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DESIGN OF HEAT-INTEGRATED PROCESS
STRUCTURES FOR HDA PLANT

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A Thesis Submitted in Partial Fulfillment of the Requirements
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กสิณ พรพิทักษ์ธรรม: การออกแบบโครงสร้างกระบวนการที่มีการเบ็ดเสร็จทางด้านความร้อนสำหรับโรงงานไฮโดรดีอัลคิลเลชัน. (DESIGN OF HEAT-INTEGRATED PROCESS STRUCTURES FOR HDA PLANT) อ.ที่ปรึกษาวิทยานิพนธ์หลัก: ผศ. ดร. มนตรี วงศ์ศรี, 225 หน้า.

ในกระบวนการทางเคมีสิ่งที่สำคัญที่สุด คือการควบคุมกระบวนการเพื่อให้ระบบดำเนินงานไปด้วยความปลอดภัยและได้ผลิตภัณฑ์ตามเป้าหมายที่ต้องการ ซึ่งในกระบวนการทางอุตสาหกรรมเคมีส่วนใหญ่จะแบ่งหน่วยการผลิตออกเป็นสองส่วนใหญ่ๆ ด้วยกัน คือ ส่วนของการเกิดปฏิกิริยา และส่วนของการแยก ทั้งสองส่วนต้องการระบบควบคุมที่ทำให้กระบวนการดำเนินการได้อย่างเหมาะสมทางเศรษฐศาสตร์ ในงานวิจัยนี้ได้ทำการปรับปรุงโครงสร้างในส่วนของการแยกของกระบวนการไฮโดรดิอัลคิเลชันและนำข่ายงานเครื่องแลกเปลี่ยนความร้อนมาใช้เพื่อช่วยในการประหยัดพลังงาน นั้นคือวัตถุประสงค์ของงานวิจัยเรา โดยกระบวนการไฮโดรดิอัลคิเลชันแบบเดิมนั้นมีการกลั่นแบบเรียงลำดับซึ่งจะถูกแทนที่ด้วยการออกแบบใหม่ จากนั้นได้ออกแบบข่ายงานเครื่องแลกเปลี่ยนความร้อนใหม่อีก 4 แบบ การออกแบบโครงสร้างการควบคุมแบบแพลนท์ไวด์สำหรับกระบวนการที่มีการเบ็ดเสร็จพลังงานถูกออกแบบโดยการใช้วิธีการส่งผ่านความแปรปรวนของ (Wongsri, M., 1990) และการออกแบบเส้นทางเดินของความร้อน (heat pathway heuristics) ของ (Wongsri, M. and Hermawan Y.D., 2005) ตามลำดับ ในการประเมินสมรรถนะโครงสร้างการควบคุมได้ใช้ตัวรบกวนกระบวนการ 2 ชนิด ได้แก่ การรบกวนทางความร้อนและการรบกวนอัตราการไหลของสาร การประเมินสมรรถนะกระบวนการที่มีการเบ็ดเสร็จทางด้านพลังงานและโครงสร้างการควบคุมที่สภาวะเชิงพลวัต ซึ่งสามารถทำได้ด้วยการจำลองกระบวนการด้วยโปรแกรมไฮซีส

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KASIN PRONPITAKTHUM: DESIGN OF HEAT-INTEGRATED PROCESS STRUCTURES FOR HDA PLANT. THESIS PRINCIPAL ADVISOR: ASST. PROF. MONTREE WONGSRI, D.Sc., 225 pp.

The most important task in a chemical process is controlling the process to achieve of process safety, production rates, and product quality. Generally, a chemical plant consists of reaction sections and separation section. Both sections need to control system for the safety and economical in plant. In this research, modification of the separation section of HDA process and the use of heat exchanger network (HEN) to save energy are proposed. The traditional HDA distillation train is replaced by a new design. Furthermore, four new heat exchanger networks (HENs) are developed. The plantwide control structures are designed using the disturbance load propagation method (Wongsri, M., 1990) and heat pathway heuristics (Wongsri, M. and Hermawan Y.D., 2005), respectively. Two kinds of disturbances: thermal and material disturbances are used in evaluation of the plantwide control structures. The performances of the heat integrated plants (HIPs) and the control structures evaluated dynamically by commercial software HYSYS.

Department……Chemical Engineering…	Student's signature……Kasin P.……
Field of study…Chemical Engineering…	Principal Advisor's signature……Montree……
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Finally, I would like to dedicate this dissertation to my family for their edification, inspiration, encouragement 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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NOMENCLATURES

ALn	The typical HDA process Alternative n; (n = 1, 2, 3and 4)
HIPn	Heat-Integrated Process of HDA plant n; (n = 1, 2, 3and 4)
CS1	Design of control structure 1
CS2	Design of control structure 2
CS3	Design of control structure 3
D	Disturbance
MER	Maximum energy recovery
DMER	Dynamic maximum energy recovery
e	error
Ci	Cold stream
Hi	Hot stream
HEN	Heat exchanger network
HIP	Heat integrated process or Heat integrated plant
HPH	Heat pathway heuristics
IAE	Integral absolute error
kF	The kinetic expression for the reaction
P	Pressure, (bar)
r	Reaction rate of reaction
RHEN	Resilient heat exchanger network
T	Temperature, ^{o}C
W	The heat capacity flowrate units of kw/hr-^{o}C
B	Bottom product flow rate (kmol/h)
D	Column distillate flow rate (kmol/h)
F	Feed flow rate (kmol/h)
Nt	Total number of theoretical trays
Nrt	Tray number of feed section
Pb	Column bottom pressure (bar)
Pd	Column top pressure (bar)
Qc	Heat duty of condenser (kw/h)
Qr	Heat duty of reboiler (kw/h)
Tb	Column bottoms temperature (°C)
Td	Column top temperature (°C)
COC	Annual Capital Costs
COP	Annual Operating Costs
TAC	Total Annual Costs

CHAPTER I

INTRODUCTION

This chapter introduces the importance and reasons for research, research objectives, scope of research and procedure

1.1 Importance and reasons for research

At present, there is growing pressure to reduce capital investment, working capital and operating cost, and to improve safety and environmental concerns. Therefore, design engineers have to achieve these tasks by, for example, eliminating many surge tanks, increasing the number of recycle streams or introducing energy integration for both existing and new plants. Nevertheless, recycle streams and energy integrations introduce a feedback of material and energy among units upstream and downstream. They also interconnect separate unit operations and create a path for disturbance propagation. Therefore, a plantwide process control strategy is required to operate an entire plant in order to achieve its objectives.

Essentially, the plantwide control problem is how to develop the control loops needed to operate an entire process and to achieve its design objectives. The problem is extremely complex and is very much opened. There are a combinatorial number of possible choices and alternative strategies to control and manage the disturbance load entering the process. It is recognized that one key tool to be used in designing more effective control structures is dynamic simulation. With the aid of simulation, both research and industrial practitioners can test their ideas and gain insight into process behavior that would not normally be intuitive given the complexity of an entire process design. Unfortunately for the research world, much plantwide information is proprietary and not available in open literature.

However, in recent years several good case studies have been published that allow testing of new control ideas on the level of complexity seen in a typical industrial chemical manufacturing plant. One such study that has been presented by Luyben and Tyreus (1998) provides detailed process and rating information for the manufacture of hydrodelakylation of toluene (HDA) process. In addition, the authors propose and test an entire control structure using their nine-step approach to plantwide control design. HDA process of toluene to benzene consists of a reactor, furnace, vapor-liquid separator, recycle compressor, heat exchangers and distillations. This plant is a realistically complex chemical process. It is considering that the energy integration for realistic and large processes is meaningful and useful, it is essential to design a control strategy for process associate with energy integration, so it can be operated well. So many controls of heat-integrated systems have been studied by several workers. Terrill and Douglas (1987a, b, c) have proposed six HEN alternatives for the HDA process, in which their energy saving ranges between 29 % and 43 %. Further, study of plantwide process control has also been done by several authors. Luyben et al. (1998) presented a general heuristic design procedure for plantwide process control

This study uses heat exchanger network (HEN) to save energy in the hydrodealkylation (HDA) plant and control structure will be designed using disturbance load propagation method (Wongsri,1990) and Luyben heuristic design method (1999), respectively. So the main objective of this study is to use plantwide control strategies to develop the new control structures for the HDA process with heat-integrated process structures schemes that are designed to achieve the control objective and reduce the cost of production. In this work, the performances of the heat exchanger network (HEN) are designed and their control structures are evaluated via commercial software HYSYS to carry out both steady state and dynamic simulations.

1.2 Objective of the Research

The objectives of this work are listed below:

1. To design Heat-integrated processes (HIPs) structures for HDA plant (based on alternative 1, 2, 3 and 4 Douglas, J. M. 1987)
2. To design control structures for heat exchanger networks (HENs) with heat-integrated process structures in HDA plant
3. To evaluate performance of design heat-integrated process structures for HDA plant.

1.3 Scopes of the Research

The Scope of this work are listed below:

1. Description and data of HDA plant are obtained from Douglas, J. M. (1988), William L. Luyben, Bjorn D. Tyreus, and Michael L. Luyben (1998), and William L. Luyben (2002).
2. The heat exchanger network with control structures of the HDA plant are programmed using a commercial process simulator HYSYS for control structure performance tests.
3. The design heat-integrated process structures for HDA plant is obtained from Terrill and Douglas 1987 (alternative 1, 2, 3 and 4)
4. The design control structures for heat-integrated process HDA plant are design using Luyben's heuristics method.

1.4 Contributions of the Research

The contributions of this work are as follows:

1. The new plantwide control structures for typical of HDA process alternatives 1, 2, 3 and 4.
2. The new heat-integrated processes (HIPs) structures for HDA process HIP1, 2, 3 and 4.
3. The new plantwide control structures with heat-integrated processes (HIPs) structures for HDA process HIP1, 2, 3 and 4.
4. Process flow diagrams of HDA process with heat-integration process HIP1, 2, 3 and 4 have been simulated.

1.5 Research Procedures

The procedures of this research are as follows:

1. Study plantwide process control theory.
2. Study HDA process and related information.

3. Study and Design heat-integrated processes (HIPs) structures for HDA process of HIP1, 2, 3 and 4 by using HEN heuristics.
4. Steady state simulation of heat-integrated processes (HIPs) structures of HDA process HIP1, 2, 3 and 4.
5. Dynamic simulation of heat-integrated processes (HIPs) structures of HDA process HIP1, 2, 3 and 4.
6. Development of the new design plantwide heat-integrated process structures for HDA process HIP1, 2, 3 and 4.
7. Dynamic simulation for the heat-integrated process structures for HDA process HIP1, 2, 3 and 4.
8. Evaluate and analyze of the dynamic performance of the heat-integrated processes (HIPs) structures.
9. Conclusion of the thesis.

1.6 Research Contents

This thesis is divided into seven chapters.

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

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

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

Chapter IV describes the process description and the design of heat exchanger networks for the typical of HDA plant.

Chapter V the strategy to design of heat integrated plant (HIP) of HDA process is proposed.

Chapter VI the three new plantwide control structures and dynamic simulation for the heat integrated plant (HIP) structure of HDA process are present.

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

CHAPTER II

LITERATURE REVIEW

Our purpose of this chapter is to present a review of the previous work on the conceptual design of chemical process, heat exchanger networks (HENs) and plantwide control design.

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

In a series of papers, Fisher et al. (1988 a, b, c) presented a study of the interface between design and control including process controllability, process operability and selecting a set of controlled variables. At the preliminary stages of a process design, most plants are uncontrollable. That is normally there are not enough manipulative variables in the flowsheet to be able to satisfy all of the process constraints and to optimize all of the operating variables as disturbances enter the plant. In order to develop a systematic procedure for controllability analysis, Fisher et al. (1988a) used the design decision hierarchy described by Douglas (1985) as the decomposition procedure and considered HDA process as a case study. Where at some levels, that are level 1, 2 and 3, the process is uncontrollable, but controllable at level 4 and level 5. If the available manipulated variables are compared with the constraints and operating variables introduced at each level, the preliminary controllability criterion can often be satisfied. Beside controllability analysis, Fisher et al. (1988b) also focused on operability analysis. The goal of operability analysis is to ensure that there is an adequate amount of equipment over design so that they could satisfy the process constraints and minimize a combination of the operating costs and over design costs over the entire range of anticipated process disturbances. They also followed the same hierarchical procedure to develop operability analysis. For HDA process, the operability decisions were encountered at each level. Fisher et al. (1988c) proposed steady state control structure for HDA process using an optimum steady state control analysis. They found the values of manipulated variables (that minimize the total operating costs for various values of the disturbances) and used it to define the controlled variables.

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

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

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

2.2 Heat Exchanger Network (HEN)

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

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

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

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

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

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

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

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

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

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

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

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

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

2.3 Design and Control of Energy-Integrated Plants

Renanto Handogo and W. L. Luyben (1987) studied the dynamics and control of heat-integrated reactor/column system. An exothermic reactor was the heat source, and a distillation column reboiler was the heat sink. Two types of heat-integration systems were examined: indirect and direct heat integration. Both indirect and direct heat integration systems are found in industry. In the indirect heat-integration system, steam generation was used as the heating medium for the reboiler. The direct heat integration system used the reactor fluid to directly heat the column. The indirect heat-integration system was found to have several advantages over the direct heat integration system in term of its dynamic performance. Both systems were operable for both large and small temperature differences between the reactor and column base.

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

M.L. Luyben, B.D. Tyreus, and W.L. Luyben (1997) presented A general heuristic design procedure is presented that generates an effective plantwide control structure for an entire complex process flowsheet and not simply individual units. The nine steps of the proposed procedure center around the fundamental principles of plantwide control: energy management; production rate; product quality; operational, environmental and safety constraints; liquid-level and gas-pressure inventories; makeup of reactants; component balances; and economic or process optimization. Application of the procedure is illustrated with three industrial examples: the vinyl acetate monomer process, the Eastman plantwide-control process, and the HDA process. The procedure produced a workable plantwide control strategy for a given process design. The control system was tested on a dynamic model built with TMODS, Dupont's in-house simulator.

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

Wongsri and Kietawarin (2002) presented a comparison among 4 control structuresdesigned for withstanding disturbances that cause production rate change of HDA process. The changes had been introduced to the amount of toluene and feed temperature before entering the reactor. Compared with the reference control structure using a level control to control toluene quantity in the system, the first control scheme measured toluene flow rate in the process and adjusted the fresh toluene feed rate. This structure resulted in faster dynamic response than the reference structure. The second control scheme was modified from the first scheme by adding a cooling unit to control the outlet temperature from the reactor , instead of using internal process flow. The result was to reduce material and separation ratio fluctuations within the process. The product purity was also quite steadily. In the third control scheme, a ratio control was introduced to the secon control scheme for controlling the ratio of hydrogen and toluene within the process. This scheme showed that it could withstand large disturbances. Dynamic study showed that the control structure had significant effect on process behavior. A good system control should quickly respond to disturbances and adjust itself to steady state while minimizing the deviation of the product quality. The control structures were compared with reference on plantwide process control book, Luyben 1998, the result was performance of these structures higher than reference.

Wongsri and Thaicharoen (2004) presented 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 structures 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 reboiler 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. The energy used in CS6 control structure is less than CS1 and CS4.

Wongsri and Hermawan Y.D. (2005) studied the control strategies for energy-integrated HDA plant (i.e. alternatives 1 and 6) based on the heat pathway heuristics (HPH), i.e. selecting an appropriate heat pathway to carry associated load to a utility unit, so that the dynamic MER can be achieved with some trade-off. In they work, a selective controller with low selector switch (LSS) is employed to select an appropriate heat pathway through the network. The new control structure with the LSS has been applied in the HDA plant alternatives 1 and 6. The designed control structure is evaluated based on the rigorous dynamic simulation using the commercial software HYSYS. The study reveals that, by selecting an appropriate heat pathway through the network, the utility consumptions can be reduced according to the input heat load disturbances; hence the dynamic MER can be achieved.

Kunajitpimol (2006) presented the resilient heat exchanger networks to achieve dynamic maximum energy recovery, plantwide control structures, and control strategies are designed for Butane Isomerization plant. The control difficulties associated with heat integration are solved by adding auxiliary utilities which is kept minimal. Four alternatives of heat exchanger networks (HEN) designs of the Butane Isomerization plant are proposed. They used the heat from the reactor effluent stream to provide the heat for the column reboiler. The energy saved is 24.88% from the design without heat integration, but the additional capital is 0.67 % due to adding of a process to process exchanger and an auxiliary utility exchanger to the process. The plantwide control configuration of heat-integrated plant is designed following Luyben’s heuristic method. Various heat pathways throughout the network designed using Wongsri’s disturbance propagation method to achieve 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:

Page 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

Bob Richardson 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

Greg 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.

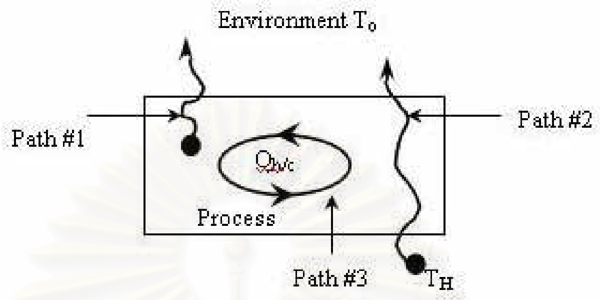


Figure 3.1 Heat pathways

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.

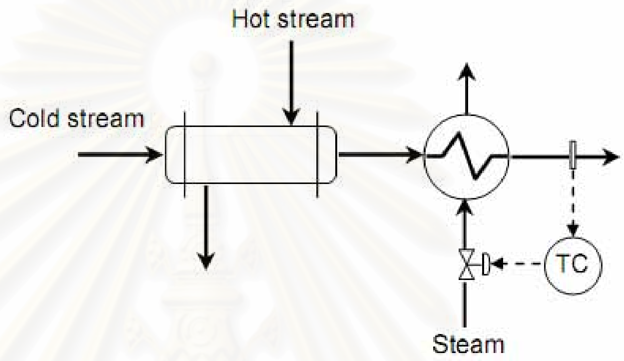


Figure 3.2 Control of process-to-process heat exchanger using the auxiliary utility.

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.

