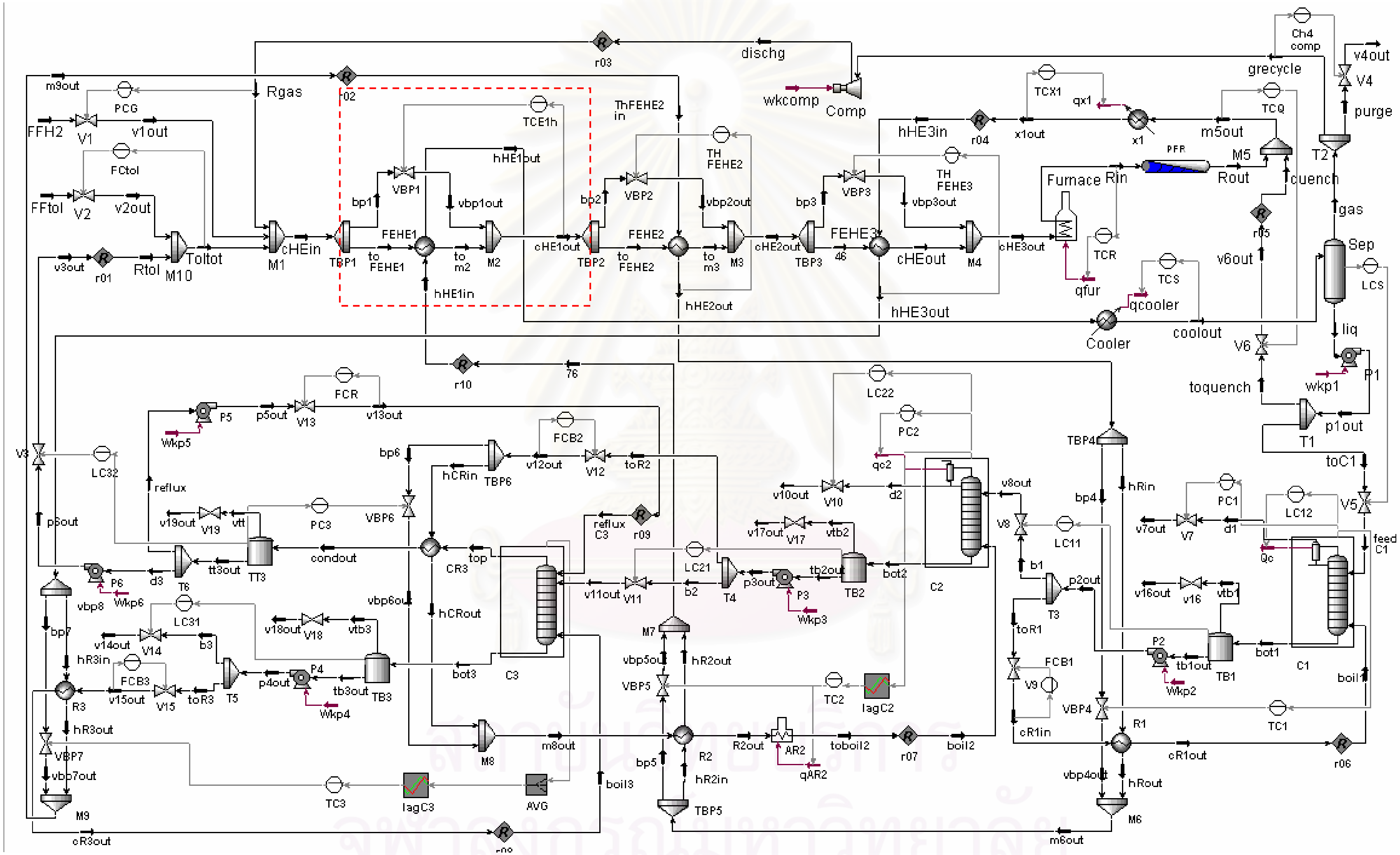

Figure 6.51 Control Structure with outlet cold temperature control at FEHE1

6.7.1 Change in the Heat Load Disturbance of Cold Stream by Decreasing Inlet Temperature of Reactor Feed Stream

In order to make this disturbance load, the fresh toluene feed temperature is decreased from 30 to 20^{o}C as the inlet temperature of C1 stream is decreased (figures 6.52 - 6.54).

Figures 6.52 - 6.54 show heat pathway to shift this disturbance load to cooler or furnace for control structure with LSS at FEHE1, control structure with outlet hot temperature control at FEHE1 and control structure with outlet cold temperature control at FEHE1, respectively.

Table 6.19 the Heat Duty of Utility Units for Change in the Heat load Disturbance of Cold Stream by Decreasing Inlet Temperature of Reactor Feed Stream

Utility Unit	heat duty (kW)			
	Normal operating condition	Decreasing inlet temperature of reactor feed stream		
		CS with LSS	CS with outlet hot temperature control	CS with outlet cold temperature control
Qfurnace	5626	5626	5682	5626
Qcooler	2617	2564	2621	2563
Qcond. C1	166.8	166.8	166.8	166.8
Qcond. C2	4035	4035	4035	4035
QAR2	-	-	-	-
Total	12444.8	12391.8	12504.8	12390.8

As can be seen in table.6.19, control structure with LSS can decrease duty at cooler because the selective control with LSS will select the appropriate heat pathway to shift this disturbance load to cooler in order to achieve dynamic maximum energy recovery (DMER) so total utility duty is less than normal operation about 0.45 %. For control structure with outlet hot temperature control, this disturbance load is shifted to furnace so the duty of furnace increases and total utility duty of this control structure is higher than control structure with LSS. For control structure with outlet cold temperature control, all the thermal disturbance load in C1 is shifted to cooler as the appropriate heat pathway for this disturbance case so the result for this control structure is same as control structure with LSS.

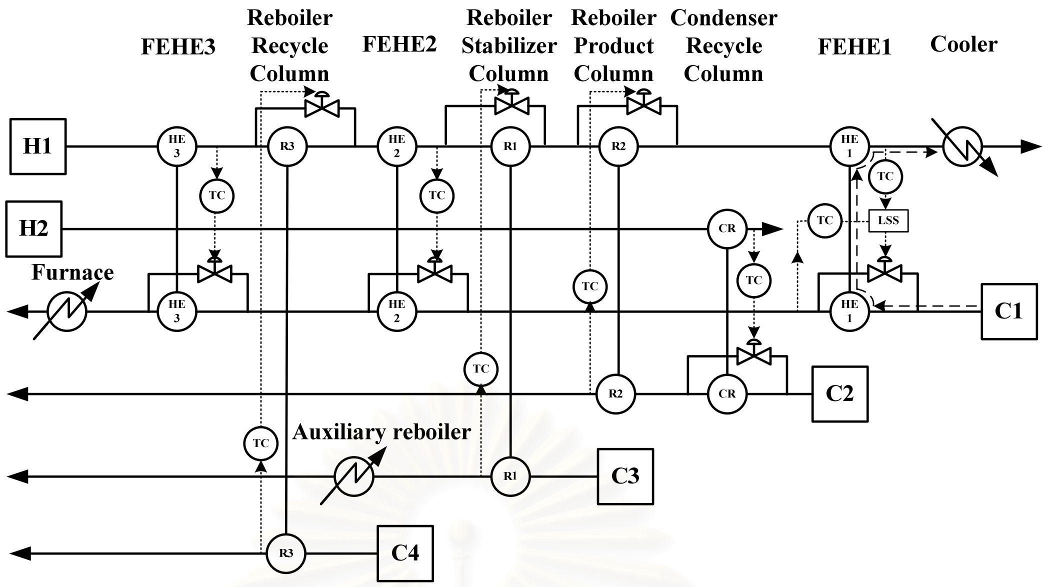

Figure 6.52 the Heat Pathway of Control Structure with LSS at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Decreasing Inlet Temperature of Reactor Feed Stream

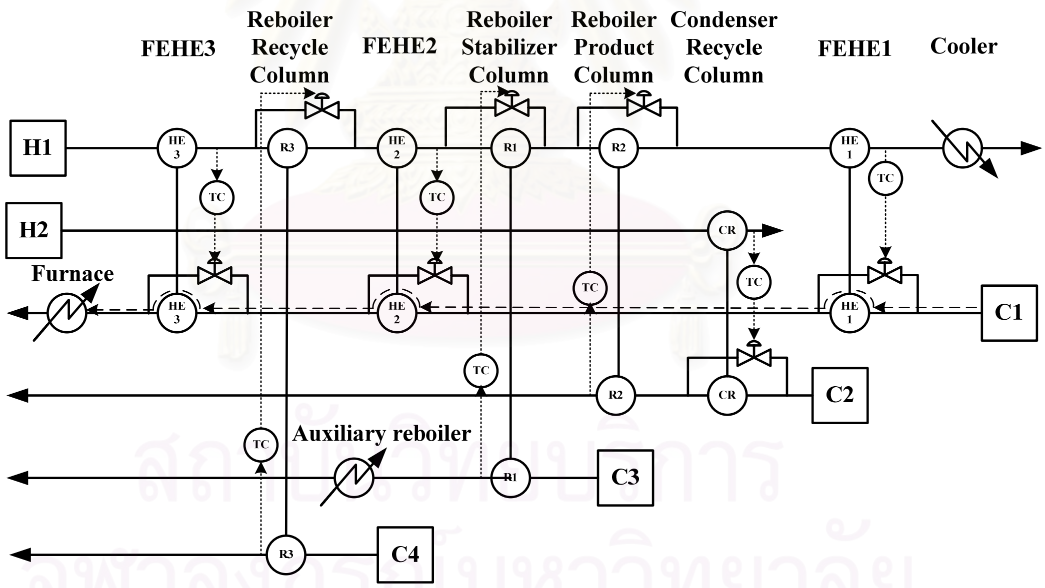

Figure 6.53 the Heat Pathway of Control Structure with Outlet Hot Temperature control at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Decreasing Inlet Temperature of Reactor Feed Stream

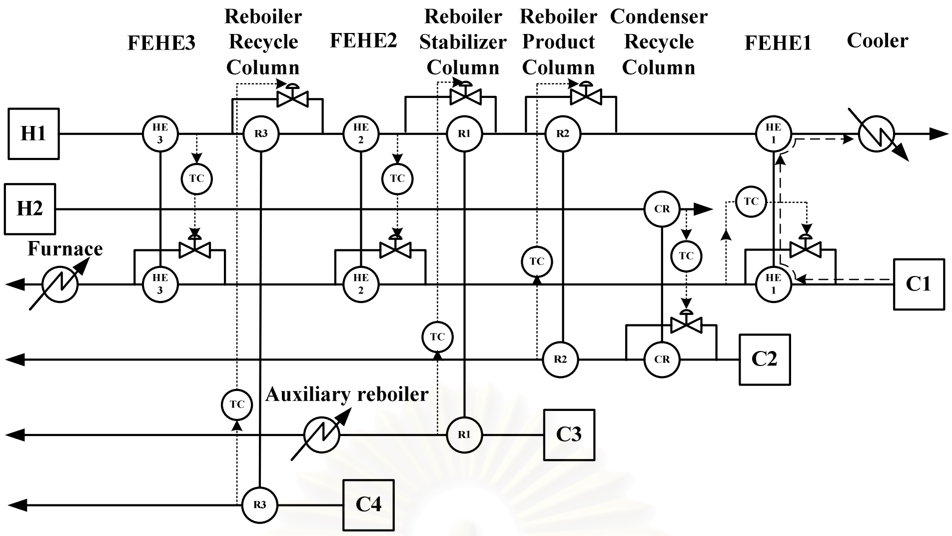

Figure 6.54 the Heat Pathway of Control Structure with Outlet Cold Temperature Control at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Decreasing Inlet Temperature of Reactor Feed Stream

6.7.2 Change in the Heat Load Disturbance of Cold Stream by Increasing Inlet Temperature of Reactor Feed Stream

In order to make this disturbance, the fresh toluene feed temperature is increased from 30 to 40^{o}C as the inlet temperature of C1 stream is increased (figure 6.55 - 6.57).

Figures 6.55 - 6.57 show heat pathway to shift this disturbance load to cooler or furnace for control structure with LSS at FEHE1, control structure with outlet hot temperature control at FEHE1 and control structure with outlet cold temperature control at FEHE1, respectively.

Table 6.20 the Heat Duty of Utility Units for Change in the Heat Load Disturbance of Cold Stream by Increasing Inlet Temperature of Reactor Feed Stream

Utility Unit	heat duty (kW)			
	Normal operating condition	Increasing inlet temperature of reactor feed stream		
		CS with LSS	CS with outlet hot temp. control	CS with outlet cold temp. control
Qfurnace	5626	5570	5570	5627
Qcooler	2617	2614	2614	2671
Qcond. C1	166.8	166.8	166.8	166.8
Qcond. C2	4035	4039	4039	4039
QAR2	-	-	-	-
Total	12444.8	12389.8	12389.8	12503.8

As can be seen in table.6.20, control structure with LSS can decrease duty at cooler because the selective control with LSS will select the appropriate heat pathway to shift this disturbance load to furnace in order to achieve dynamic maximum energy recovery (DMER) and total utility duty is less than normal operation about 0.45 %. For control structure with outlet hot temperature control, all the thermal disturbance load in C1 is shifted to heater as the appropriate heat pathway for this disturbance load so the result for this control structure is same as control structure with LSS. For control structure with outlet cold temperature control, this disturbance load is shifted to cooler so the duty of cooler increases and total utility duty of this control structure is higher than other.

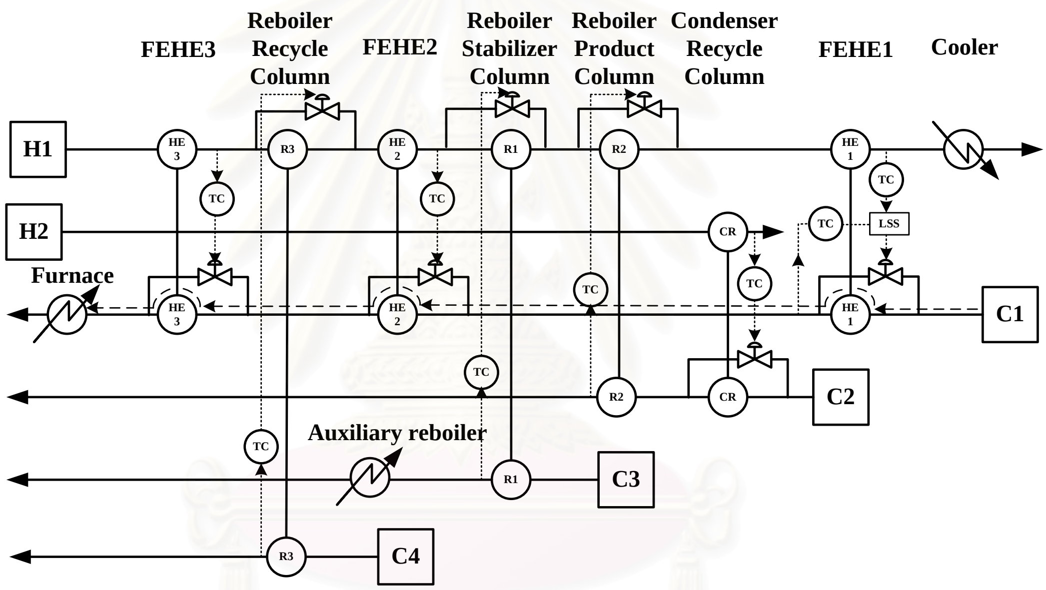

Figure 6.55 the Heat Pathway of Control Structure with LSS at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Increasing Inlet Temperature of Reactor Feed Stream