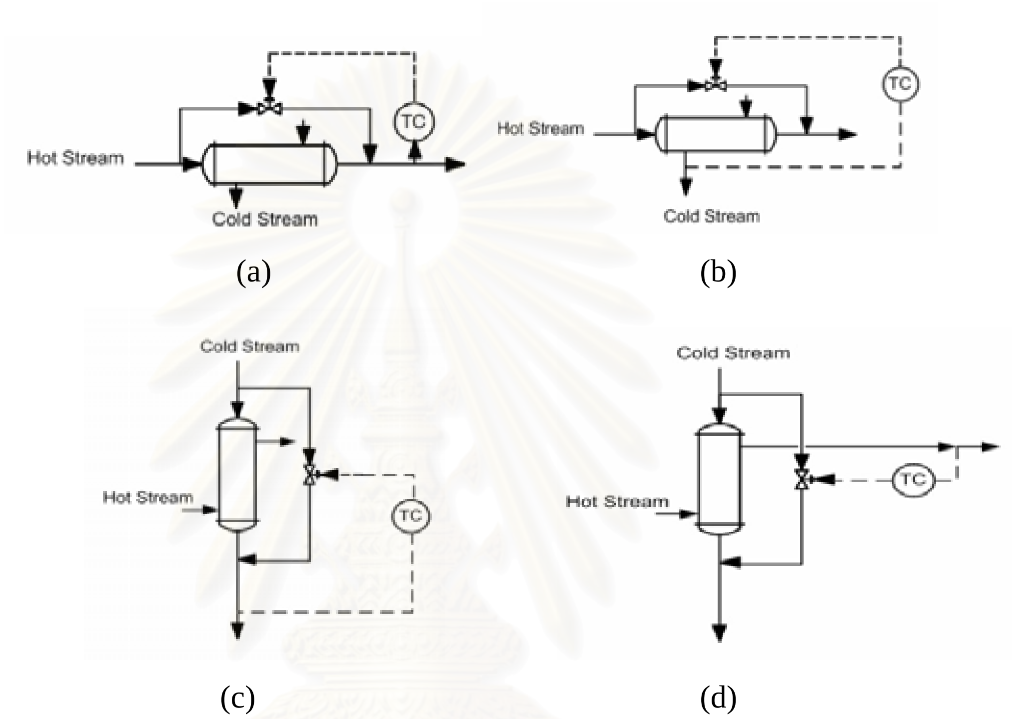


Figure 3.3: Bypass control of process-to-process heat exchangers. (a) Controlling and bypassing hot stream; (b) controlling cold stream and bypassing hot stream; (c) controlling and bypassing cold stream; (d) controlling hot stream and bypassing hot stream.

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

TYPICAL 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

$$\text{Toluene} + H_2 \rightarrow \text{benzene} + CH_4$$

$$2\ \text{Benzene} \leftrightarrow \text{diphenyl} + H_2$$

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.

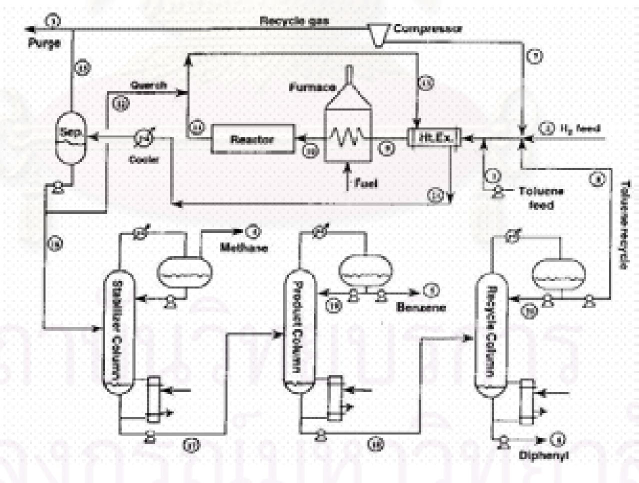


Figure 4.1 Hydrodealkylation (HDA) of Toluene Process (alternative 1)

Component physical property data for the HDA process were obtain from William L. Luyben, Bjorn D. Tyreus, Michael L. Luyben (1999)

4.2 Alternatives of Typical HDA Process

Terrill and Douglas (1987) design six different energy-saving alternatives to the base case**,** the typical HDA process alternatives 1-6 show on Figure 4.2. 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.

Table 4.1 The energy saving from the energy integration of Typical HDA process

Typical HDA process	Alternatives					
	1	2	3	4	5	6
Utilities usage for alternatives, MW	9.1	7.7	7.4	7.3	7.3	7.3
Energy savings from new HEN, %	29	40	42	43	43	43
TAC for alternatives with Base-case design values, $10^6/year	6.4	6.5	6.4	6.1	6.0	6.0

The energy saving from the energy integration fall between 29 and 43 % show on Table 4.1 This work will consider the HDA alternatives 1-4 because there are non-highly heat integrated process.

4.3 Design of Heat Exchanger Networks of Typical HDA Process

At this point, the heat exchanger network design method provide by Wongsri (1990) is used to design the heat exchanger networks for HDA process. The design procedures and definitions from previous chapters will be methods to design and compare with the preliminary stage of a process design without energy integration. The Problem Table Method is applied to find pinch temperature and reach maximum energy recovery (MER). The cost estimated will be consequence to compare and choose the best network that more optimal for the HDA process. The information for design is shown in the following Table 4.2

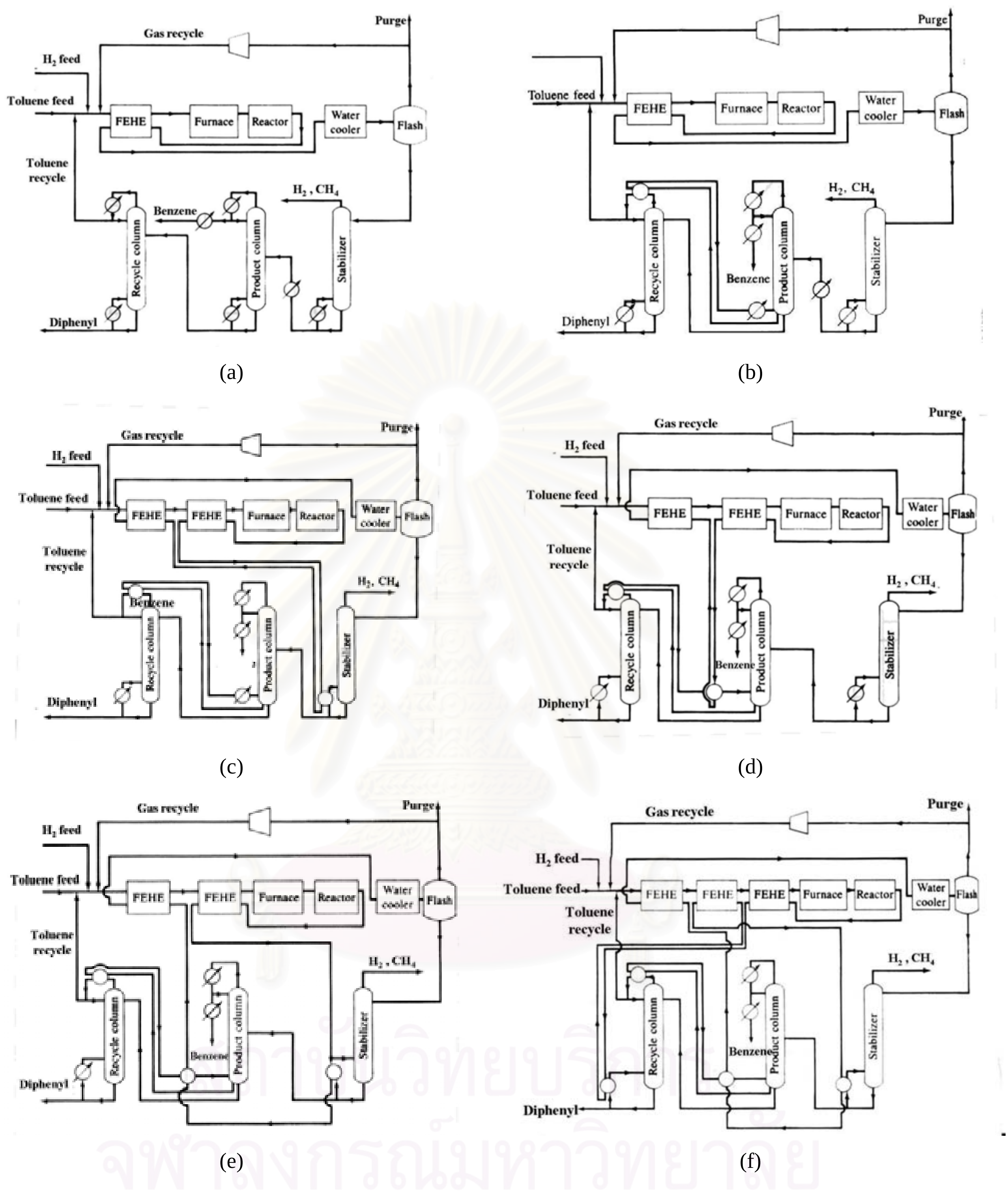


Figure 4.2 the typical of HDA process six alternatives structures: (a) alternative 1, (b) alternative 2, (c) alternative 3, (d) alternative 4, (e) alternative 5, (f) alternative 6

Table 4.2 The information for typical of HDA process

Stream Name	Tin (°C)	Tout (°C)	W	Duty (kW)
H1: Reactor Product Stream (RPS)	621	45	33	19008
H2 : Recycle Column Condenser (RCC)	183	181	200	400
C1 : Reactor Feed Stream (RFS)	65	621	32.24	17925.44
C2 : Product Column Reboiler (PCR)	145	193	91	4368
C3 : Stabilizer Column Reboiler (SCR)	190	215	59	1475
C4 : Recycle Column Reboiler (RCR)	349.5	350.7	456	547.2

4.3.1 Resilient Heat Exchanger Network of HDA process alternative 1

There are two streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 4.4. At the minimum heat load condition, the pinch temperature occurs at 621/611°C. The minimum utility requirements have been predicted 322.4 kW/hr of hot utilities and 1404.96 Btu/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.4. Figure 4.3 shows a design of resilient heat exchanger network for HDA process alternative 1. In our case as shown in Figure 4.3, the minimum temperature difference in the process-to-process-heat-exchangers ΔT_{min} is set to be 10 °C.

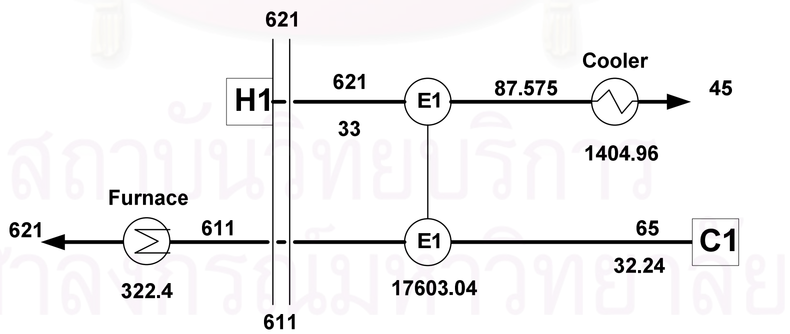


Figure 4.3 The resilient heat exchanger network alternative 1

Table 4.3 Process stream data for alternative 1

Stream Name	Tin (°C)	Tout (°C)	W	Duty (kW)
H1	621	45	33	19008
C1	65	621	32.24	17925.44

Table 4.4 Problem table for alternative 1

W		T		Δ T	sum W	Require	Interval	Cascade	sum Interval
H1	C1	hot	cold			(Heater)		(Cooler)	
0	0	631	621			Qh			
0	32.24	621	611	10	-32.24	322.4	-322.4	0	-322.4
33	32.24	75	65	546	0.76	0	414.96	414.96	92.56
33	0	45	35	30	33	414.96	990	1404.96	1082.56
								Qc	

Synthesis table for Cold End of alternative 1

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	18678	33	611	45	330	0	select BC
C1	16958.24	32.24	75	601	322.4	322.4	select
b) state 2							
H1	1719.76	33	97.11	45	330	0	to Cooler
C1							match to H1

Synthesis table for Hot End of alternative 1

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	0	33	621	621	330	330	select AH
C1	322.4	32.24	611	621	0	322.4	select
b) state 2							
H1							match to H1
C1	322.4	32.24	611	621	0	322.4	to Heater

4.3.2 Resilient Heat Exchanger Network of HDA process alternative 2

There are two streams in the networks. So we can find Pinch temperature using Problem table method as shown in Table 4.6. At the minimum heat load condition, the pinch temperature occurs at 155/145^{o}C. The minimum utility requirements have been predicted 3968 kW/hr of hot utilities and 0 kW/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.6. Figure 4.4 shows a design of resilient heat exchanger network for HDA process alternative 2. In our case as shown in Fig. 4.4, the minimum temperature difference in the process-to-process-heat-exchangers, is set to be 10 ^{o}C.

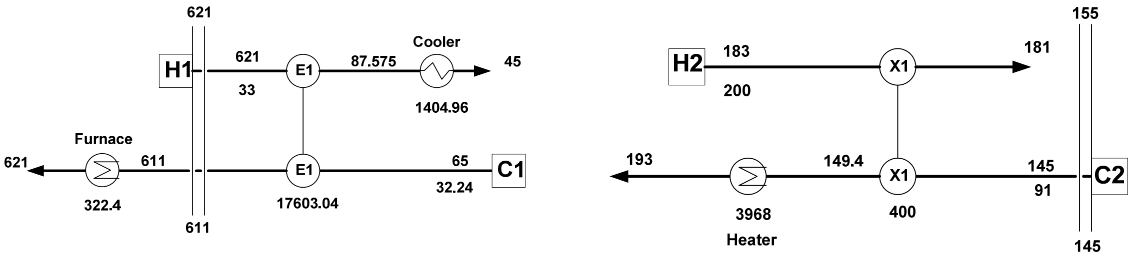


Figure 4.4 The resilient heat exchanger network alternative 2

Table 4.5 Process stream data for alternative 2

Stream Name	Tin (°C)	Tout (°C)	W	Duty (kW)
H2	183	181	200	400
C2	145	193	91	4368

Table 4.6 Problem table for alternative 2

W		T		ΔT	sum W	Require	Interval	Cascade	sum Interval
H2	C2	hot	cold			(Heater)		(cooler)	
0	0	203	193			Qh			
0	91	183	173	20	-91	3968	-1820	2148	-1820
200	91	181	171	2	109	2148	218	2366	-1602
0	91	155	145	26	-91	2366	-2366	0	-3968
								Qc	

Synthesis table for Cold End of alternative 2

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	18678	33	611	45	330	0	select BC
C1	17280.64	32.24	75	611	322.4	322.4	
C4	72.93	104.18	154.30	155	0	104.18	select
b) state 2							
H1	18500.89	33	605.63	45	330	0	to Cooler
C1	17280.64	32.24	75	611	322.4	322.4	select
C4							match to H1
c) state 3							
H1	897.85	33	72.21	45	330	0	to Cooler
C1							match to H1
C4							

Synthesis table for Hot End of alternative 2

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	0	33	621	621	330	330	select C[H]
C1	322.40	32.24	611	621	0	322.4	
b) state 2							
H1							select AH
C1	322.40	32.24	611	621	0	322.4	select

4.3.3 Resilient Heat Exchanger Network of HDA process alternative 3

There are three streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 4.8. At the minimum heat load condition, the pinch temperature occurs at 200/190^{o}C. The minimum utility requirements have been predicted 1477.44 kW/hr of hot utilities and 1085 kW/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.8. Figure 4.5 shows a design of resilient heat exchanger network for HDA process alternative 3. In our case as shown in Fig. 4.5, the minimum temperature difference in the process-to-process-heat-exchangers, $\Delta T_{\min}$ is set to be 10 ^{o}C.

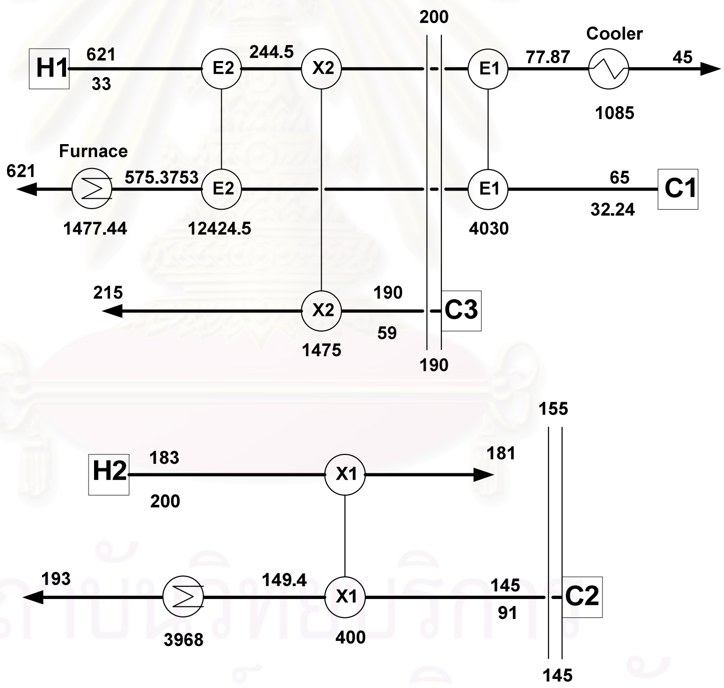


Figure 4.5 The resilient heat exchanger network alternative 3

Table 4.7 Process stream data for alternative 3

Stream Name	Tin (^{o}C)	Tout (^{o}C)	W	Duty (kW)
H1	621	45	33	19008
C1	65	621	32.24	17925.44
C3	190	215	59	1475

Table 4.8 Problem table for alternative 3

W			T		ΔT	sum W	Require	Interval	Cascade	sum Interval
H1	C1	C3	hot	cold			(Heater)		(cooler)	
0	0	0	631	621			Qh			
0	32.24	0	621	611	10	-32.24	1477.44	-322.4	1155.04	-322.4
33	32.24	0	225	215	396	0.76	1155.04	300.96	1456	-21.44
33	32.24	59	200	190	25	-58.24	1456	-1456	0	-1477.44
33	32.24	0	75	65	125	0.76	0	95	95	-1382.44
33	0	0	45	35	30	33	95	990	1085	-392.44
									Qc	

Synthesis table for Cold End of alternative 3

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	5445	33	210	45	330	0	select BC
C1	4030	32.24	75	200	322.4	322.4	select
b) state 2							
H1	1415	33	87.88	45	330	0	to Cooler
C1							match to H1

Synthesis table for Hot End of alternative 3

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	13233	33	611	210	330	330	select C[H]
C1	13573.04	32.24	200	621	0	322.4	
C3	885	59	200	215	0	590	select
b) state 2							
H1	12348	33	611	236.82	330	0	select AH
C1	13573.04	32.24	200	621	0	322.4	select
C3					0	260	to heater
c) state 3							
H1							match to C1
C1	1225.04	32.24	583.00248	621	0	652.4	to heater

4.3.4 Resilient Heat Exchanger Network of HDA process alternative 4

There are four streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 4.10. At the minimum heat load condition, the pinch temperature occurs at 155/145^{o}C. The minimum utility requirements have been predicted 3936.24 kW/hr of hot utilities and 1050.8 kW/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.10. Figure 4.6 shows a design of resilient heat exchanger network for HDA process alternative 4. In our case as shown in Fig. 4.6, the minimum temperature difference in the process-to-process-heat-exchangers, ΔT_{min} is set to be 10^{o}C.