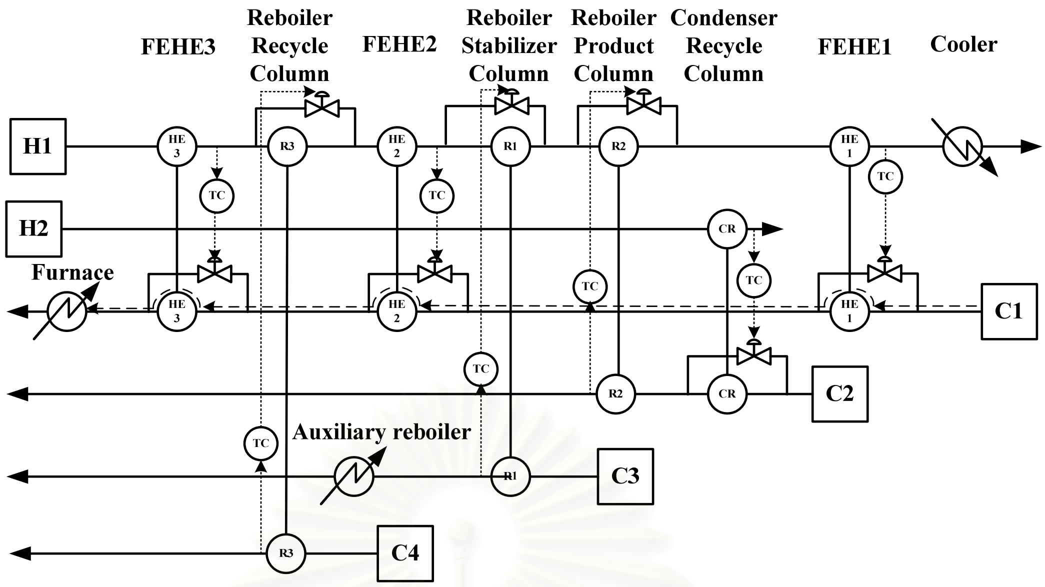

Figure 6.56 the Heat Pathway of Control Structure with Outlet Hot Temperature control at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Increasing Inlet Temperature of Reactor Feed Stream

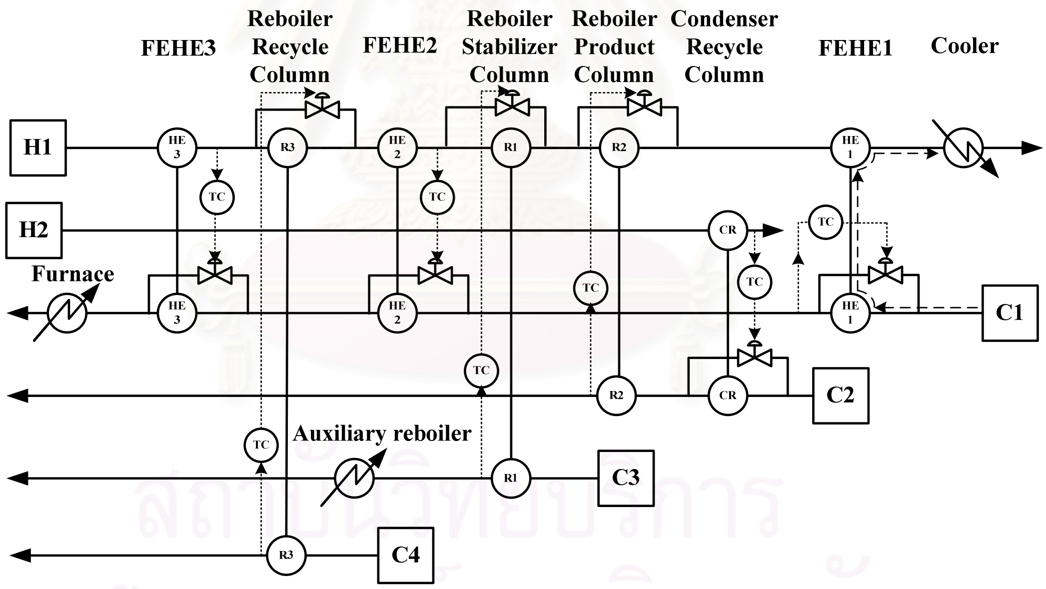

Figure 6.57 the Heat Pathway of Control Structure with Outlet Cold Temperature Control at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Increasing Inlet Temperature of Reactor Feed Stream

6.7.3 Change in the Heat Load Disturbance of Hot Stream by Decreasing Inlet Temperature of Reactor Product Stream

In order to make this disturbance, the reactor product stream temperature is decreased from 621.1 to 611.1^{o}C as the inlet temperature of H1 stream is decreased (figures 6.58 - 6.60).

Figures 6.58 - 6.60 show heat pathway to shift this disturbance load to cooler or furnace for control structure with LSS at FEHE1, control structure with outlet hot temperature control at FEHE1 and control structure with outlet cold temperature control at FEHE1, respectively.

Table 6.21 the Heat Duty of Utility Units for Change in the Heat Load Disturbance of Hot Stream by Decreasing Inlet Temperature of Reactor Product Stream

Utility Unit	heat duty (kW)			
	Normal operating condition	Increasing inlet temperature of reactor product stream		
		CS with LSS	CS with outlet hot temp. control	CS with outlet cold temp. control
Qfurnace	5626	6041	6041	6051
Qcooler	2617	2624	2624	2634
Qcond. C1	166.8	166.7	166.7	166.7
Qcond. C2	4035	4027	4027	4027
QAR2	-	-	-	-
Total	12444.8	12858.7	12858.7	12878.7

All the control structures shift this disturbance load to furnace because there are the same control loops to handle this disturbance load at FEHE3 (figures 6.58 - 6.60). As can be seen in table 6.21, the total duty of utility for control structure with LSS and control structure with outlet hot temperature control are similar and lowest heat duty but the total duty of utility for the control structure with outlet cold temperature control is a little bit higher than other.

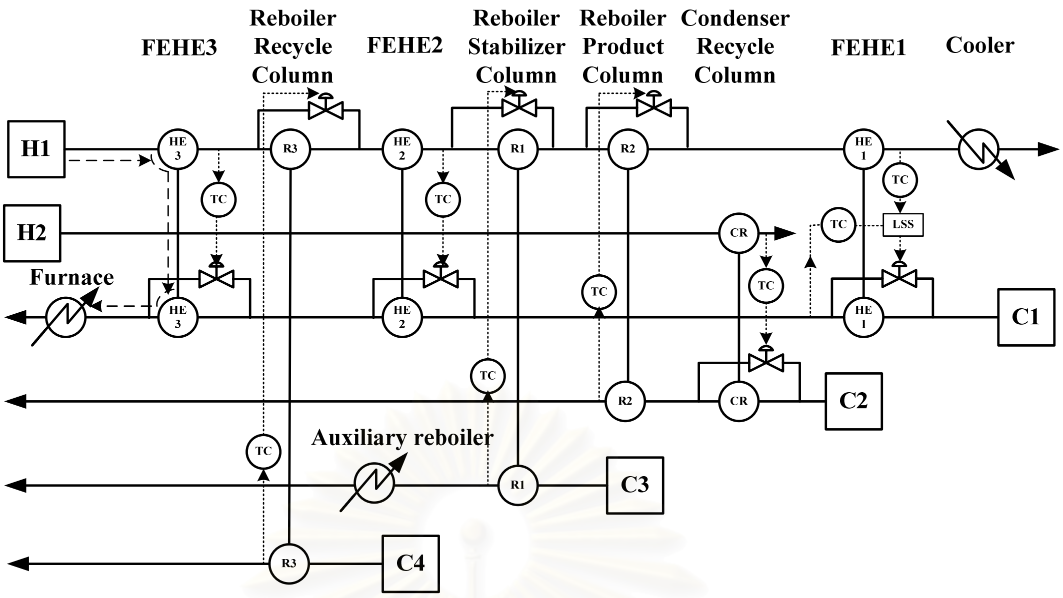

Figure 6.58 the Heat Pathway of Control Structure with LSS at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Decreasing Inlet Temperature of Reactor Product Stream

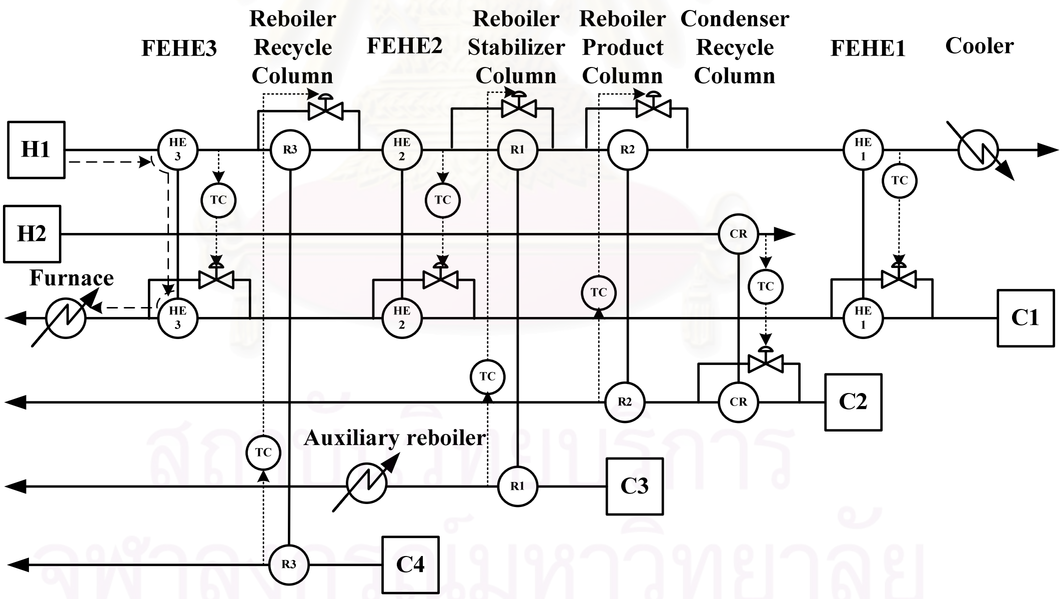

Figure 6.59 The Heat Pathway of Control Structure with outlet hot temperature control at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Decreasing Inlet Temperature of Reactor Product Stream

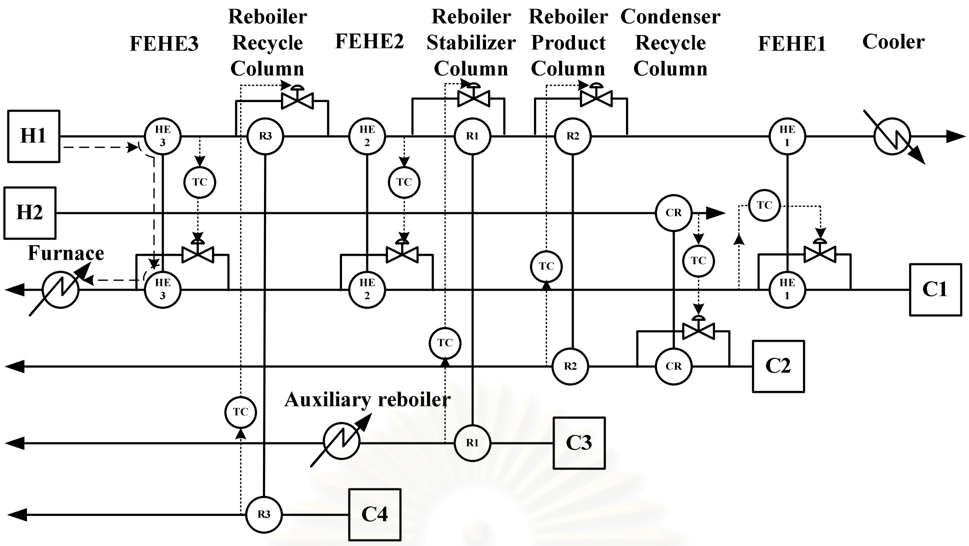

Figure 6.60 The Heat Pathway of Control Structure with Outlet Cold Temperature Control at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Decreasing Inlet Temperature of Reactor Product Stream

6.7.4 Change in the Heat Load Disturbance of Hot Stream by Increasing Inlet Temperature of Reactor Product Stream

In order to make this disturbance, the reactor product stream temperature is increased from 621.1 to 631.1^{o}C as the inlet temperature of H1 stream is increased (figures 6.61 - 6.63).

Figures 6.61 - 6.63 show heat pathway to shift this disturbance load to cooler or furnace for control structure with LSS at FEHE1, control structure with outlet hot temperature control at FEHE1 and control structure with outlet cold temperature control at FEHE1, respectively.

Table 6.22 the Heat Duty of Utility Units for Change in the Heat Load Disturbance of Hot Stream by Increasing Inlet Temperature of Reactor Product Stream

Utility Unit	heat duty (kW)			
	Normal operating condition	Increasing inlet temperature of reactor product stream		
		CS with LSS	CS with outlet hot temp. control	CS with outlet cold temp. control
Qfurnace	5626	5201	5213	5201
Qcooler	2617	2599	2611	2599
Qcond. C1	166.8	166.8	166.8	166.8
Qcond. C2	4035	4044	4043	4044
QAR2	-	-	-	-
Total	12444.8	12010.8	12033.8	12010.8

All the control structures shift this disturbance load to furnace because there are the same control loops to handle this disturbance load. As can be seen in table 6.22, the total duty of utility for control structure with LSS and control structure with outlet cold temperature control are similar and lowest heat duty but the total duty of utility for the control structure with outlet cold temperature control is a little bit higher than other.

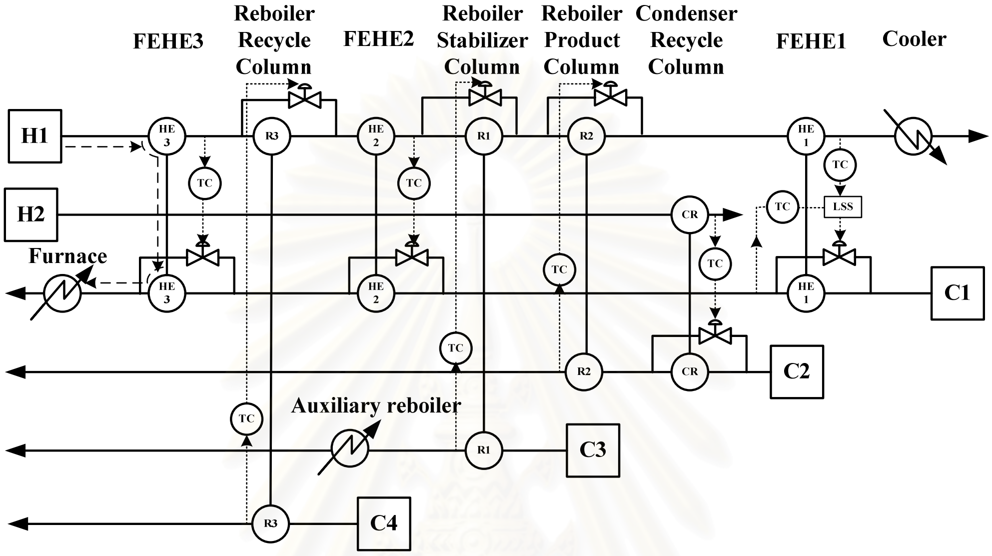

Figure 6.61 the Heat Pathway of Control Structure with LSS at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Increasing Inlet Temperature of Reactor Product Stream