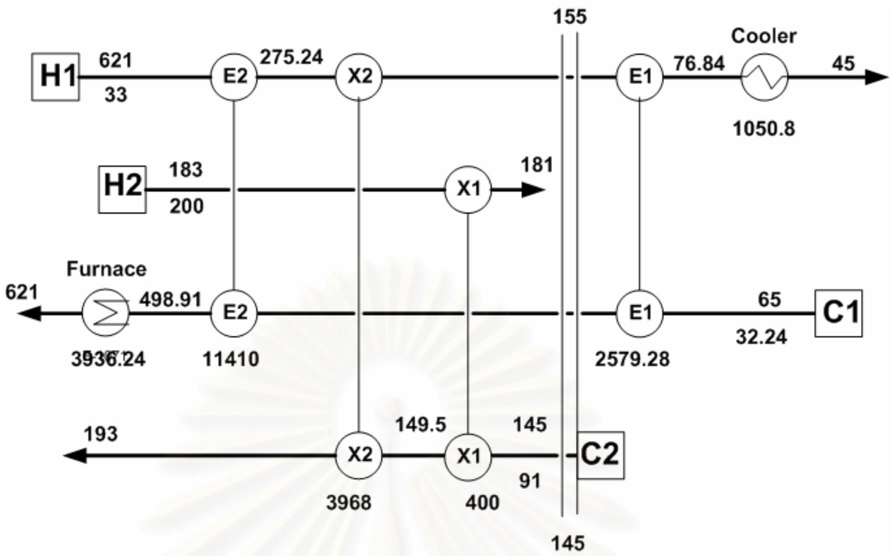


Figure 4.6 The resilient heat exchanger network alternative 4

Table 4.9 Process stream data for alternative 4

Stream Name	Tin (^{o}C)	Tout (^{o}C)	W	Duty (kW)
H1	621	45	33	19008
H2	183	181	200	400
C1	65	621	32.24	17925.4
C2	145	193	91	4368

Table 4.10 Problem table for alternative 4

W				T		ΔT	sum W	Require	Interval	Cascade	sum Interval
H1	H2	C1	C2	hot	cold			(Heater)		(cooler)	
0	0	0	0	631	621			Qh			
0	0	32.24	0	621	611	10	-32.24	3936.24	-322.4	3613.84	-322.4
33	0	32.24	0	203	193	418	0.76	3613.84	317.68	3931.52	-4.72
33	0	32.24	91	183	173	20	-90.24	3931.52	-1804.8	2126.72	-1809.52
33	200	32.24	91	181	171	2	109.76	2126.72	219.52	2346.24	-1590
33	0	32.24	91	155	145	26	-90.24	2346.24	-2346.24	0	-3936.24
33	0	32.24	0	75	65	80	0.76	0	60.8	60.8	-3875.44
33	0	0	0	45	35	30	33	60.8	990	1050.8	-2885.44
										Qc	

Synthesis table for Cold End of alternative 4

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	3630	33	155	45	330	0	select BC
C1	2256.8	32.24	75	145	322.4	322.4	select
b) state 2							
H1	1050.8	33	76.84	45	330	0	to Cooler
C1							match to H1

Synthesis table for Hot End of alternative 4

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	15048	33	611	155	330	330	
H2	200	200	182	181	200	0	select A[H]
C1	15346.24	32.24	145	621	0	322.4	
C2	3458	91	155	193	0	910	select
b) state 2							
H1	14718	33	611	165	330	330	select C[H]
H2							match to C2
C1	15023.84	32.24	155	621	0	322.4	
C2	3058	91	159.40	193	0	1110	select
c) state 3							
H1	11660	33	611	257.67	330	0	select A[H]
C1	15023.84	32.24	155	621	0	322.4	select
C2					0	780	to heater
d) state 4							
H1							match to C1
C1	3033.84	32.24	526.90	621	0	652.4	to heater

The various alternatives of heat exchanger network are designed for the HDA process, the energy saved from the alternative 1 (Base case) as in Table 4.11

Table 4.11 Energy integration for HDA process

The Typical HDA process	Alternatives			
	AL 1 (BC)	AL 2	AL 3	AL 4
Furnace	322.42	322.42	1477.44	3936.24
Cooler	1404.96	1404.96	1085	1050.8
Stabilizer column reboiler	1475	1475	0	0
Product column reboiler	4368	0	0	0
Recycle column reboiler	547.2	547.2	547.2	547.2
Heater Product reboiler	0	3968	3968	0
Hot utilities usage, (kW)	6712.62	6312.62	5992.64	4483.44
Cold utilities usage, (kW)	1404.96	1404.96	1085	1050.8
Total Hot & Cold utilities (kW)	8117.58	7717.58	7077.64	5534.24
Energy savings from RHEN, %	0	4.9	12.8	31.8

4.3.5 Alternative Structures of Typical HDA Process

Seven alternatives of heat exchanger networks (HENs) designs of the HDA process are proposed to save energy from the alternative 1 (BC) and use to evaluate performance of control structures are designed both simply heat-integrated process and complex heat-integrated process.

In Figure 4.7 show the alternative 1 of typical HDA process with simply energy integration, we used a feed-effluent heat exchanger (FEHE) to reduce the amount of fuel burned in the furnace. The heat of reaction and the heat added in the furnace are therefore removed in the flooded condenser.

In alternative 1 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 shown on Figure 4.7.

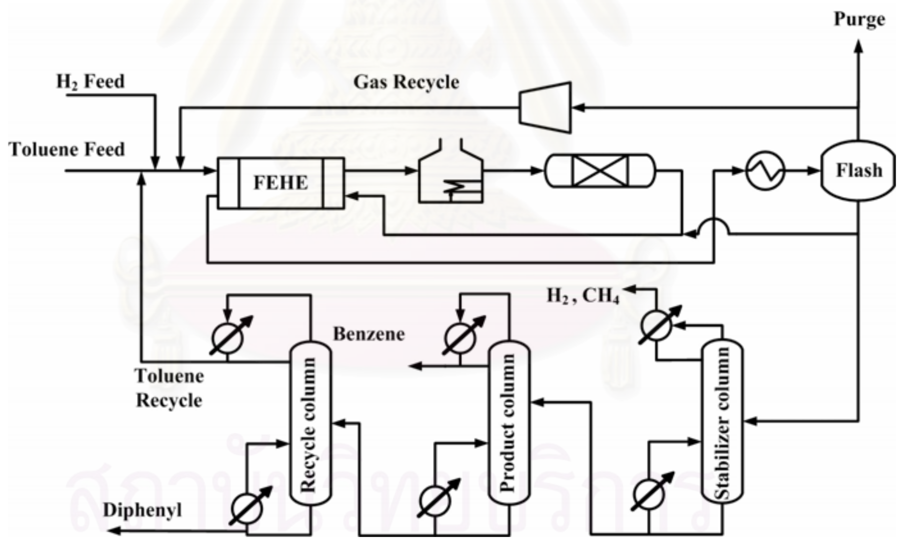


Figure 4.7 The typical HDA process alternative 1

In alternative 2 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 shown on Figure 4.8.

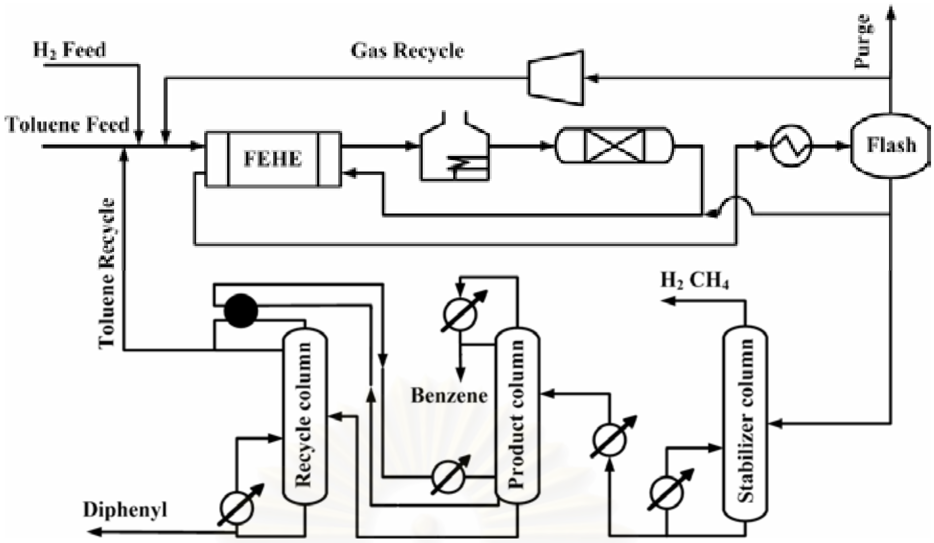


Figure 4.8 The typical HDA process alternative 2

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 shown on Figure 4.9.

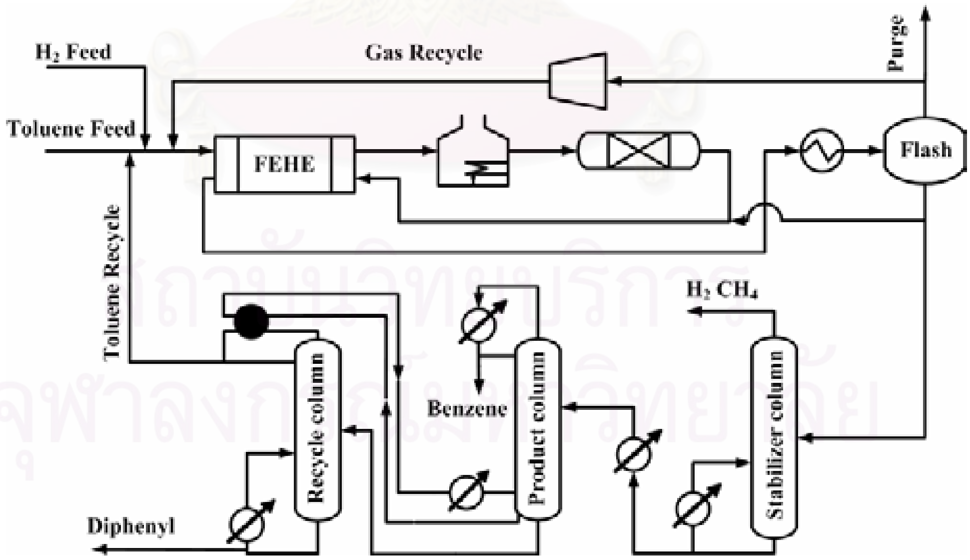


Figure 4.9 The typical HDA process alternative 3

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 shown on Figure 4.10.

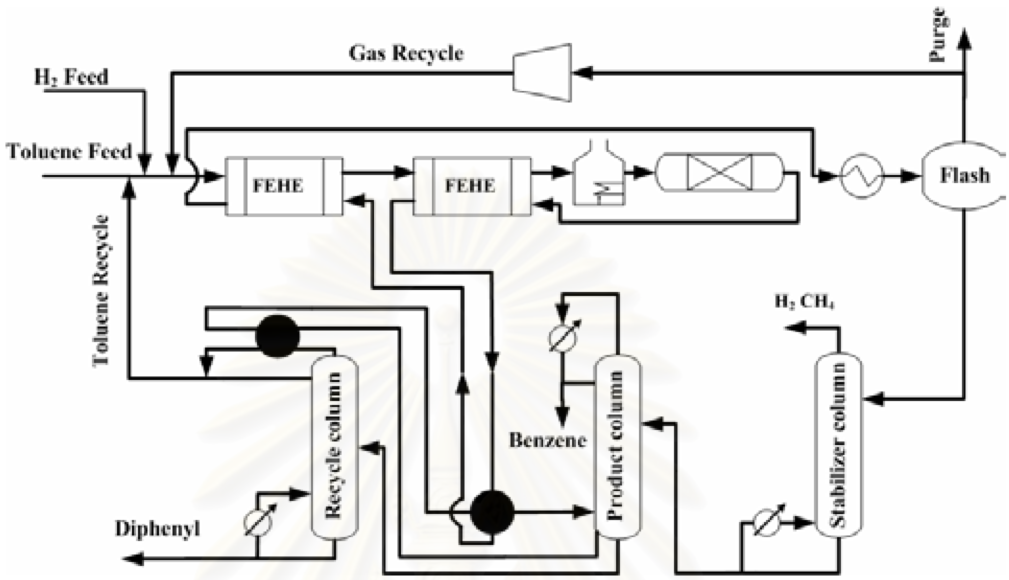


Figure 4.10 The typical HDA process alternative 4

4.4 Steady State Modeling for Typical of HDA Process

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 Typical HDA process alterative 1, 2, 3 and 4. Figures 4.11 to 4.14 show the HYSYS flowsheets of the HDA process with energy integration schemes for alternatives 1, 2, 3 and 4, respectively. 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.

4.5 Comparisons of Energy from Steady State Simulation for Typical of HDA Process

From steady state simulation results by HYSYS, the energy saved from the base case heat consumption as shown in Table 4.12.

Table 4.12 Energy integration for Typical HDA process (Steady State Simulation)

Typical HDA process	Alternatives			
	AL 1 (BC)	AL 2	AL 3	AL 4
Furnace	385.9	384.7	1721.7	2126.1
Cooler	2212.0	2307.1	2074.6	1157.2
Stabilizer Column Reboiler	1273.0	1273.1	0	1280.1
Product Column Reboiler	3461.8	0	0	0
Recycle Column Reboiler	481.5	565.1	563.7	563.7
Heater Product reboiler	0	2792.3	2792.3	0
Total Reboiler	5216.3	4630.6	3356.0	1843.8
Stabilizer Column Condenser	176.7	176.9	373.8	182.9
Product Column Condenser	4061.3	3797.1	3809.6	3805.8
Recycle Column Condenser	437.4	0	0	0
Total Condenser	4675.4	3974.0	4183.4	3988.7
Hot utilities usage, (kW)	5602.1	5015.2	5077.7	3969.9
Cold utilities usage, (kW)	6887.4	6281.0	6258.0	5145.9
Total Hot & Cold utilities (kW)	12489.5	11296.3	11335.7	9115.8
Energy savings %	0	9.6	9.2	27.0

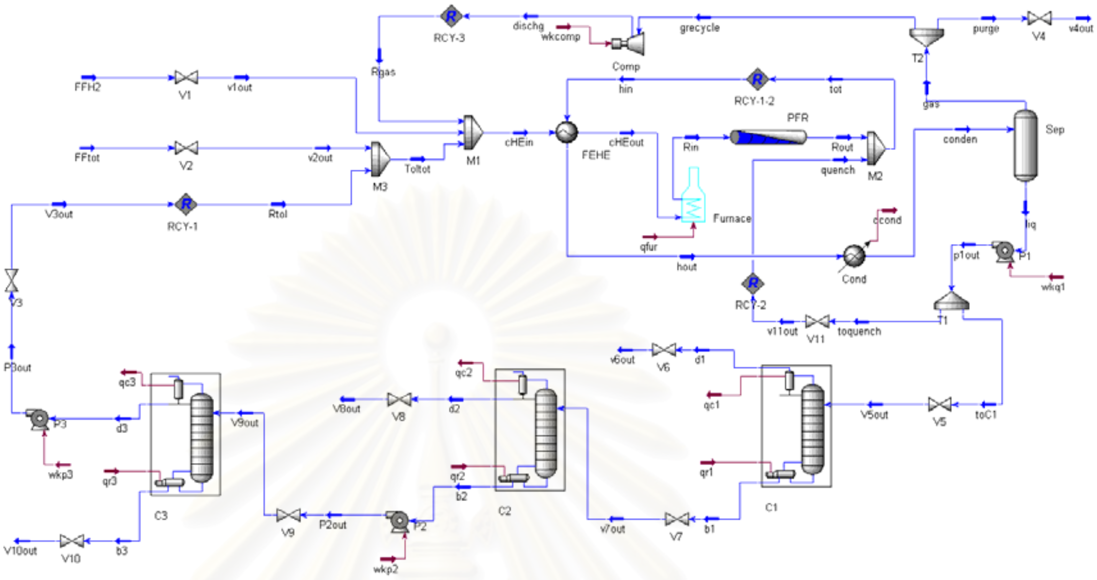


Figure 4.11 HYSYS Flowsheet of the Steady State Modeling of Typical HDA Process Alternative 1

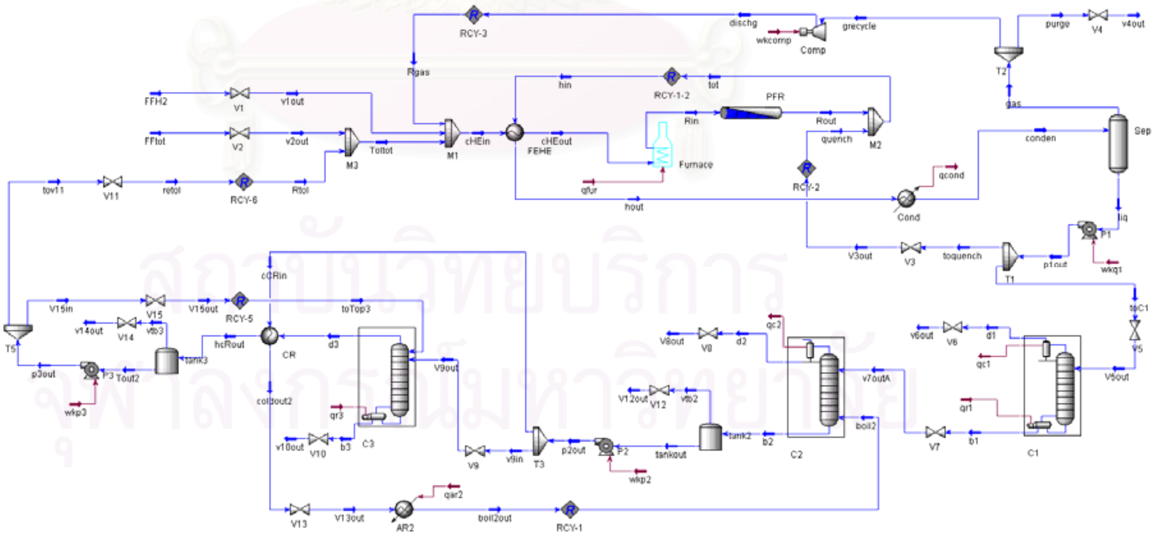


Figure 4.12 HYSYS Flowsheet of the Steady State Modeling of Typical HDA Process Alternative 2