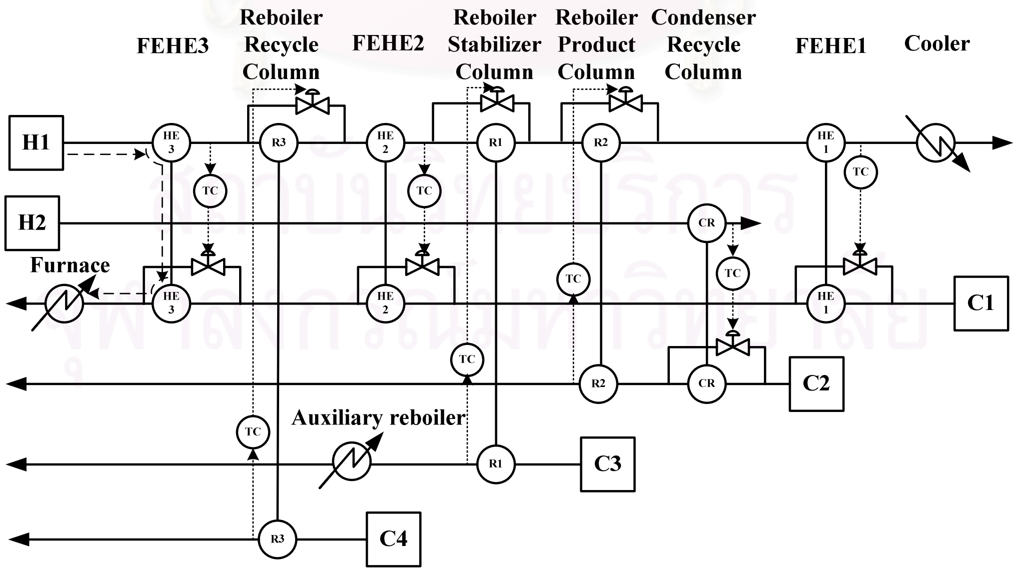

Figure 6.62 the Heat Pathway of Control Structure with outlet hot temperature control at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Increasing Inlet Temperature of Reactor Product Stream

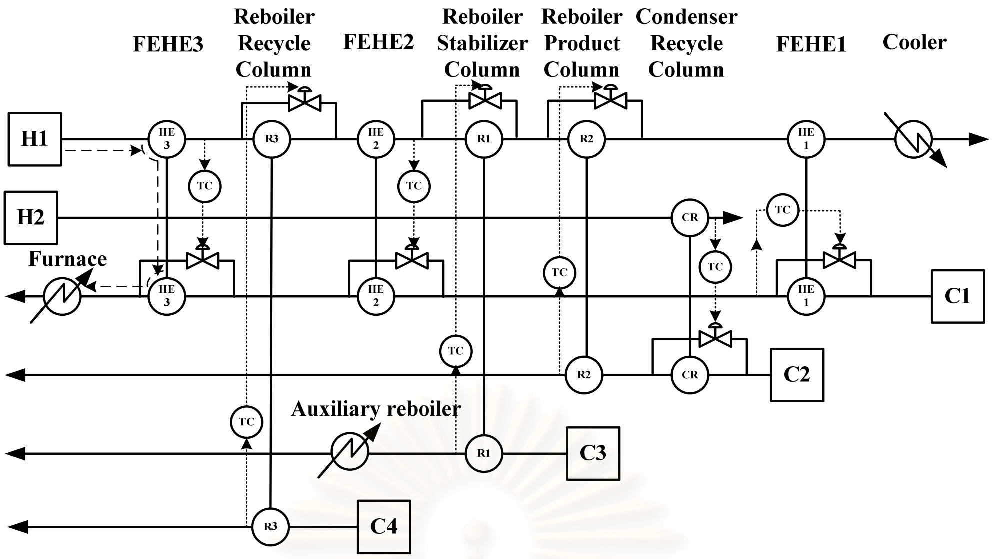

Figure 6.63 the Heat Pathway of Control Structure with Outlet Cold Temperature Control at FEHE1 for Changing in the Heat Load Disturbance of Cold Stream by Increasing Inlet Temperature of Reactor Product Stream

6.7.5 Change in the Total Toluene Feed Flowrates by Decreasing Flowrates

In order to make this disturbance, the total toluene feed flowrates is decreased from 172.3 to 162.3 kgmole/hr.

Table 6.23 shows the heat duty of utility units for change in the total toluene feed flowrates by decreasing flowrates

Table 6.23 the Heat Duty of Utility Units for Change in the Total Toluene Feed Flowrates by Decreasing Flowrates

Utility Unit	heat duty (kW)			
	Normal operating condition	Increasing inlet temperature of reactor product stream		
		CS with LSS	CS with outlet hot temp. control	CS with outlet cold temp. control
Qfurnace	5626	5540	5599	5540
Qcooler	2617	2531	2589	2531
Qcond. C1	166.8	182.2	182.1	182.2
Qcond. C2	4035	3835	3831	3835
QAR2	-	-	-	-
Total	12444.8	12088.2	12201.1	12088.2

As can be seen in table 6.22, the total duty of utility for control structure with LSS is lowest heat duty compare with other because the selective control with LSS will select the appropriate the heat pathway to achieve dynamic maximum energy recovery (DMER).

6.7.6 Change in the Total Toluene Feed Flowrates by Increasing Flowrates

In order to make this disturbance, the total toluene feed flowrates is increased from 172.3 to 182.3 kgmole/hr.

Table 6.23 shows the heat duty of utility units for change in the total toluene feed flowrates by increasing flowrates

Table 6.24 the Heat Duty of Utility Units for Change in the Total Toluene Feed Flowrates by Increasing Flowrates

Utility Unit	heat duty (kW)			
	Normal operating condition	Increasing inlet temperature of reactor product stream		
		CS with LSS	CS with outlet hot temp. control	CS with outlet cold temp. control
Qfurnace	5626	5669	5669	5707
Qcooler	2617	2646	2646	2684
Qcond. C1	166.8	159.3	159.3	159.2
Qcond. C2	4035	4258	4258	4255
QAR2	-	-	-	-
Total	12444.8	12732.3	12732.3	12805.2

Again, as can be seen in table 6.23, the total duty of utility for control structure with LSS is lowest heat duty compare with other because the selective control with LSS will select the appropriate the heat pathway to achieve dynamic maximum energy recovery (DMER).

For the all disturbance load testing, the control structure with LSS in selective controller is the best control structure for energy management because the dynamic maximum energy recovery can be achieved for all disturbance load testing.

CHAPTER VII

CONCLUSIONS AND RECOMMENDATIONS

7.1. Conclusion

The plantwide process control problem is to develop a control strategy for a complex and integrated process that satisfies the plant's design objectives. The cost of process is one of the important plant's design objectives therefore the design of the highly heat-integrated HDA process alternatives 5 and 6 with minimum auxiliary utility unit are developed in this research. The commercial software HYSYS was utilized to carry out both the steady state and dynamic simulations.

At steady state design, the design procedure for evaluating the minimum number of auxiliary utility unit is proposed in this researcher as follows:

1. Determined the expected disturbances and their magnitudes

2. Design the worst case conditions

   The two worst case conditions are selected to be the design condition for evaluating the minimum number of auxiliary utility unit. The first worst case condition is the condition that hot process streams have minimum heat supply and the cold streams have maximum heat demands (minimum heating condition). This condition will be used to evaluate the minimum number of auxiliary heating unit. The second case condition is the condition that cold process streams have maximum heat supply and the cold streams have minimum heat demand (maximum cooling condition). This condition will be used to evaluate the minimum number of auxiliary cooling unit.

3. Design heat pathways for the worst case conditions

   The objective of design heat pathways is to achieve dynamic maximum energy recover (DMER) therefore the disturbance loads are shifted to cooler for minimum heating condition and to heater for maximum cooling condition.

4. Estimated the minimum auxiliary utility units

The worst condition and the appropriate heat pathway are specified in the heat exchanger network at steady state design in order to find the minimum number of auxiliary utility units that we need to employ for the heat exchanger network. The minimum numbers of auxiliary heating units are evaluated form testing minimum heating condition and the minimum numbers of auxiliary cooling units are evaluated form testing maximum cooling condition.

This design procedure is applied to the highly heat integrated HDA process alternatives 5 and 6. From the steady state simulation result, our design guarantees that the HDA process with highly heat integrated alternatives 5 and 6 are workable by adding only one auxiliary utility unit instead of three and four as suggested by Luyben (1999), respectively.

At dynamic simulation, this work presents new three plantwide designed control structures. The major control loop of three new designs for energy management is same. The LSS is employed at the first feed effluent heat exchanger (FEHE1) in odder to shift the thermal disturbance load in appropriate heat pathway to cooler or heater and DMER can be achieved. The hot outlet temperature of FEHE that will be hot inlet temperature of reboiler (R1 or R3) is controlled in odder to prevent the propagation of thermal disturbance to separation section and to guarantee the workable highly HDA process with minimum auxiliary utility units. For the optimum operation of auxiliary utility units, the split range control is employed to minimize heat load of auxiliary utility units.

From the result, we can see that the highly heat integrated HDA process (Alternatives 5 and 6) with auxiliary utility units suggested by Luyben and with minimum auxiliary utility units are given the similar dynamic performance for all control structures.

Since the major of control loop is similar, the dynamic response and dynamic performance of the three new control structures are slightly deference. As can be seen, CS1 is the best control structure for handle disturbances due to it gives better control performances.

Moreover, the effect of a selective controller with low selective switch (LSS) to the energy management is studied in this research. For the all disturbance load testing, the control structure with LSS in selective controller is the best control structure for energy management because the dynamic maximum energy recovery can be achieved for all disturbance load testing.

7.2 Recommendation

In this work, the dynamic maximum energy recover (DMER) is considered to design the workable highly heat integrated process with minimum auxiliary. If DMER is not considered. We may design the highly heat integrated process without auxiliary utility unit by modifying the control structure for example in figure 7.1 however there will need more heating and cooling requirement at cooler and heater. This control system will work as follows: If the target temperature of cold stream C2 can't achieve, the bypass of heat exchanger E1 will decrease to increase heat transfer at E2. If the bypass of heat exchanger E1 is full close but the target temperature of cold stream C2 still can't achieve. The bypass of heat exchanger E1 will open to decrease heat transfer at E1 in order to increase heat transfer at E2.

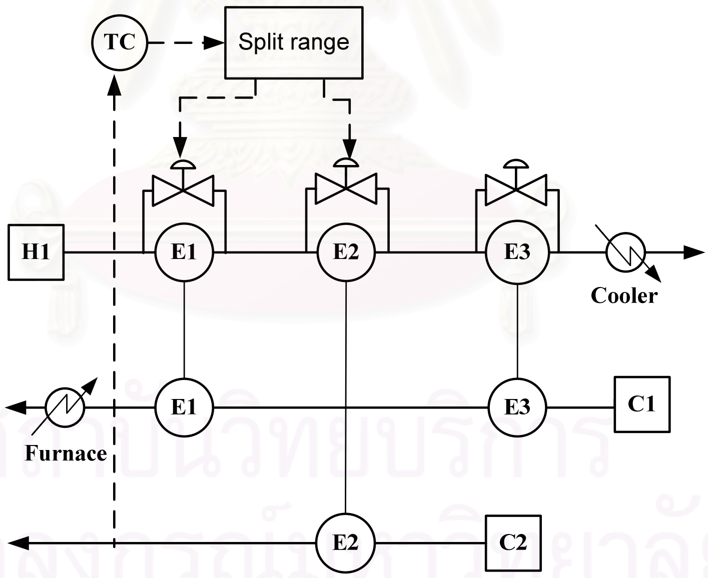

Figure 7.1 The example modifying control structure for the process with out auxiliary utility unit

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APPENDICES

APPENDIX A

PROCESS STREAM DATA OF HDA PROCESS

Table A.1 Process Stream Data of HDA Process Alternative 5 at Steady State simulation

	Fresh toluene	Fresh hydrogen	Purge gas	Stabilizer gas	Benzene product	Diphenyl product
Stream name	FFtol	FFH2	purge	d1	d2	b3
Temperature [C]	30.00	30.00	45.00	51.50	105.56	349.16
Pressure [kPa]	4309.22	4309.22	3130.22	1034.00	206.84	545.32
MolarFlow[kgmole/hr]	132.66	222.43	217.73	8.71	123.80	2.87
H_2, mole fraction	0	0.97	0.3966	0.0884	0	0
CH_4	0	0.03	0.5919	0.8690	0	0
C_6H_6	0	0	0.0104	0.0420	0.9997	0
C_7H_8	1	0	0.0012	0.0006	0.0003	0.00026
$C_{12}H_{10}$	0	0	0	0	0	0.99974

Table A.1 Continued

	Gas recycle	Toluene recycle	FEHE1 Cold in	FEHE1 Cold out	FEHE1 Cold out	Reactor inlet
Stream name	dischg	d3	cHE1in	cHE1out	cHE2out	Rin
Temperature [C]	70.08	180.13	66.42	141.00	513.10	621.11
Pressure [kPa]	3964.49	520.00	3964.49	3861.07	3757.64	3468.06
MolarFlow[kgmole/hr]	1606.73	39.59	2001.41	2001.41	2001.41	2001.41
H_2, mole fraction	0.3966	0	0.4262	0.4262	0.4262	0.4262
CH_4	0.5919	0	0.4785	0.4785	0.4785	0.4785
C_6H_6	0.0104	0.0006	0.0083	0.0083	0.0083	0.0083
C_7H_8	0.0012	0.9993	0.0870	0.0870	0.0870	0.0870
$C_{12}H_{10}$	0	0	0	0	0	0

Table A.1 Continued

	Reactor outlet	Quench	FEHE2 Hot in	FEHE2 Hot out	FEHE1 Hot in	FEHE1 Hot out
Stream name	Rout	quench	hHE2in	hHE2out	hR2out	hHE1out
Temperature [C]	665.91	45.40	621.16	299.95	191.32	106.91
Pressure [kPa]	3350.85	3350.85	3350.85	3299.14	3195.72	3144.01
MolarFlow[kgmole/hr]	2001.41	49.72	2049.34	2049.34	2049.34	2049.34
H_2, mole fraction	0.3616	0.0044	0.3535	0.3535	0.3535	0.3535
CH_4	0.5445	0.0432	0.5317	0.5317	0.5317	0.5317
C_6H_6	0.0714	0.7098	0.0871	0.0871	0.0871	0.0871
C_7H_8	0.0210	0.2261	0.0259	0.0259	0.0259	0.0259
$C_{12}H_{10}$	0.0014	0.0165	0.0018	0.0018	0.0018	0.0018