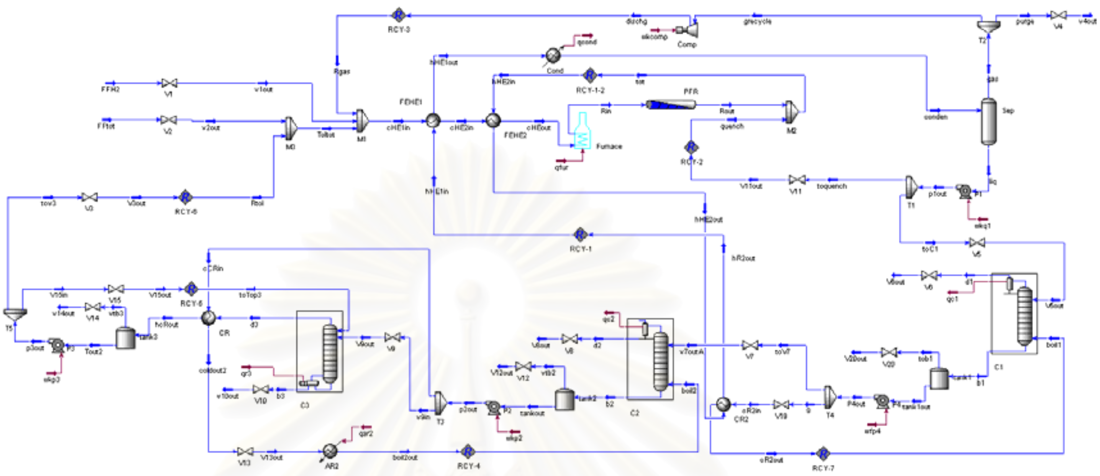


Figure 4.13 HYSYS Flowsheet of the Steady State Modeling of Typical HDA Process Alternative 3

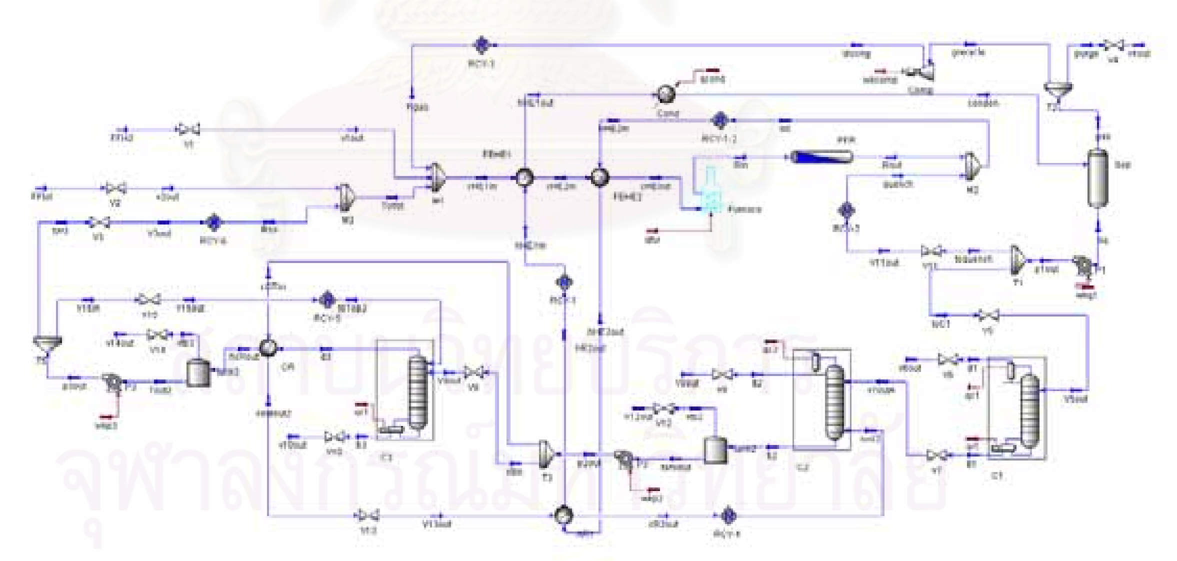


Figure 4.14 HYSYS Flowsheet of the Steady State Modeling of Typical HDA Process Alternative 4

CHAPTER V

DESIGN OF HEAT-INTEGRATED PROCESS STRUCTUERS FOR HDA PLANT

5.1 Design of Heat-Integrated Plant for HDA Process

Before getting to heat-integrated plant (HIP), the following assumptions are made. (1) The flowrate and composition of the streams in reaction part is similar as the typical process. (2) The flowrate and composition of product stream, fresh feed stream, recycle stream and purge stream are similar as the typical process.

From the assumptions, in this work we design only the separation part of the process. The stabilizer column separate hydrogen, methane and benzene as the overhead product, and benzene is the desired product from the product column bottom. After that, in the recycle column, toluene is separated from diphenyl, as the distillate and recycled back. In Figure 5.1 shows the separation section in the heat-integrated plant (HIP).

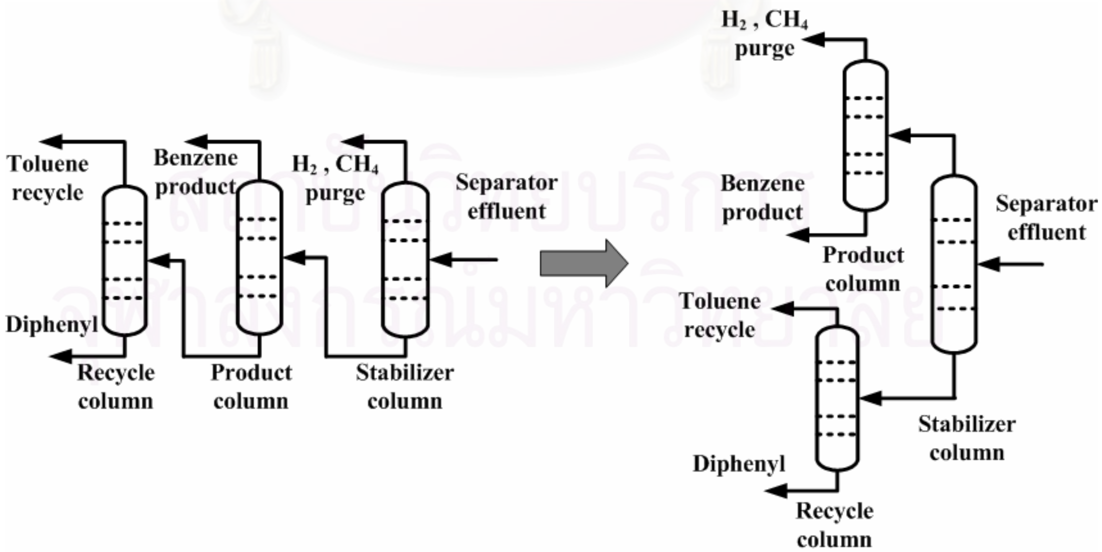


Figure 5.1 Separation section in the heat-integrated plant (HIP)

Then, a rigorous simulation of the different configurations was performed by using the commercial software HYSYS, already utilized in the simulation of the existing plant, was utilized to simulate all the units. The Peng and Robinson (PR) equation, which was successfully used in the plant simulation, was adopted for the calculation of the activity coefficients in liquid phase, whereas ideal behaviour was assumed for the vapour phase. In all the examined sequences, the number of stages obtained by the EOS shortcut method was utilized.

Table 5.1 Parameters and the results from shortcut of distillation design

Parameters	Typical HDA process (AL1)			Heat-integrated plant (HIP1)		
	Col.1	Col.2	Col.3	Col.1	Col.2	Col.3
F	174.90	166.21	42.44	175.15	132.66	42.49
D	8.69	123.77	39.56	132.66	9.06	39.60
B	166.21	42.44	2.88	42.49	123.60	2.89
Tdi	51.05	93.33	133.33	154.79	55.56	137.62
Tbi	189.64	137.78	260.00	200.67	153.26	291.92
Pdi	10.34	2.068	2.068	6.84	6.14	2.068
Pbi	10.38	2.221	2.108	7.05	6.16	2.108
Nt	6	27	7	36	3	7
Nrt	3	15	5	20	3	4
Qci	176.72	4061.26	437.40	2664.23	1097.69	425.58
Qri	1273.01	3461.76	481.49	4494.11	175.97	321.34
Total Qc		4675.38			4187.51	
Total Qr		5216.26			4991.42	

We design the columns by using shortcut of distillation from HYSYS. Table 5.1 shows the parameters and the results from the design of shortcut distillation. When we compare with the typical design, the results are different, in the stabilizer column and product column but recycle column resemble in typical process. So we decide to use the column spec similar in the previous design (Luyben et al. 1998). Next, we trial and error to find the feed tray location by use the initial feed location from Kirkbride's equation. Table 5.2 shows the operating condition for the three configurations of the process, and Figure 5.2 shows the flow sheet of the heat-integrated plant (HIP1).

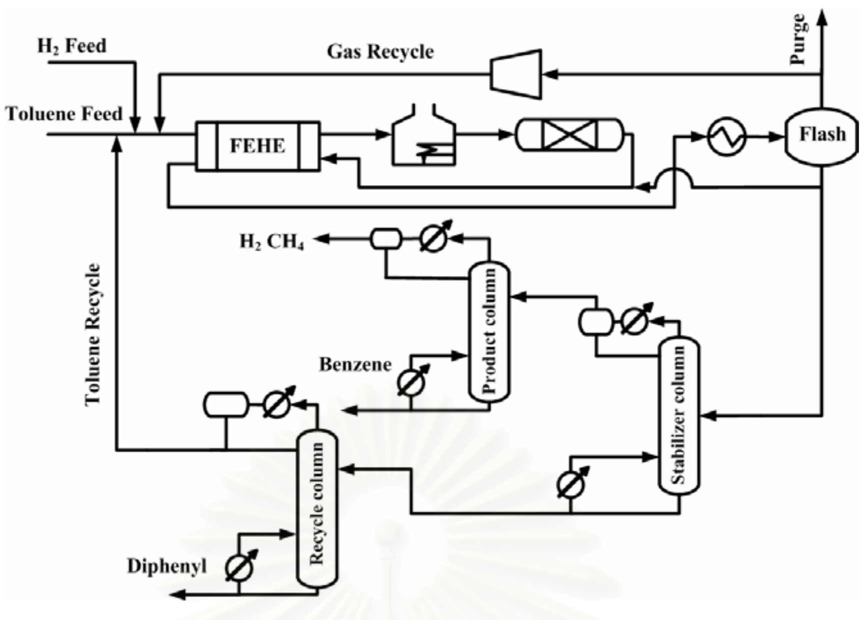


Figure 5.2 The heat-integrated plant (HIP1) of HDA process.

Table 5.2 The operating condition for the three configurations (steady state simulation)

Flowsheet 1. Stabilizer column (Col.1)	Typical HDA process (AL1)	Heat-integrated plant (HIP1)
No. of trays	6	36
Feed tray	3	20
Distillate flowrate (kgmol/h)	8.69	132.66
Hydrogen in distillate (mole frac.)	0.0884	0.0058
Methane in distillate (mole frac.)	0.8692	0.0570
Benzene in distillate (mole frac.)	0.0420	0.9368
Toluene in distillate (mole frac.)	0.0003	0.0003
Diphenyl in distillate (mole frac.)	0	0
Bottom flowrate (kgmol/h)	166.21	42.49
Hydrogen in distillate (mole frac.)	0	0
Methane in distillate (mole frac.)	0	0
Benzene in distillate (mole frac.)	0.7446	0.0009
Toluene in distillate (mole frac.)	0.2381	0.9311
Diphenyl in distillate (mole frac.)	0.0173	0.0680

2. Product column (Col.2)	Typical HDA process (AL1)	Heat-integrated plant (HIP1)
No. of trays		
Feed tray	15	3
Distillate flowrate (kgmol/h)	123.77	9.06
Hydrogen in distillate (mole frac.)	0	0.0849
Methane in distillate (mole frac.)	0	0.8351
Benzene in distillate (mole frac.)	0.9997	0.0800
Toluene in distillate (mole frac.)	0.0003	0

Diphenyl in distillate (mole frac.)	0	0
Bottom flowrate (kgmol/h)	42.44	123.60
Hydrogen in distillate (mole frac.)	0	0
Methane in distillate (mole frac.)	0	0
Benzene in distillate (mole frac.)	0.0006	0.9997
Toluene in distillate (mole frac.)	0.9315	0.0003
Diphenyl in distillate (mole frac.)	0.0679	0

3. Recycle column (Col.3)	Typical HDA process (AL1)	Heat-integrated plant (HIP1)
No. of trays	7	7
Feed tray	5	4
Distillate flowrate (kgmol/h)	39.56	39.60
Hydrogen in distillate (mole frac.)	0	0
Methane in distillate (mole frac.)	0	0
Benzene in distillate (mole frac.)	0.0006	0.0010
Toluene in distillate (mole frac.)	0.9993	0.9990
Diphenyl in distillate (mole frac.)	0	0
Bottom flowrate (kgmol/h)	2.88	2.89
Hydrogen in distillate (mole frac.)	0	0
Methane in distillate (mole frac.)	0	0
Benzene in distillate (mole frac.)	0	0
Toluene in distillate (mole frac.)	0.0003	0.0003
Diphenyl in distillate (mole frac.)	0.9997	0.9997

5.1.1 Determination of the operating pressures of the units

The operating pressures of the units in a thermodynamic configuration are determined by the given cold utility. First, if there is a condenser in a column, then the operating pressure of this column is determined in such a way that its top vapor could be condensed by the given cold utility. This is done by the calculation of the bubble point of the top vapor stream. The temperature of the top stream is determined based on the temperature of the cold utility and the given minimum approach temperature in the condenser. The minimum approach temperature is given based on the heuristics of King (1980). Then, the bottom pressure of the column is determined based on the calculated number of the theoretical trays and the given pressure drop for a single tray. The pressure drop for a single tray is given based on the heuristics of Kister (1992).

5.1.2 Economic Evaluation of Distillation Columns

With the above designed parameters, the synthesis of the above sequence design distillation flowsheets can be implemented based on the economic evaluation. The economic evaluation is based on the total annual cost of a flowsheet where the operating cost is calculated based on the cold and hot utility consumptions, while the capital cost is a sum of the costs of columns, condensers and reboilers. The capital cost of columns, condensers and reboilers is estimated based on the correlations and data provided by Douglas (1988).While the operating cost is calculated based on the heating and cooling loads and the unit costs of heating and cooling utilities.

Table 5.3 Energy and Total annual cost saving compared with the typical process (steady state simulation)

Description	Typical HDA process (AL1)			Heat-integrated plant (HIP1)		
	Col.1	Col.2	Col.3	Col.1	Col.2	Col.3
Duty of condenser x10^2(MJ/h)	6.3	146.0	15.8	95.9	39.5	15.3
Duty of reboiler x10^2(MJ/h)	45.8	125.0	17.3	162.0	6.3	11.6
Total Duty of condenser x10^2(MJ/h)	168.1			150.7		
Total Duty of reboiler x10^2(MJ/h)	188.1			179.9		
Annual operating costs (COP) x10^6($)	8235.8			6840.8		
Annual capital costs (COC) x10^2($)	659.0			755.8		
Total Annual Costs (TAC) x10^6($)	8235.8			6840.9		
Energy saving (%)	0			7.2		
TAC saving (%)	0			16.9		

The cost correlations are based on Douglas (1988). The capital cost is annualized over a period which is often referred to as plant life time and the operating time is assumed to be 8000 h per year. All the capital costs and operating cost were actualized by the CAPCOST program, automatically calculates the purchase price and the bare module factor corrections for pressure and materials. The results of energy and total annual cost saving compare with the typical process (steady state simulation) are shown in Table 5.3, the new sequence we can save the energy usage 7.2% and the total annual cost (TAC) 16.9% compared with the typical process.

5.2 Design of Resilient Heat Exchanger Networks (RHEN) for Heat-Integrated Plant (HIP) of HDA Process

At this point, the heat exchanger network design method provide by Wongsri (1990) is used to design the heat exchanger networks for heat-integrated plant (HIP) of HDA process. The design procedures and definitions from previous chapters will be methods to design and compare with the preliminary stage of a process design without energy integration. The Problem Table Method is applied to find pinch temperature and reach maximum energy recovery (MER). The cost estimated will be consequence to compare and choose the best network that more optimal for the heat-integrated plant of HDA process. The information for design is shown in the following Table 5.4

Table 5.4 The information of heat-integrated plant (HIP) of HDA Process

Stream Name	Tin (^{o}C)	Tout (^{o}C)	W	Duty (kW)
H1: Reactor Product Stream (RPS)	621	45	33	19008.0
C1 : Reactor Feed Stream (RFS)	65	621	32.24	17925.4
C2 : Stabilizer Column Reboiler (SCR)	200.93	239.40	116.40	4477.9
C3 : Recycle Column Reboiler (RCR)	290.59	291.9	244.43	320.2
C4 : Product Column Reboiler (PCR)	153.3	155	104.18	177.1

5.2.1 Design of RHEN for heat-integrated plant (HIP1)

There are two streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 5.6. At the minimum heat load condition, the pinch temperature occurs at 621/611^{o}C. The minimum utility requirements have been predicted 322.4 kW/hr of hot utilities and 1404.96 Btu/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 5.6. Figure 5.3 shows a design of resilient heat exchanger network for HDA process HIP1. In our case as shown in Fig. 5.3, the minimum temperature difference in the process-to-process-heat-exchangers ΔT_{min} is set to be 10 ^{o}C.

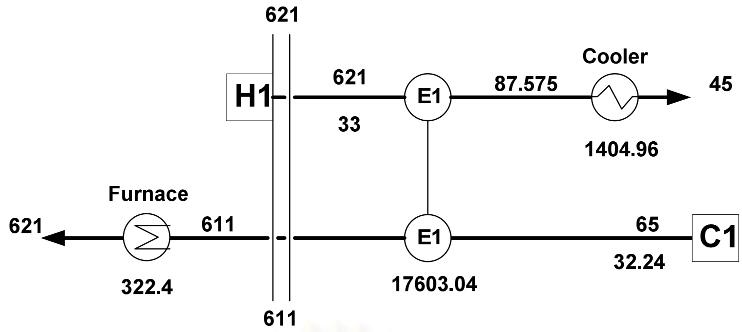


Figure 5.3 The resilient heat exchanger network, HIP1

Table 5.5 Process stream data for HIP1

Stream Name	Tin (°C)	Tout (°C)	W	Duty (kW)
H1	621	45	33	19008
C1	65	621	32.24	17925.44

Table 5.6 Problem table for HIP1

W		T		Δ T	sum W	Require	Interval	Cascade	sum Interval
H1	C1	hot	cold			(Heater)		(Cooler)	
0	0	631	621			Qh			
0	32.24	621	611	10	-32.24	322.4	-322.4	0	-322.4
33	32.24	75	65	546	0.76	0	414.96	414.96	92.56
33	0	45	35	30	33	414.96	990	1404.96	1082.56
								Qc	

Synthesis table for Cold End of HIP1

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	18678	33	611	45	330	0	select BC
C1	16958.24	32.24	75	601	322.4	322.4	select
b) state 2							
H1	1719.76	33	97.11	45	330	0	to Cooler
C1							match to H1

Synthesis table for Hot End of HIP1

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	0	33	621	621	330	330	select AH
C1	322.4	32.24	611	621	0	322.4	select
b) state 2							
H1							match to H1
C1	322.4	32.24	611	621	0	322.4	to Heater

5.2.2 Design of RHEN for heat-integrated plant (HIP2)

There are three streams in the networks. So we can find Pinch temperature using Problem table method as shown in Table 5.8. At the minimum heat load condition, the pinch temperature occurs at 621/611°C. The minimum utility requirements have been predicted 322.40 kW/hr of hot utilities and 1227.85 kW/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 5.8. Figure 5.4 shows a design of resilient heat exchanger network for HDA process HIP2. In our case as shown in Fig. 5.4, the minimum temperature difference in the process-to-process-heat-exchangers, is set to be 10 °C.

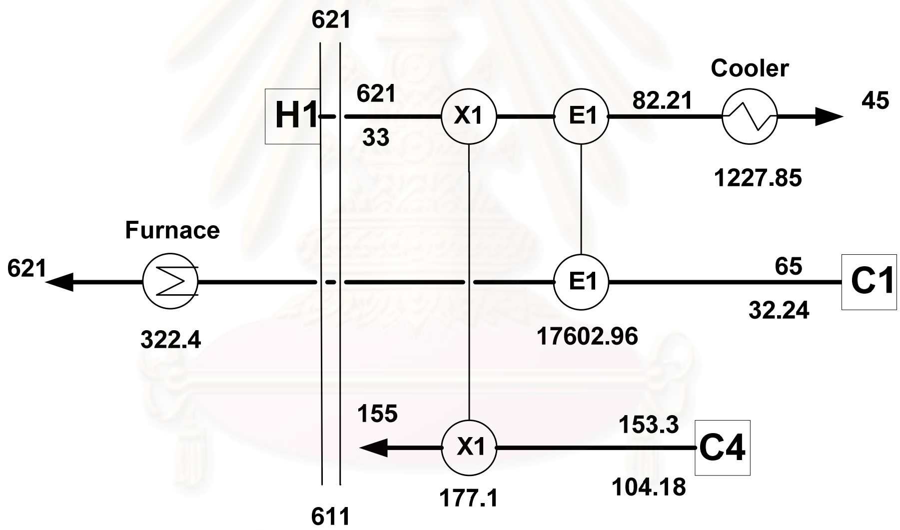


Figure 5.4 The resilient heat exchanger network, HIP2

Table 5.7 Process stream data for HIP2

Stream Name	Tin (°C)	Tout (°C)	W	Duty (kW)
H1: Reactor Product Stream (RPS)	621	45	33	19008.0
C1 : Reactor Feed Stream (RFS)	65	621	32.24	17925.4
C4 : Product Column Reboiler (PCR)	153.3	155	104.18	177.1

Table 5.8 Problem table for HIP2

W			T		ΔT	sum W	Require	Interval	Cascade	sum Interval
H1	C1	C3	hot	cold			(Heater)		(Cooler)	
0	0	0	631	621			Qh			
0	32.24	0	621	611	10	-32.24	322.40	-322.4	0.00	-322.4
33	32.24	0	165.00	155.00	456.00	0.76	0.00	346.56	346.56	24.16
33	32.24	104.18	163.30	153.30	1.70	-103.42	346.56	-175.81	170.746	-151.65
33	32.24	0	75	65	88.30	0.76	170.746	67.11	237.85	-84.55
33	0	0	45	35	30	33	237.85	990	1227.85	905.45
									Qc	

Synthesis table for Cold End of HIP2

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	18678	33	611	45	330	0	select BC
C1	17280.64	32.24	75	611	322.4	322.4	
C4	72.93	104.18	154.30	155	0	104.18	select
b) state 2							
H1	18500.89	33	605.63	45	330	0	to Cooler
C1	17280.64	32.24	75	611	322.4	322.4	select
C4							match to H1
c) state 3							
H1	897.85	33	72.21	45	330	0	to Cooler
C1							match to H1
C4							

Synthesis table for Hot End of HIP2

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	0	33	621	621	330	330	select C[H]
C1	322.40	32.24	611	621	0	322.4	
b) state 2							
H1							select AH
C1	322.40	32.24	611	621	0	322.4	select

5.2.3 Design of RHEN for heat-integrated plant (HIP3)

There are three streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 5.10. At the minimum heat load condition, the pinch temperature occurs at 350.59/290.59^{o}C. The minimum utility requirements have been predicted 399.09 kW/hr of hot utilities and 1161.45 kW/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 5.10. Figure 5.5 shows a design of resilient heat exchanger network for HDA process HIP3. In our case as shown in Fig. 5.5, the minimum temperature difference in the process-to-process-heat-exchangers, ΔT_{min} is set to be 10 ^oC.

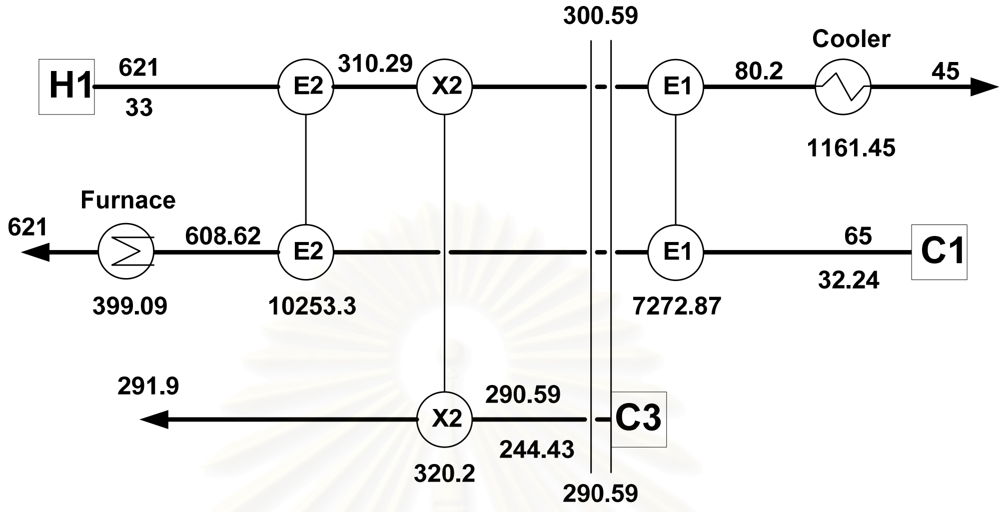


Figure 5.5 The resilient heat exchanger network, HIP3

Table 5.9 Process stream data for HIP3

Stream Name	Tin (^oC)	Tout (^oC)	W	Duty (kW)
H1: Reactor Product Stream (RPS)	621	45	33	19008.0
C1 : Reactor Feed Stream (RFS)	65	621	32.24	17925.4
C3 : Recycle Column Reboiler (RCR)	290.59	291.9	244.43	320.2

Table 5.10 Problem table for HIP3

W			T		ΔT	sum W	Require	Interval	Cascade	sum Interval
H1	C1	C3	hot	cold			(Heater)		(Cooler)	
0	0	0	631	621			Qh			
0	32.24	0	621	611	10	-32.24	399.09	-322.4	76.69	-322.4
33	32.24	0	301.90	291.90	319.10	0.76	76.69	242.52	319.21	-79.88
33	32.24	244.43	300.59	290.59	1.31	-243.67	319.21	-319.21	0	-399.09
33	32.24	0	75	65	225.59	0.76	0	171.45	171.45	-227.64
33	0	0	45	35	30	33	171.45	990	1161.45	762.36
									Qc	

Synthesis table for Cold End of HIP3

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	8104.47	33	290.59	45	330	0	select BC
C1	6950.62	32.24	75	290.59	322.4	322.4	select
b) state 2							
H1	831.45	33	70.20	45	330	0	to Cooler
C1							match to H1

Synthesis table for Hot End of HIP3

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	10573.53	33	611	290.59	330	330	select C[H]
C1	10287.78	32.24	301.90	621	0	322.4	
C3	75.77	244.43	291.59	291.90	0	244.43	select
b) state 2							
H1	10253.33	33	611	300.29	330	0	select AH
C1	10287.78	32.24	301.90	621	0	322.4	select
C3					0	85.57	to Heater
c) state 3							
H1							match to C1
C1	34.46	32.24	619.93	621	0	652.4	to Heater

5.2.4 Design of RHEN for heat-integrated plant (HIP4)

There are three streams in the network. So we can find Pinch temperature using Problem table method as shown in Table 5.12. At the minimum heat load condition, the pinch temperature occurs at 210.93/200.93^{o}C. The minimum utility requirements have been predicted 4488.65 kW/hr of hot utilities and 1093.31 kW/hr of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 5.12. Figure 5.6 shows a design of resilient heat exchanger network for HDA process HIP4. In our case as shown in Fig. 5.6, the minimum temperature difference in the process-to-process-heat-exchangers, ΔT_{min} is set to be 10^{o}C.

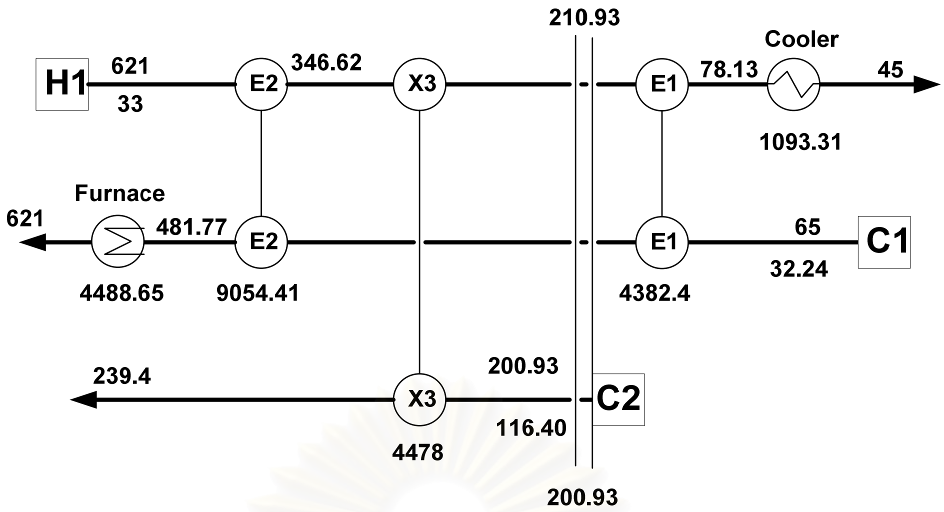


Figure 5.6 The resilient heat exchanger network, HIP4

Table 5.11 Process stream data for HIP4

Stream Name	Tin (°C)	Tout (°C)	W	Duty (kW)
H1: Reactor Product Stream (RPS)	621	45	33	19008.0
C1 : Reactor Feed Stream (RFS)	65	621	32.24	17925.4
C2 : Stabilizer Column Reboiler (SCR)	200.93	239.40	116.40	4477.9

Table 5.12 Problem table for HIP4

W			T		ΔT	sum W	Require	Interval	Cascade	sum Interval
H1	C1	C2	hot	cold			(Heater)		(Cooler)	
0	0	0	631	621			Qh			
0	32.24	0	621	611	10	-32.24	4488.65	-322.4	4166.25	-322.4
33	32.24	0	249.40	239.40	371.60	0.76	4166.25	282.42	4448.67	-39.98
33	32.24	116.4	210.93	200.93	38.47	-115.64	4448.67	-4448.67	0	-4488.65
33	32.24	0	75	65	135.93	0.76	0	103.31	103.31	-4385.35
33	0	0	45	35	30	33	103.31	990	1093.31	-3395.35
									Qc	

Synthesis table for Cold End of HIP4

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	5145.69	33	200.93	45	330	0	select BC
C1	4059.98	32.24	75	200.93	322.4	322.4	select
b) state 2							
H1	763.31	33	68.13	45	330	0	to Cooler
C1							match to H1

Synthesis table for Hot End of HIP4

stream	load	W	T1	T2	D1	D2	Action
a) state 1							
H1	13532.31	33	611	200.93	330	330	select C[H]
C1	13220.66	32.24	210.93	621	0	322.4	
C2	4361.51	116.4	201.93	239.40	0	116.4	select
b) state 2							
H1	9054.40	33	611	336.62	330	0	select AH
C1	13220.66	32.24	210.93	621	0	322.4	select
C2					0	213.6	to Heater
c) state 3							
H1							match to C1
C1	4166.25	32.24	491.77	621	0	652.4	to Heater

The various alternatives of heat exchanger network are designed for the heat-integrated plant (HIP) of HDA process, the energy saved from the HIP1 as in Table 5.13

Table 5.13 Energy integration for the heat-integrated plants (HIPs) of HDA process

Heat-integrated plants (HIPs) HDA process	Alternatives			
	HIP 1	HIP 2	HIP3	HIP4
Furnace	322.4	322.4	399.1	4488.7
Cooler	1405.0	1227.9	1161.5	1093.3
Stabilizer column reboiler	4477.9	4477.9	4477.9	0
Product column reboiler	177.1	0	177.1	177.1
Recycle column reboiler	320.2	320.2	0	320.2
Hot utilities usage, (kW)	5297.6	5120.5	5054.1	4986.0
Cold utilities usage, (kW)	1405.0	1227.9	1161.5	1093.3
Total Hot & Cold utilities (kW)	6702.6	6348.4	6215.5	6079.3
Energy savings from HIP, %	0	5.3	7.3	9.3

5.3 Design of Heat-Integration Plant Structure of HDA Process

Seven the heat exchanger networks (HENs) designs of the heat-integrated plant structure HDA process are proposed to save energy from the HIP1 and use to evaluate performance of control structures are designed both simply heat-integrated process and complex heat-integrated process.

In Figure 5.7 show the HIP1 of heat-integrated plant structure of HDA process with simply energy integration, we used a feed-effluent heat exchanger (FEHE) to reduce the amount of fuel burned in the furnace. The heat of reaction and the heat added in the furnace are therefore removed in the flooded condenser.

In HIP1 The simplest of these designs heat consumption by making the reactor preheater larger and the furnace smaller shown on Figure 5.7.

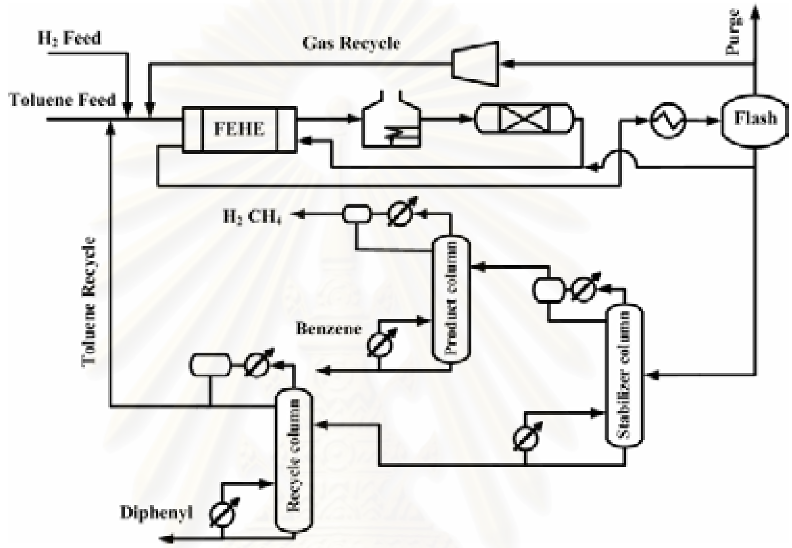


Figure 5.7 The heat-integrated plant structure of HDA process, HIP1

In HIP2 is the same as HIP1, except that product stream from reactor send to product column reboiler and next go on FEHE to preheat feed stream shown on Figure 5.8.

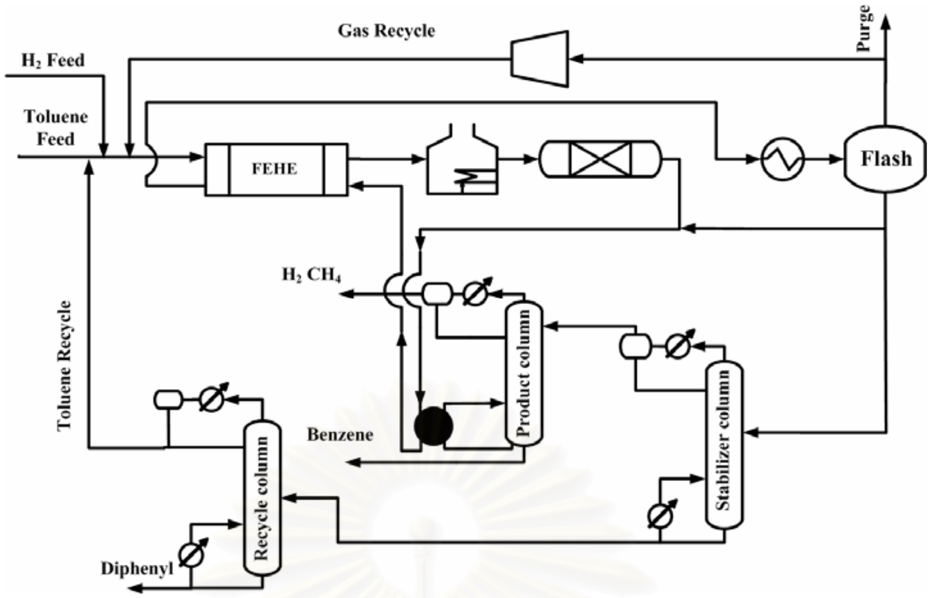


Figure 5.8 The heat-integrated plant structure of HDA process, HIP2

In HIP3 part of the heat in the reactor effluent stream is used to drive the recycle column reboiler shown on Figure 5.9.