Table A.1 Continued

	Reb1 cold in	Reb1 cold out	Reb1 hot in	Con/Reb cold in	Reb2 cold in	Reb2 cold out
Stream name	cR1in	cR1out	hR1 in	cCRin	cCRout	cR2out
Temperature [C]	190.01	203.00	257.72	30.00	30.00	45
Pressure [kPa]	1244.90	1038.05	3247.43	4378.17	4378.17	3211.58
MolarFlow[kgmole/hr]	182.67	182.67	1946.87	130.00	222.72	219.62
$H_{2,}$mole fraction	0	0	0.3535	0	0.97	0.3980
CH_4	0	0	0.5317	0	0.03	0.5906
C_6H_6	0.7446	0.7446	0.0871	0	0	0.0102
C_7H_8	0.2381	0.2381	0.0259	1	0	0.0012
$C_{12}H_{10}$	0.0173	0.0173	0.0018	0	0	0

Table A.1 Continued

	Con/Reb hot in	Separator inlet	Separator Gas out	Stabilizer feed	Stabilizer bottoms	Product bottoms
Stream name	top	coolout	gas	toC1	b1	b2
Temperature [C]	51.05	45.00	45.00	45.26	190.05	143.92
Pressure [kPa]	1034.21	3130.22	3130.22	3654.22	1451.74	689.48
MolarFlow[kgmole/hr]	8.75	2049.34	1824.45	175.17	166.25	42.45
$H_{2,}$mole fraction	0.0890	0.3535	0.3966	0.0044	0	0
CH_4	0.8686	0.5317	0.5919	0.0432	0	0
C_6H_6	0.0420	0.0871	0.0104	0.7097	0.7446	0.0006
C_7H_8	0.0003	0.0259	0.0012	0.2262	0.2381	0.9316
$C_{12}H_{10}$	0	0.0018	0	0.0165	0.0173	0.0678

Table A.1 Continued

	Column reflux			Column boilup		
	C1	C2	C3	C1	C2	C3
Temperature [C]	51.50	105.56	180.26	203.00	165.14	165.14
Pressure [kPa]	1034.00	206.84	540.00	1038.05	222.11	222.11
MolarFlow[kgmole/hr]	14.82	375.71	131.99	182.46	389.82	389.82
$H_{2,}$mole fraction	0.0003	0	0	0	0	0
CH_4	0.0214	0	0	0	0	0
C_6H_6	0.9421	0.9997	0.0006	0.7445	0.0006	0.0006
C_7H_8	0.0362	0.0003	0.9993	0.2382	0.9316	0.9316
$C_{12}H_{10}$	0	0	0	0.0173	0.0678	0.0678

Table A.2 Process stream data of HDA process alternative 6 at Steady State simulation

	Fresh toluene	Fresh hydrogen	Purge gas	Stabilizer gas	Benzene product	Diphenyl product
Stream name	FFtol	FFH2	purge	d1	d2	b3
Temperature [C]	30.00	30.00	45.00	51.50	105.56	349.22
Pressure [kPa]	4309.22	4309.22	3130.22	1034.00	206.84	798.20
MolarFlow[kgmole/hr]	132.66	222.43	217.73	8.71	123.80	2.88
$H_{2,}$mole fraction	0	0.97	0.3966	0.0884	0	0
CH_4	0	0.0300	0.5919	0.8690	0	0
C_6H_6	0	0	0.0104	0.0420	0.9997	0
C_7H_8	132.6630	0	0.0012	0.0006	0.0003	0.0001
$C_{12}H_{10}$	0	0	0	0	0	0.9999

Table A.2 Continued

	Gas recycle	Toluene recycle	FEHE1 Cold in	FEHE1 Cold out	FEHE2 Cold out	FEHE3 Cold out
Stream name	dischg	d3	cHE1in	cHE1out	cHE2out	cHE3out
Temperature [C]	70.08	180.13	66.42	145.20	246.20	470.00
Pressure [kPa]	3964.49	520.00	3964.49	3895.54	3826.59	3757.64
MolarFlow[kgmole/hr]	1606.73	39.59	2001.41	2001.41	2001.41	2001.41
mole fraction $H_{2,}$	0.3966	0	0.4262	0.4262	0.4262	0.4262
CH_4	0.5919	0	0.4785	0.4785	0.4785	0.4785
C_6H_6	0.0104	0.0006	0.0083	0.0083	0.0083	0.0083
C_7H_8	0.0012	0.9993	0.0870	0.0870	0.0870	0.0870
$C_{12}H_{10}$	0	0.0001	0	0	0	0

Table A.2 Continued

	Reactor inlet	Reactor outlet	Quench	FEHE3 hot in	FEHE3 hot out	Reb3 hot out
Stream name	Rin	Rout	quench	hHE3in	hHE3out	hR3out
Temperature [C]	621.11	665.91	45.40	621.16	435.70	390.31
Pressure [kPa]	3468.06	3350.85	3350.85	3350.85	3316.38	3281.91
MolarFlow[kgmole/hr]	2001.41	2001.41	49.72	2049.34	2049.34	2049.34
$H_{2,}$mole fraction	0.4262	0.3616	0.0044	0.3535	0.3535	0.3535
CH_4	0.4785	0.5445	0.0432	0.5317	0.5317	0.5317
C_6H_6	0.0083	0.0714	0.7098	0.0871	0.0871	0.0871
C_7H_8	0.0870	0.0210	0.2261	0.0259	0.0259	0.0259
$C_{12}H_{10}$	0	0.0014	0.0165	0.0018	0.0018	0.0018

Table A.2 Continued

	FEHE2 hot out	Reb1 hot out	Reb2 hot out	FEHE1 hot out	Separator inlet	Separator gas out
Stream name	hHE2out	hR1out	hR2out	hHE1out	cool out	gas
Temperature [C]	304.25	262.22	195.37	106.67	45.00	45.00
Pressure [kPa]	3247.43	3212.96	3178.48	3144.01	3130.22	3130.22
MolarFlow[kgmole/hr]	2049.34	1946.87	1946.87	2049.34	2049.34	1824.45
$H_{2,}$mole fraction	0.3535	0.3535	0.3535	0.3535	0.3535	0.3966
CH_4	0.5317	0.5317	0.5317	0.5317	0.5317	0.5919
C_6H_6	0.0871	0.0871	0.0871	0.0871	0.0871	0.0104
C_7H_8	0.0259	0.0259	0.0259	0.0259	0.0259	0.0012
$C_{12}H_{10}$	0.0018	0.0018	0.0018	0.0018	0.0018	0

Table A.2 Continued

	Stabilizer feed	Stabilizer bottoms	Product bottoms	Reb1 cold out	Reb1 cold out	Con/Reb cold in
Stream name	toC1	b1	b2	cR1in	cR1out	cCRin
Temperature [C]	45.26	190.05	143.92	190.01	203.00	143.96
Pressure [kPa]	3654.22	1451.74	689.48	1244.90	1038.05	551.58
MolarFlow[kgmole/hr]	175.17	166.25	42.45	182.67	182.67	370.33
$H_{2,}$mole fraction	0.0044	0	0	0	0	0
CH_4	0.0432	0	0	0	0	0
C_6H_6	0.7097	0.7446	0.0006	0.7446	0.7446	0.0006
C_7H_8	0.2262	0.2381	0.9316	0.2381	0.2381	0.9316
$C_{12}H_{10}$	0.0165	0.0173	0.0678	0.0173	0.0173	0.0678

Table A.2 Continued

	Con/Reb cold out	Reb2 cold out	Reb3 cold in	Reb3 cold out	Con/Reb hot in	Con/Reb hot out
Stream name	cCRout	cR2out	cR3in	cR3out	top	condout
Temperature [C]	165.13	165.14	349.22	353.00	182.18	180.13
Pressure [kPa]	344.74	222.11	660.30	544.21	540.00	520.00
MolarFlow[kgmole/hr]	370.33	389.82	135.66	135.66	181.58	181.58
$H_{2,}$ mole fraction	0	0	0	0	0	0
CH_4	0	0	0	0	0	0
C_6H_6	0.0006	0.0006	0.0000	0.0000	0.0006	0.0006
C_7H_8	0.9316	0.9316	0.0001	0.0001	0.9993	0.9993
$C_{12}H_{10}$	0.0678	0.0678	0.9999	0.9999	0.0001	0.0001