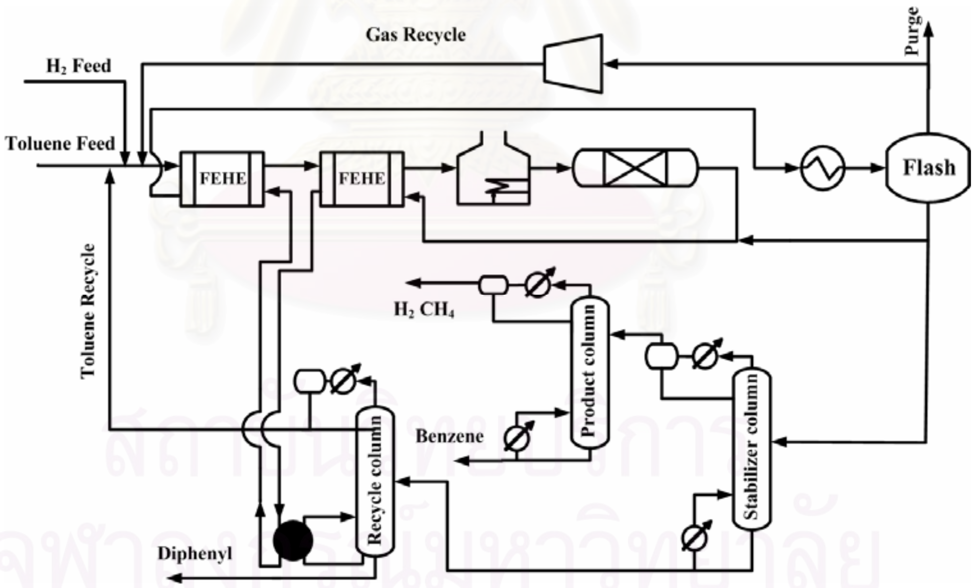


Figure 5.9 The heat-integrated plant structure of HDA process, HIP3

In HIP4 the reactor effluent is used to drive the stabilizer column reboiler shown on Figure 5.10.

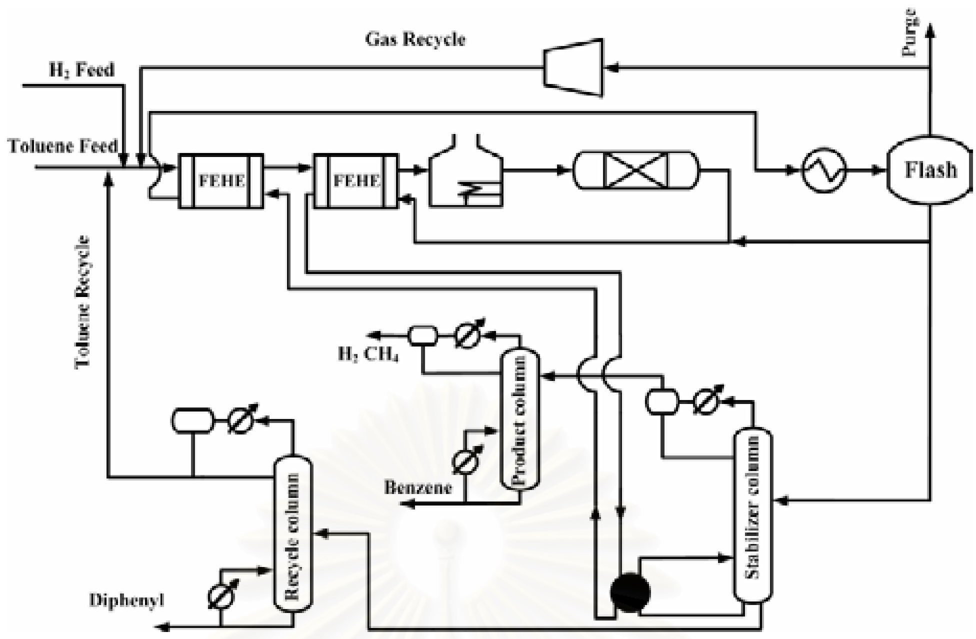


Figure 5.10 The heat-integrated plant structure of HDA process, HIP4

5.4 Steady State Model of Heat-Integrated Plant Structure of HDA Process

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 heat-integrated plant HDA process HIP 1, 2, 3 and 4. Figures 5.11 to 5.14 show the HYSYS flowsheets of the HDA process with heat-integration process schemes for HIP1, 2, 3 and 4, respectively. The data for the selected streams for these alternatives are not included in this chapter but listed in Appendix A. The data and specifications for the equipments are summarized in Appendix B. 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.

5.5 Comparisons of Energy from Steady State Simulation of Heat-Integrated Plant of HDA Process

From steady state simulation results by HYSYS, the energy saved from the HIP1 (base case) heat consumption as shown in Table 5.14. Energy saving for heat-integrated plant structures compared with typical HDA process show on Figure 5.11 and Table 5.15. In Our work, we are design heat-integrated process (HIP) of HDA process can save energy between 5.7 – 24.7% from typical of HDA process alternative1 (BC)

Table 5.14 Energy integration for heat-integrated plant of HDA process (Steady State Simulation)

Heat-integrated plant structure HDA process	Alternatives			
	HIP 1(BC)	HIP 2	HIP 3	HIP 4
Furnace	386.1	386.1	1890.8	3735.6
Cooler	2214.4	2022.2	476.0	995.6
Stabilizer Column Reboiler	4494.1	4492.0	4497.6	0
Product Column Reboiler	176.0	0	179.0	178.7
Recycle Column Reboiler	321.3	323.6	0	320.9
Total Reboiler	4991.4	4815.7	4676.6	499.6
Stabilizer Column Condenser	2664.2	2665.2	2665.1	2651.6
Product Column Condenser	1097.7	1096.5	1102.5	1099.9
Recycle Column Condenser	425.6	428.9	424.1	425.7
Total Condenser	4187.5	4190.7	4191.6	4177.1
Hot utilities usage, (kW)	5377.5	5201.7	6567.4	4235.2
Cold utilities usage, (kW)	6401.9	6212.9	4667.7	5172.7
Total Hot & Cold utilities (kW)	11779.4	11414.6	11235.1	9407.9
Energy savings %	0	3.1	4.6	20.1

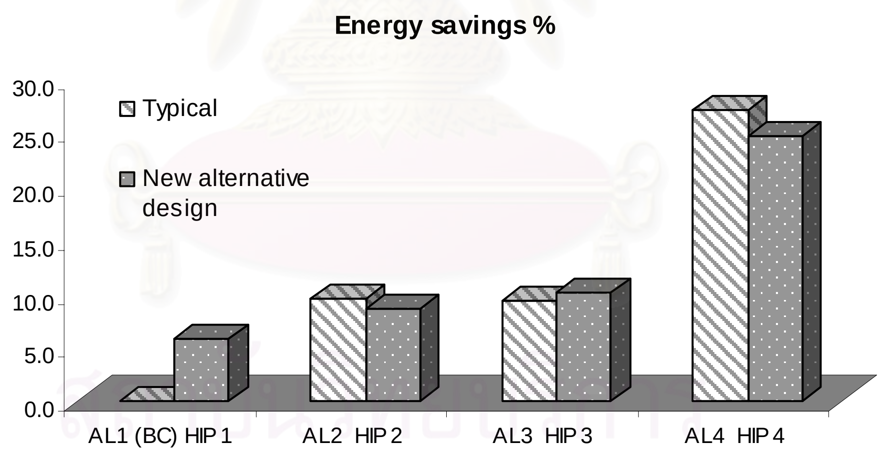


Figure 5.11 Energy saving for heat-integrated plant structures compared with typical HDA process (Steady State Simulation)

Table 5.15 Energy integration for heat-integrated plant structures compared with typical HDA process (Steady State Simulation)

Description	Alternatives							
	AL1 (BC)	AL2	AL3	AL4	HIP 1	HIP 2	HIP 3	HIP 4
Furnace	385.9	384.7	1721.7	2126.1	386.1	386.1	1890.8	3735.6
Cooler	2212.0	2307.1	2074.6	1157.2	2214.4	2022.2	476.0	995.6
Stabilizer Column Reboiler	1273.0	1273.1	0	1280.1	4494.1	4492.0	4497.6	0
Product Column Reboiler	3461.8	0	0	0	176.0	0	179.0	178.7
Recycle Column Reboiler	481.5	565.1	563.7	563.7	321.3	323.6	0	320.9
Heater Product Reboiler	0	2772.9	2792.1	2772.9	0	0	0	0
Total Reboiler	5216.3	4630.6	3356.0	1843.8	4991.4	4815.7	4676.6	499.6
Stabilizer Column Condenser	176.7	176.9	373.8	182.9	2664.2	2665.2	2665.1	2651.6
Product Column Condenser	4061.3	3797.1	3809.6	3805.8	1097.7	1096.5	1102.5	1099.9
Recycle Column Condenser	437.4	0	0	0	425.6	428.9	424.1	425.7
Total Condenser	4675.4	3974.0	4183.4	3988.7	4187.5	4190.7	4191.6	4177.1
Hot utilities usage, (kW)	5602.1	5015.2	5077.7	3969.9	5377.5	5201.7	6567.4	4235.2
Cold utilities usage, (kW)	6887.4	6281.0	6258.0	5145.9	6401.9	6212.9	4667.7	5172.7
Total Hot & Cold utilities (kW)	12489.5	11296.3	11335.7	9115.8	11779.4	11414.6	11235.1	9407.9
Energy savings %	**0**	**9.6**	**9.2**	**27.0**	**5.7**	**8.6**	**10.0**	**24.7**

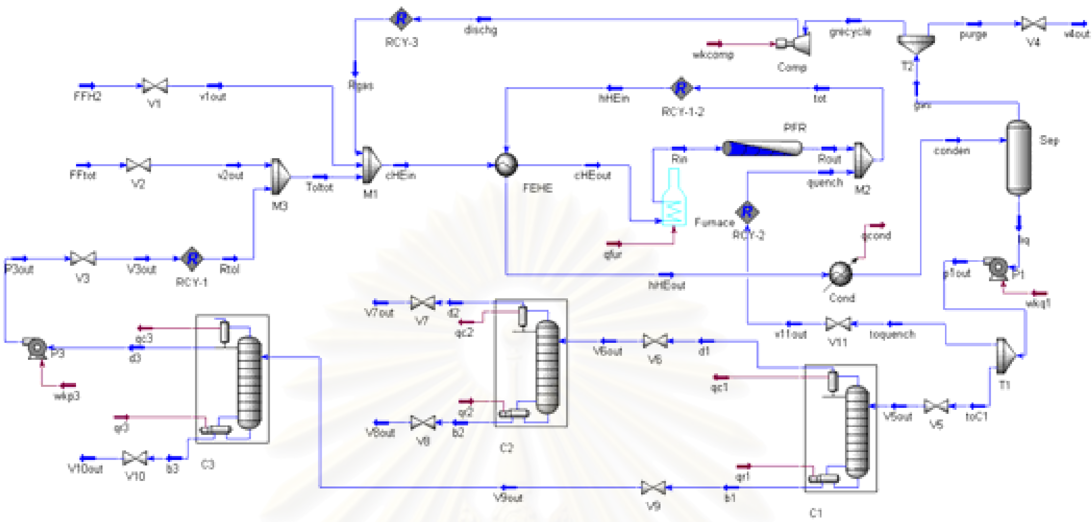


Figure 5.12 HYSYS Flowsheet of the Steady State Model for heat-integrated plant structure of HDA Process, HIP1

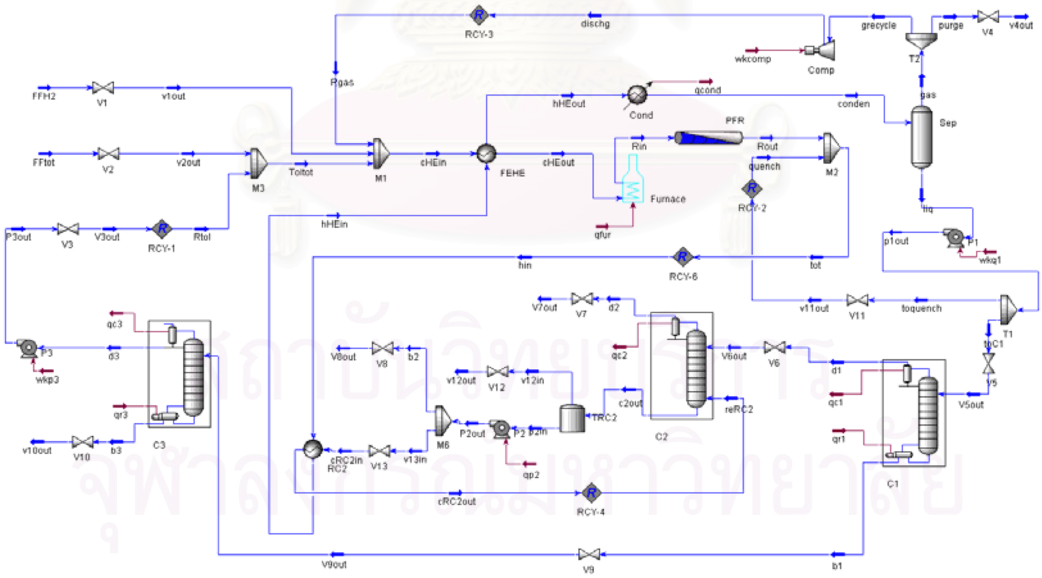


Figure 5.13 HYSYS Flowsheet of the Steady State Model for heat-integrated plant structure of HDA Process, HIP2

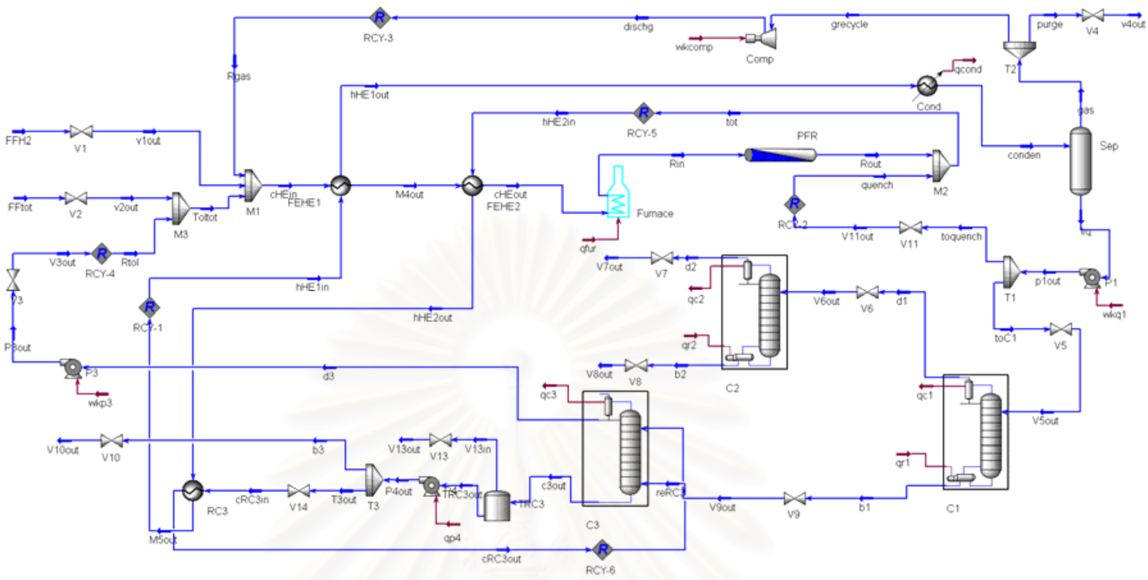


Figure 5.14 HYSYS Flowsheet of the Steady State Model for heat-integrated plant structure of HDA Process, HIP3

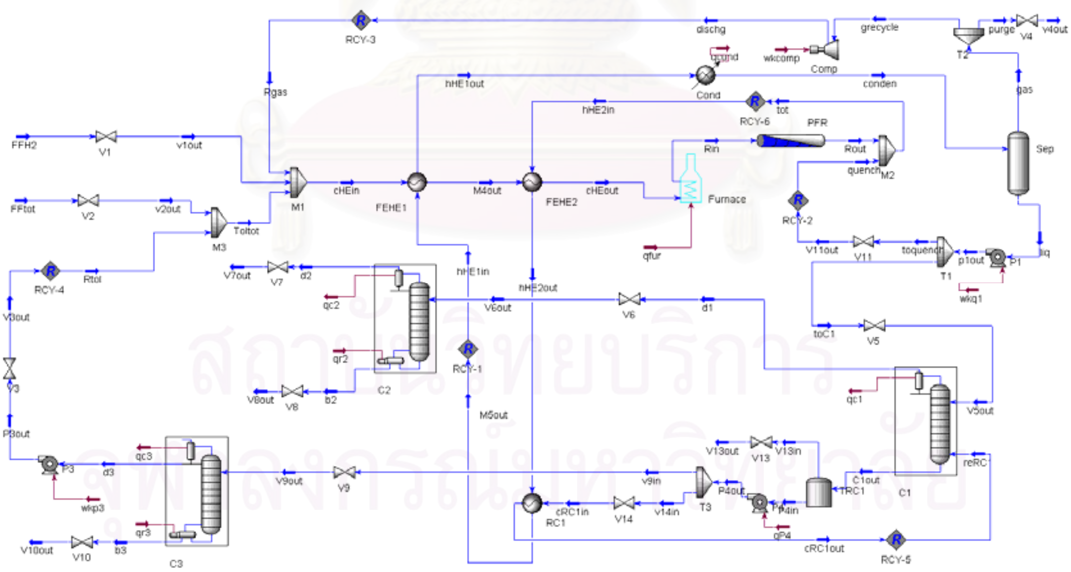


Figure 5.15 HYSYS Flowsheet of the Steady State Model for heat-integrated plant structure of HDA Process, HIP4

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 energy integrated process. Moreover, the three new designed control structures are also compared between typical and heat-integrated plant (HIP) of HDA process based on rigorous dynamic simulation by using the commercial software HYSYS.

6.1 Plantwide Control Strategies

The plantwide control structures in the typical and heat-integrated plant (HIP) of HDA process are designed based on the heuristic design procedure given by Luyben et al. (1999) and discussed below.

Step1. Establish Control Objectives

For this process, the essential is to produce pure benzene while minimizing yield losses of hydrogen and diphenyl. The reactor feed ratio of hydrogen to aromatics must be greater than 5:1. The reactor effluent gas must be quenched to 1150^{o}F

Step2. Determine Control Degree of Freedom

There are 23 control degrees of freedom. They include; two fresh feed valves for hydrogen and toluene, purge valve, separator base and overhead valves, cooler cooling water valve, liquid quench valve, furnace fuel valve, stabilizer column steam, bottoms, reflux, cooling water, and vapor product valves; product column steam, bottoms, reflux, distillate, and cooling water valves; and recycle column steam, bottoms, reflux, distillate, and cooling water valves.

Step3. Establish Energy management system

The product benzene is produced from the exothermic reaction between hydrogen and toluene at 1158^{o}F. 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 for preventing the benzene yield decreases from the side reaction. In the reference control structure, quenched stream is used for control temperature at the design value and for saving cost from the cooling utility. However, this method makes the path of disturbance propagation to the separation section, so the product purity control must be tighter because of component inventories changing. The alternative way is using of the heuristic laws; Montree (2000) introduces about the energy management that “Decreasing the effect of heat integration in the process can be done by remove the energy as much as possible”. Therefore, the cooling utility should be used for controlling the reactor exit temperature and preventing the disturbance propagation to the separation section as the second control structure. Another energy control loop is using of the cooling utility for removing excess heat from the heat exchanger to reach the optimal temperature in the separator.

Step4. Set Production Rate

There are not constrained to set production either by supply or demand, then the production rate can be set by benzene production. Considering of the kinetics equation is found that the three variables alter the reaction rate; pressure, temperature and toluene concentration which is the limiting agent.

• Pressure control of the compressor operates at maximum capacity for yield purposes.

• Reactor inlet temperature is controlled by specify the reactant fresh feed rate and reactant composition into the reactor. The reactor inlet temperature is constrained below 1300^{o}F for preventing the cracking reaction that produces undesired byproduct.

• Toluene inventory can be controlled in liquid level at the top of recycle column is measured to change toluene feed flow.

Step5. Control Product Quality and Handle Safety, Operational, and Environmental

Constraints Benzene purity must be maintained at 99.97 percent for this research. Any methane that leaves in the bottoms of the stabilizer column contaminates the benzene product. The separation in the stabilizer column prevents the problem by using a temperature to set column stream rate (boilup). Toluene in the overhead of the product column also affects benzene quality. Benzene purity can be controlled by manipulating the column steam rate (boilup) to maintain temperature in the column.

Step6. Control Inventories and Fix a Flow in Every Recycle Loop

In most processes a flow control should be present in all recycle loops. This is a simple and effective way to prevent potentially large changes in recycle flows, while the process is perturbed by small disturbance. We call this high sensitivity of the recycle flowrates to small disturbances the “snowball effect”. There are two recycle control loops.

Four pressures and seven liquid levels must be controlled in this process. For the pressures, there are in the gas loop and in the three distillation columns. In the gas loop, the separator overhead valve is opened and run the compressor at maximum gas recycle rate to improve yield so the gas loop control is related to the purge stream and fresh hydrogen feed flow. In the stabilizer column, vapor product flow is used to control pressure. In the product and recycle columns, pressure control can be achieved by manipulating cooling water flow to regulate overhead condensation rate.

For liquid loops, there are a 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. The stabilizer column overhead level is controlled with cooling water flow and base level is controlled with bottoms flow. In the product column, distillate flow controls overhead receiver level and bottoms flow controls base level. In the recycle column, control structure use the fresh toluene feed flow to control level. The base level of recycle column is controlled by manipulating the column steam flow because it has much larger effect than bottoms flow.

Step7. Check Component Balances

Component balances control loops consists of:

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

• Diphenyl is removed in the bottoms stream from the recycle column, where steam flow controls base level.

• The inventory of benzene is accounted for by temperature and overhead receiver level control in the product column.

• Toluene inventory is accounted for by level control in the recycle column overhead receiver.

• Gas loop pressure control accounts for hydrogen inventory.

Step8. Control Individual Unit Operations

The rest degrees of freedom are assigned for control loops within individual units. These include:

• Cooling water flow to the cooler controls process temperature to the separator.

• Refluxes to the stabilizer, product, and recycle columns are flow controlled.

Step9. 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).

6.2 Energy Management of Heat-Integrated Process

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

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

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

6.2.1 Design of Heat Pathways and HEN control configuration for AL 1 and AL2

The design of the heat pathways for AL1and AL2 shown in Figure 6.1 shifts the positive and negative disturbance loads of C1 to cooler. Thus, the negative disturbance load of a cold stream will result in decrease of the cooler duty which is good. The negative disturbance load will result in increase of the cooler duty which is ruled by ΔT_{min} constraint. The negative or positive disturbance load of H1 is directed to cooler; the cooler duty of corresponding column is increased or decreased accordingly.

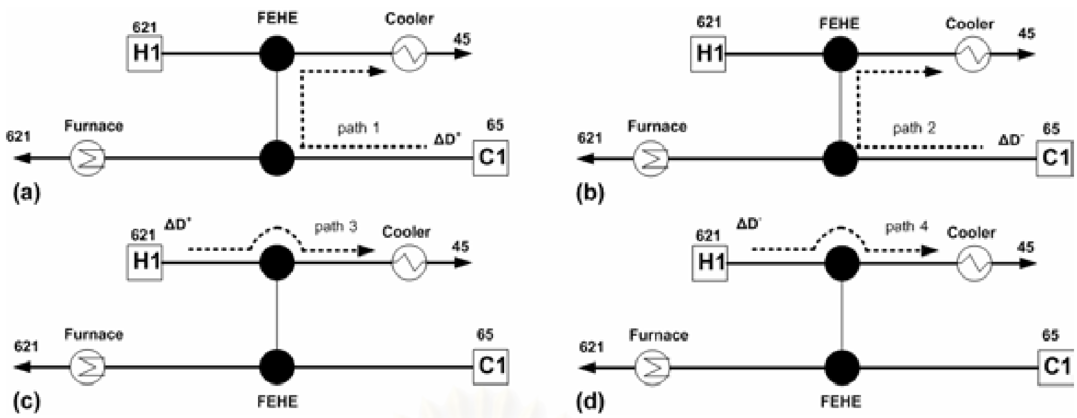


Figure 6.1 Heat pathways through AL1 and AL2, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_1 to the cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_1 to the cooler, (c) path 3 is used to shift the positive disturbance load of the hot stream H_1 to cooler, (d) path 4 is used to shift the negative disturbance load of the hot stream H_1 to cooler

From designed the heat pathways for AL1and AL2, we can design the control configurations as show in Figure 6.2.

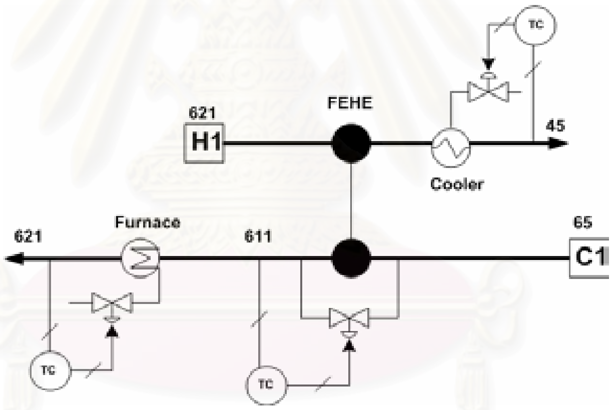


Figure 6.2 Control configurations of AL1 and AL2

6.2.2 Design of Heat Pathways and HEN control configuration for AL2 and AL3

The design of the heat pathways for AL2 and AL3 shown in Figure 6.3 shifts the positive and negative disturbance loads of C2 to heater. Thus, the positive disturbance load of a cold stream will result in decrease of the heater duty which is good. The negative disturbance load will result in increase of the heater duty which is ruled by ΔT_{min} constraint. The negative or positive disturbance load of H2 is directed to the heater; the heater duty of corresponding column is increased or decreased accordingly.

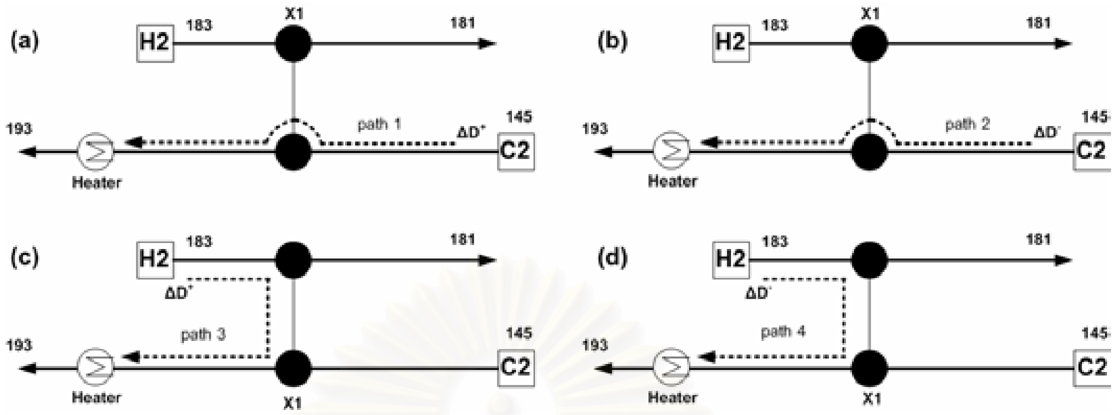


Figure 6.3 Heat pathways through AL2 and AL3, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_2 to heater, (b) path 2 is used to shift the negative disturbance load of the cold stream C_2 to heater, (c) path 3 is used to shift the positive disturbance load of the hot stream H_2 to heater, (d) path 4 is used to shift the negative disturbance load of the hot stream H_2 to heater

These control systems involve one manipulated variable and one controlled variable and work as follows: Figure 6.4

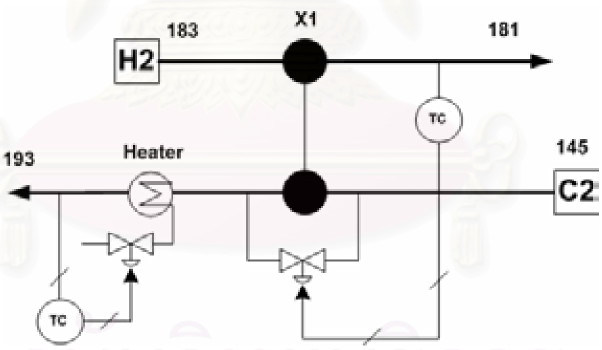


Figure 6.4 Control configurations of AL2 and AL3

6.2.3 Design of Heat Pathways and HEN control configuration for AL3

The design of the heat pathways for AL3 shown in Figure 6.5 shifts the positive and negative disturbance loads of H1 to furnace. Thus, the negative disturbance load of a cold stream will result in decrease of the furnace duty which is good. The negative disturbance load will result in increase of the furnace duty which is ruled by ΔT_{min} constraint. Both negative and positive disturbance loads of C1 and C3 are shifted to the cooler.

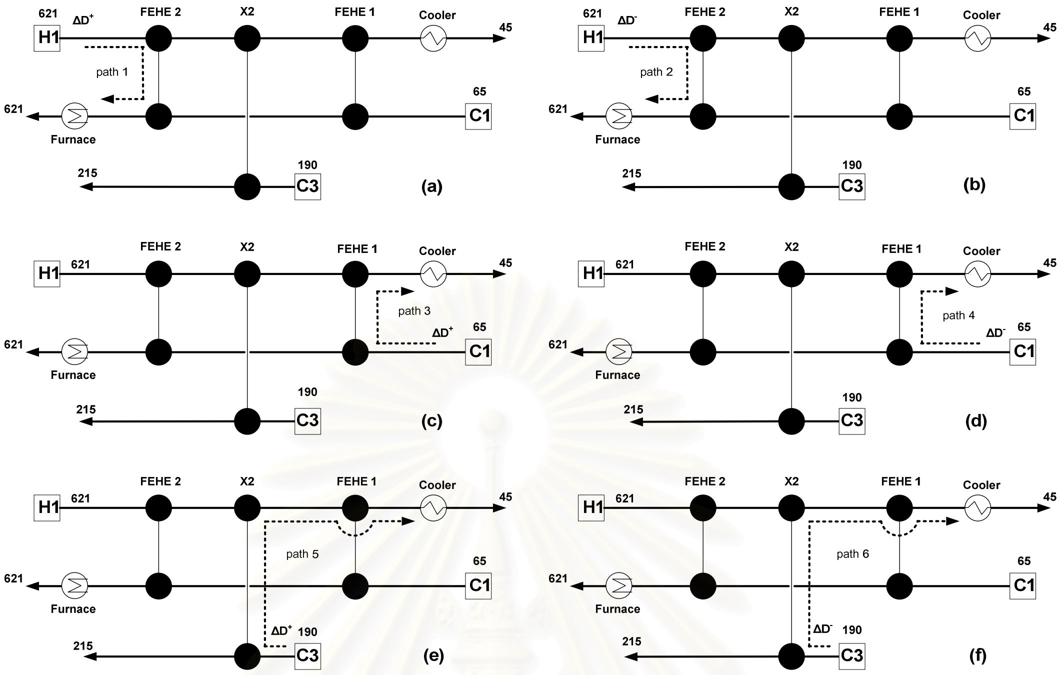


Figure 6.5 Heat pathways through AL3, where: (a) path 1 is used to shift the positive disturbance load of the hot stream H_1 to the furnace, (b) path 2 is used to shift the negative disturbance load of the hot stream H_1 to the furnace , (c) path 3 is used to shift the positive disturbance load of the cold stream C_1 to cooler, (d) path 4 is used to shift the negative disturbance load of the cold stream C_1 to cooler, (e) path 5 is used to shift the positive disturbance load of the cold stream C_3 to cooler and (f) path 6 is used to shift the negative disturbance load of the cold stream C_3 to cooler

From designed the heat pathways for AL3, we can design the control configurations as show in Figure 6.6.

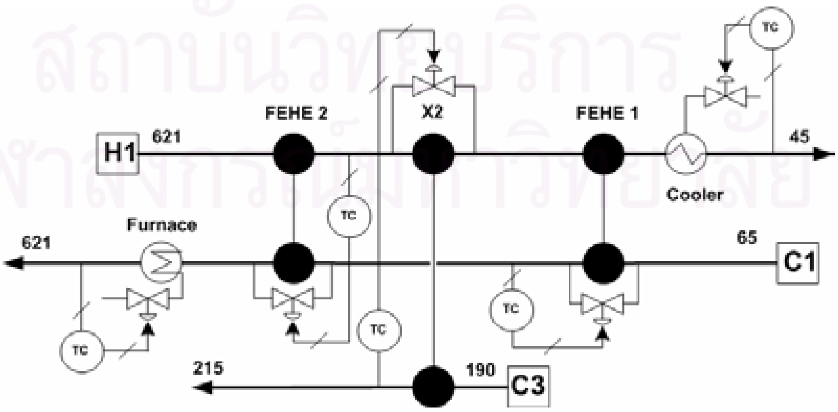


Figure 6.6 Control configurations of AL3

6.2.4 Design of Heat Pathways and HEN control configuration for AL4

The design of the heat pathways for AL4 shown in Figure 6.7 shifts the positive and negative disturbance loads of H1 to furnace and H2 to cooler. Thus, the negative disturbance load of a cold stream will result in decrease of the furnace duty

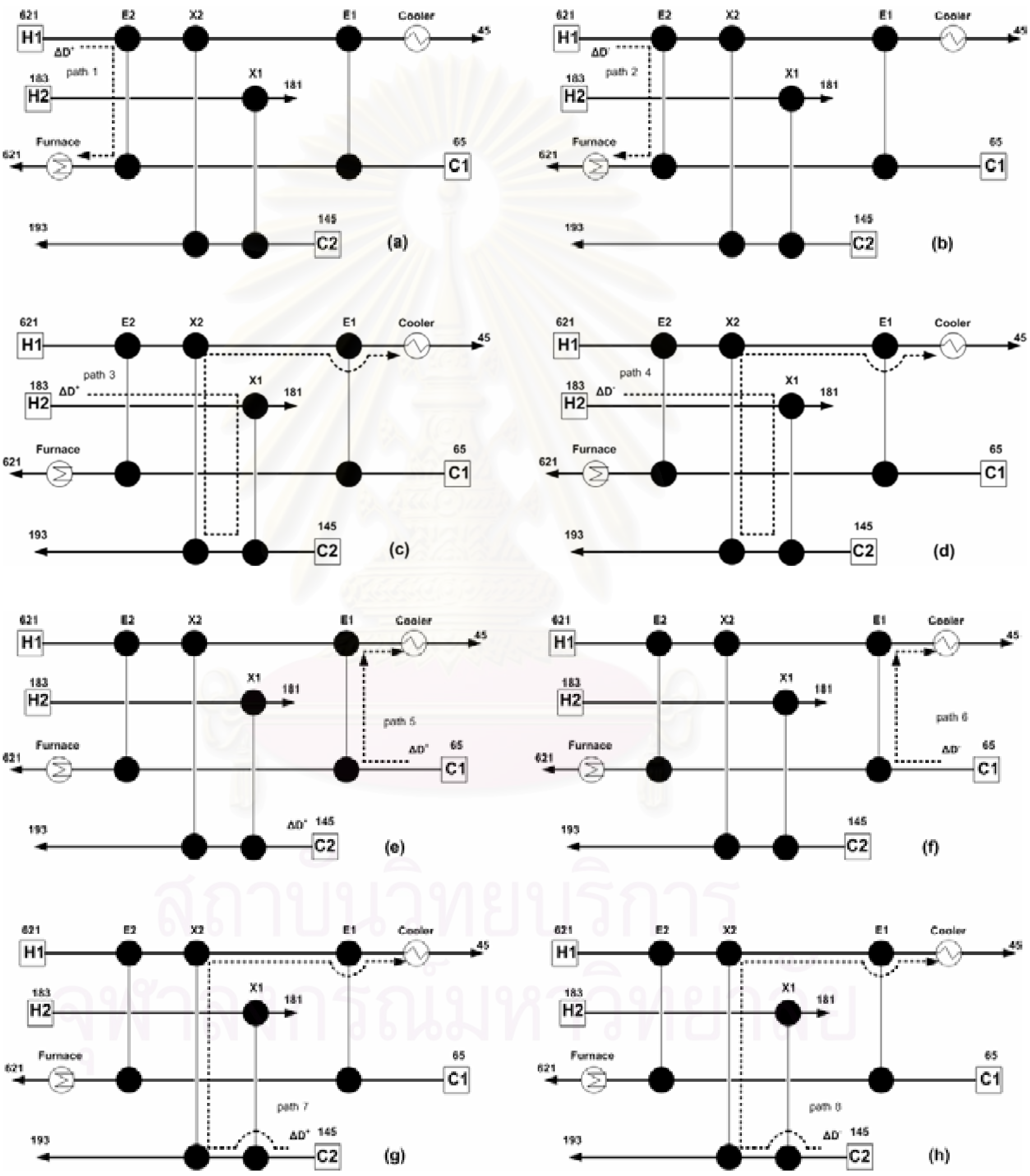


Figure 6.7 Heat pathways through AL4, where: (a) path 1 is used to shift the positive disturbance load of the hot stream H_1 to the furnace, (b) path 2 is used to shift the negative disturbance load of the hot stream H_1 to the furnace , (c) path 3 is used to shift the positive disturbance load of the hot stream H_2 to cooler, (d) path 4 is used to shift the negative disturbance load of the hot stream H_2 to cooler, (e) path 5 is used to

shift the positive disturbance load of the cold stream C_1 to cooler and (f) path 6 is used to shift the negative disturbance load of the cold stream C_1 to cooler, (g) path 7 is used to shift the positive disturbance load of the cold stream C_2 to cooler and (h) path 8 is used to shift the negative disturbance load of the cold stream C_2 to cooler

which is good. The negative disturbance load will result in increase of the furnace duty which is ruled by ΔT_{min} constraint. Both negative and positive disturbance loads of C1 and C2 are shifted to the cooler.

From designed the heat pathways for AL4, we can design the control configurations as show in Figure 6.8.