Table A.2 Continued

	Column reflux			Column boilup		
	C1	C2	C3	C1	C2	C3
Temperature [C]	51.50	105.56	180.26	203.00	165.14	353.00
Pressure [kPa]	1034.00	206.84	540.00	1038.05	222.11	544.21
MolarFlow[kgmole/hr]	14.82	375.71	142.00	182.46	389.82	135.66
mole fraction H_2,	0.0003	0	0	0	0	0
CH_4	0.0214	0	0	0	0	0
C_6H_6	0.9421	0.9997	0.0006	0.7445	0.0006	0
C_7H_8	0.0362	0.0003	0.9993	0.2382	0.9316	0.0002
$C_{12}H_{10}$	0	0	0.0001	0.0173	0.0678	0.9998

APPENDIX B

EQUIPMENT AND DATA SPECIFICATION OF HDA PROCESS

Table B.1 Equipment data and Specifications of HDA Process

Equipments	Specifications	Alternative	
		5	6
Reactor	Diameter (m)	17.374	17.374
	Length (m)	2.905	2.905
	Number of tube	1	1
Furnace	Tube volume (m^3)	8.5	8.5
Cooler	Tube volume (m^3)	8.5	8.5
Separator	Liquid volume (m^3)	1.13	1.13
FEHE1*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	3.718×10^5	4.573×10^5
	Number of shells in series	10	10
	Tube passes per shell	2	2
FEHE2*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	3.898×10^5	9.884×10^4
	Number of shells in series	10	10
	Tube passes per shell	2	2
FEHE3*	Shell volume (m^3)	-	14.16
	Tube volume (m^3)	-	14.16
	UA (kJ/C-h)	-	2.222×10^5
	Number of shells in series	-	10
	Tube passes per shell	-	2
Reboiler1 (R1)*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	6.980×10^5	8.844×10^4
	Number of shells in series	10	10
	Tube passes per shell	2	2
Reboiler2 (R2)*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	1.517×10^5	1.721×10^5
	Number of shells in series	10	10
	Tube passes per shell	2	2

Table B.1 Continued

Equipments	Specifications	Alternative	Equipments
		5	6
Reboiler3 (R3)*.	Shell volume (m^3)	-	14.16
	Tube volume (m^3)	-	14.16
	UA (kJ/C-h)	-	1.818x 10^5
	Number of shells in series	-	10
	Tube passes per shell	-	2
Condensor/Reboiler (CR)*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	4.11x 10^5	2.919x 10^5
	Number of shells in series	10	10
	Tube passes per shell	2	2
Tank Bottom C1(TB1)**	Vesel volume (m^3)	7.08	7.08
Tank Bottom C2(TB2)**	Vesel volume (m^3)	8.50	8.50
Tank Bottom C3(TB3)**	Vesel volume (m^3)	-	1.42
Tank Top C3 (TT3)***	Vesel volume (m^3)	2.83	2.83

* Simulation by heat exchanger

** Simulation by drum to accumulate liquid from the bottom of the three columns

*** Simulation by drum to accumulate condensate from the top of column C3

Table B.2 Column Specifications of HDA process alternative 5

Stream name	Stabilizer Column	Product Column	Recycle Column
Model	Refluxed Absorber	Refluxed Absorber	Reboiled Absorber
Number of theoretical tray	6	27	7
Feed tray	3	15	5
Pressure (kPa)	1034.25	206.85	540.00
Diameter (m)	1.067	1.981	0.762
Weir length (m)	0.8842	1.405	0.5563
Weir height (m)	0.0508	0.0508	0.0508
Tray spacing (m)	0.6096	0.6096	0.6096
Tray type	Sieve	Sieve	Sieve
Reboiler vol. (m3)	-	-	1.41
Condenser vol. (m3)	0.28	9.06	-
Specification 1	Benzene mole fraction in overhead = 0.042	Toluene mole fraction in overhead = 0.0003	-
Specification 2	-	-	Toluene mole fraction in overhead = 0.00026

Table B.3 Column Specifications of HDA process alternative 6

Stream name	Stabilizer Column	Product Column	Recycle Column
Model	Refluxed Absorber	Refluxed Absorber	Absorber
Number of theoretical tray	6	27	7
Feed tray	3	15	5
Pressure (kPa)	1034.25	206.85	540.00
Diameter (m)	1.067	1.981	0.6096
Weir length (m)	0.8842	1.405	0.4544
Weir height (m)	0.0508	0.0508	0.0508
Tray spacing (m)	0.6096	0.6096	0.6096
Tray type	Sieve	Sieve	Sieve
Reboiler vol. (m3)	-	-	-
Condenser vol. (m3)	0.28	9.06	-
Specification 1	Benzene mole fraction in overhead = 0.042	Toluene mole fraction in overhead = 0.0003	-
Specification 2	-	-	-

APPENDIX C

TUNING OF CONTROL STRUCTURES

C.1 Tuning Controllers

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

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

C.2 Tuning Flow, Level and Pressure Loops

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

Most level controllers should use proportional-only action with a gain of 1 to 2. This provides the maximum amount of flow smoothing. Proportional control means there will be steady state offset (the level will not be returned to its setpoint value). However, maintaining a liquid level at a certain value is often not necessary when the liquid capacity is simply being used as surge volume. So the recommended tuning of a level controller is K_C = 2. Most pressure controllers can be fairly easily tuned. The process time constant is estimated by dividing the gas volume of the system by the volumetric flowrate of gas flowing through the system. Setting the integral time equal to about 2 to 4 times the process time constant and using a reasonable controller gain usually gives satisfactory pressure control. Typical pressure controller tuning constants for columns and tanks are K_C = 2 and τ_I = 10 minutes.

C.3 Relay- Feedback Testing

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

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

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

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

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

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

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

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

$$K_C = K_U / 2.2 \tag{2}$$

$$\tau_I = P_U / 1.2 \tag{3}$$

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

$$K_C = K_U / 3.2 \tag{4}$$

$$\tau_I = 2.2 P_U \tag{5}$$

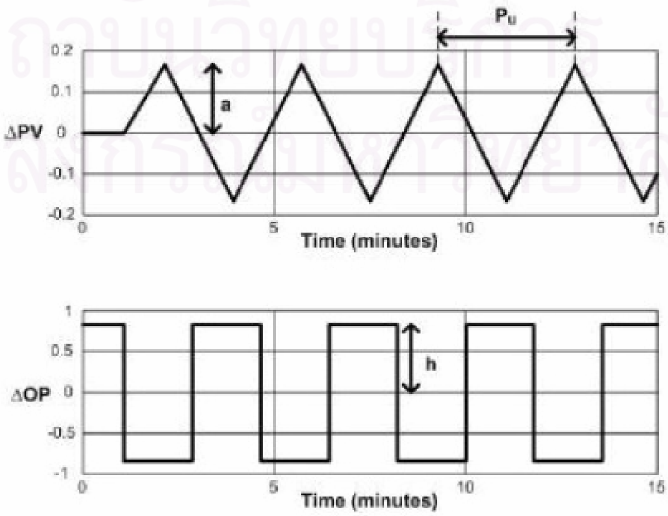

Figure C.1 Input and Output from Relay-Feedback Test

C.4 Inclusion of Lags

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

Table C.1 Typical measurement lags

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

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

Realistic dynamic simulations require that we explicitly include lags and/or dead times in all the important loops. Usually this means controllers that affect Product quality or process constraint. Table B.1 summarizes some recommended lags to include in several different types of control loops.

VITA

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