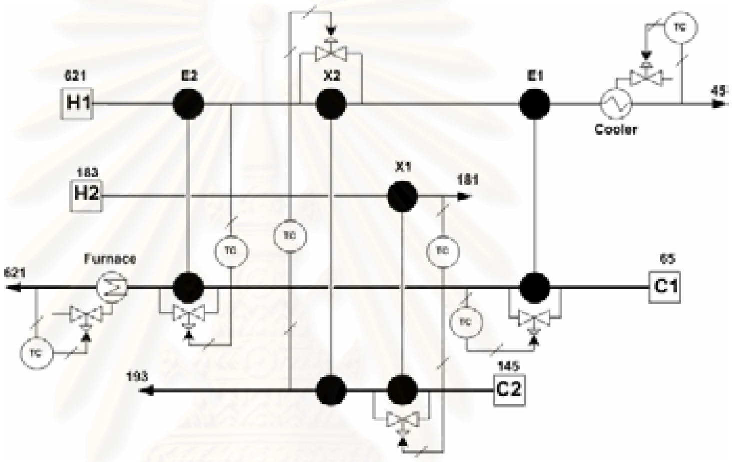


Figure 6.8 Control configurations of AL4

6.2.5 Design of Heat Pathways and HEN control configuration for HIP1

The design of the heat pathways for HIP1 shown in Figure 6.9 shifts the positive and negative disturbance loads of C1 to cooler. Thus, the negative disturbance load of a cold stream will result in decrease of the cooler duty which is good. The negative disturbance load will result in increase of the cooler duty which is ruled by ΔT_{min} constraint. The negative or positive disturbance load of H1 is directed to cooler; the cooler duty of corresponding column is increased or decreased accordingly.

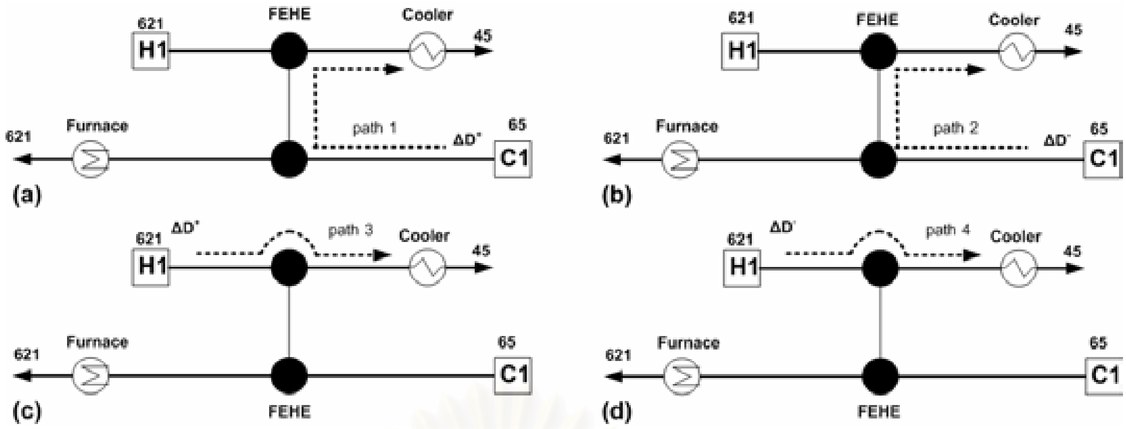


Figure 6.9 Heat pathways through HIP1, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_1 to the cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_1 to the cooler, (c) path 3 is used to shift the positive disturbance load of the hot stream H_1 to cooler, (d) path 4 is used to shift the negative disturbance load of the hot stream H_1 to cooler

From designed the heat pathways for HIP1, we can design the control configurations as show in Figure 6.10.

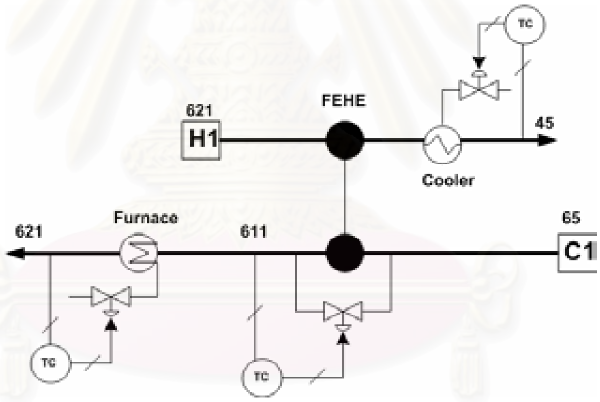


Figure 6.10 Control configurations of HIP1

6.2.6 Design of Heat Pathways and HEN control configuration for HIP2

The design of the heat pathways for HIP2 shown in Figure 6.11 shifts the positive and negative disturbance loads of H1 to cooler. Thus, the negative disturbance load of a cold stream will result in decrease of the furnace duty which is good. The negative disturbance load will result in increase of the furnace duty which is ruled by ΔT_{min} constraint. Both negative and positive disturbance loads of C1 and C4 are shifted to the cooler.

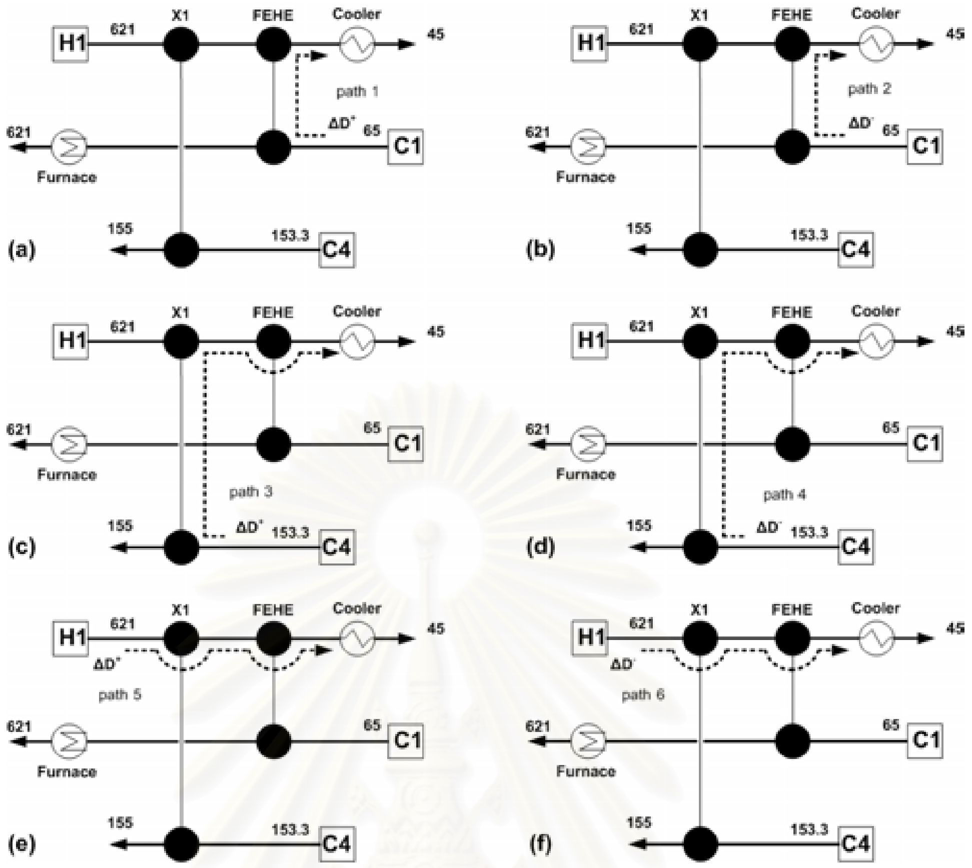


Figure 6.11 Heat pathways through HIP2, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_1 to cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_1 to the cooler, (c) path 3 is used to shift the positive disturbance load of the cold stream C_4 to cooler, (d) path 4 is used to shift the negative disturbance load of the cold stream C_4 to the cooler, (e) path 5 is used to shift the positive disturbance load of the hot stream H_1 to cooler, (f) path 6 is used to shift the negative disturbance load of the hot stream H_1 to cooler

From designed the heat pathways for HIP2, we can design the control configurations as show in Figure 6.12.

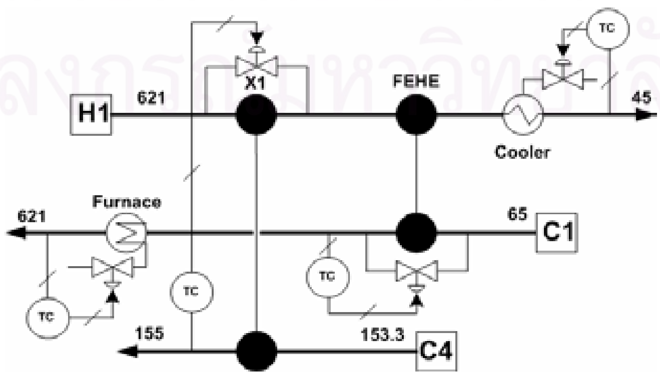


Figure 6.12 Control configurations of HIP2

6.2.7 Design of Heat Pathways and HEN control configuration for HIP3

The design of the heat pathways for HIP3 shown in Figure 6.13 shifts the positive and negative disturbance loads of H1 to the furnace. Thus, the negative disturbance load of a cold stream will result in decrease of the furnace duty which is good. The negative disturbance load will result in increase of the furnace duty which is ruled by ΔT_{min} constraint. Both negative and positive disturbance loads of C1 and C3 are shifted to the cooler.

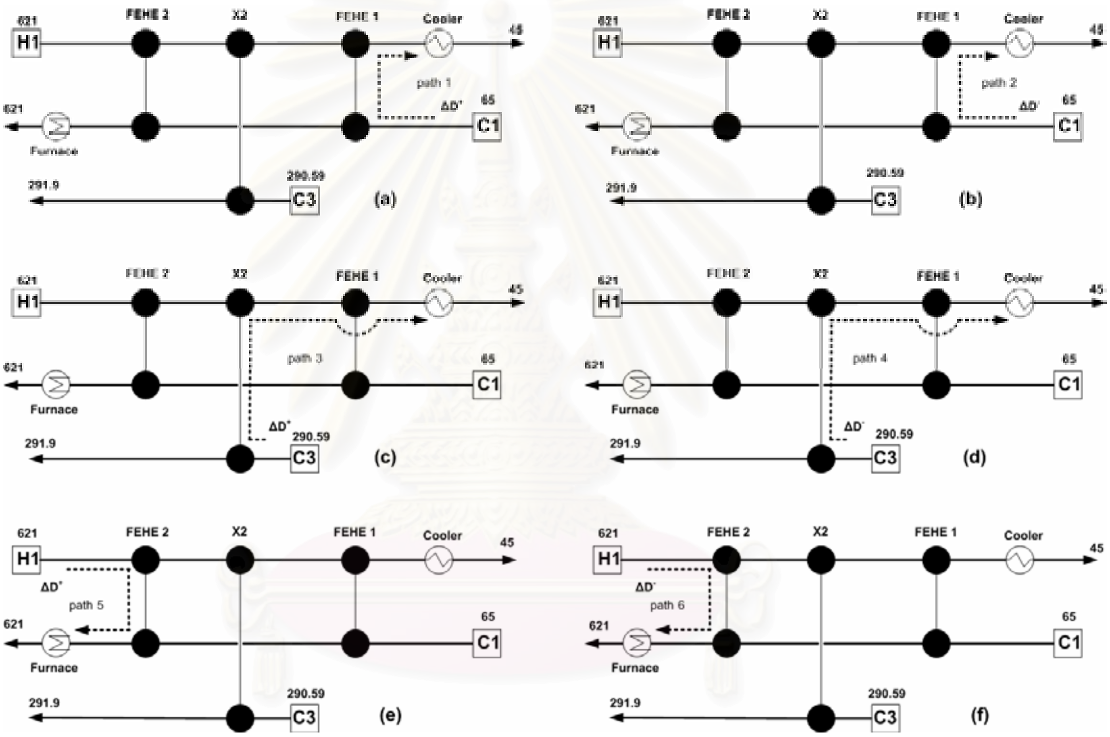


Figure 6.13 Heat pathways through HIP3, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_1 to cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_1 to the cooler, (c) path 3 is used to shift the positive disturbance load of the cold stream C_3 to cooler, (d) path 4 is used to shift the negative disturbance load of the cold stream C_3 to cooler, (e) path 5 is used to shift the positive disturbance load of the hot stream H_1 to the furnace, (f) path 6 is used to shift the negative disturbance load of the hot stream H_1 to the furnace

From designed the heat pathways for HIP3, we can design the control configurations as show in Figure 6.14.

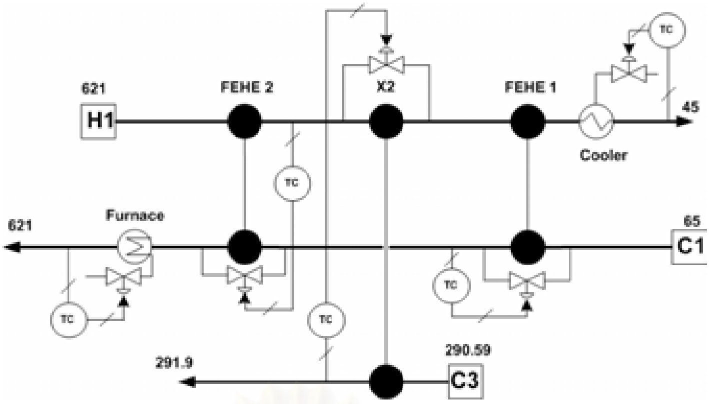


Figure 6.14 Control configurations of HIP3

6.2.8 Design of Heat Pathways and HEN control configuration for HIP4

The design of the heat pathways for HIP4 shown in Figure 6.15 shifts the positive and negative disturbance loads of H1 to the furnace. Thus, the negative

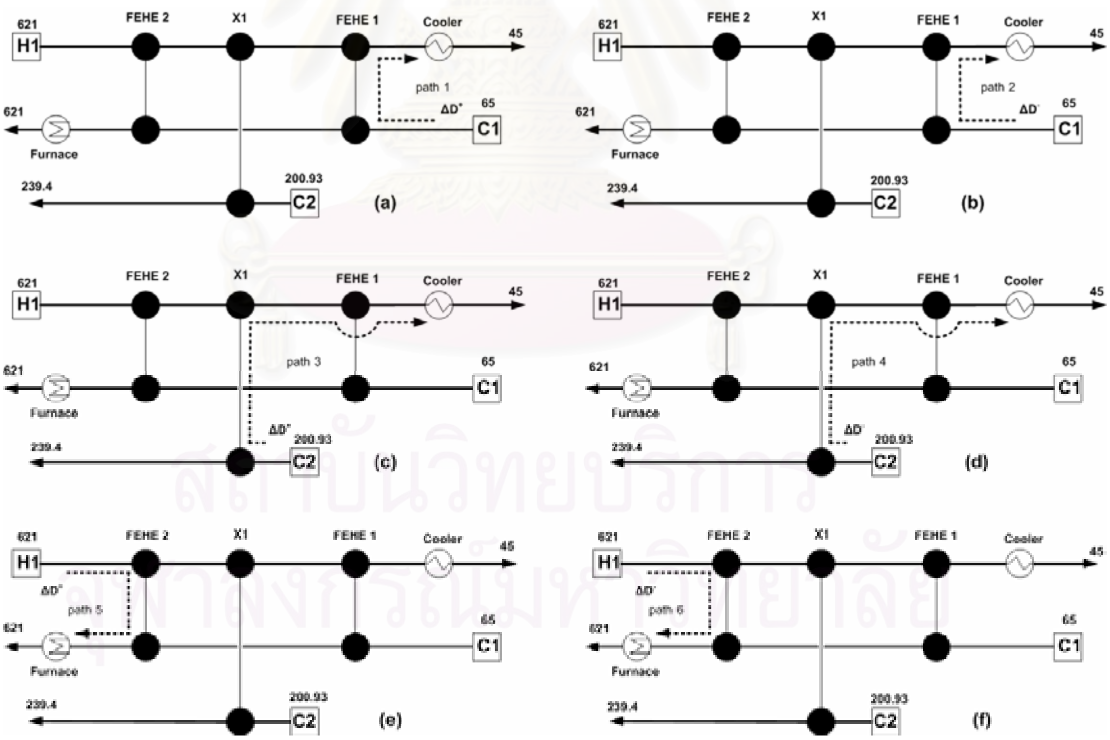


Figure 6.15 Heat pathways through HIP4, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C_1 to cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C_1 to the cooler, (c) path 3 is used to shift the positive disturbance load of the cold stream C_2 to cooler, (d) path 4 is used to shift the negative disturbance load of the cold stream C_2 to cooler, (e) path 5 is used to shift

the positive disturbance load of the hot stream H_1 to the furnace, (f) path 6 is used to shift the negative disturbance load of the hot stream H_1 to the furnace

disturbance load of a cold stream will result in decrease of the furnace duty which is good. The negative disturbance load will result in increase of the furnace duty which is ruled by ΔT_{min} constraint. Both negative and positive disturbance loads of C1 and C2 are shifted to the cooler.

From designed the heat pathways for HIP4, we can design the control configurations as show in Figure 6.16.

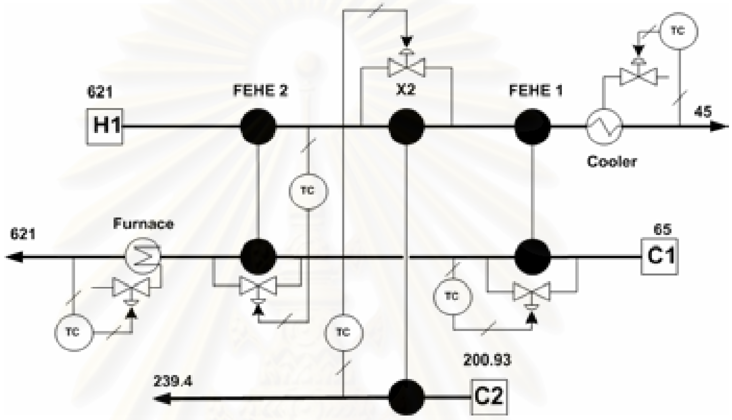


Figure 6.16 Control configurations of HIP4

6.3 Design of Plantwide Control Structure

In this work, the HDA process is designed by considering several control objectives (Luyben et al., 1999); they are: achieving a specified production rates of essential pure benzene (99.97 %-mole); achieving a ratio of hydrogen to aromatic greater than 5:1 in the reactor feed; quenching reactor effluent to a temperature of 621 oC to prevent coking and by-product formation in the heat exchanger. The plantwide control structures in the heat-integrated processes are designed based on the heuristic design procedure given by Luyben et al. (1999).

The major loops are the same as those used in Luyben et al. (1999), but we have designed two new loops for FEHE and three new loops for the three columns. In the literature (e.g. Luyben et al., 1999), a bypass control and an auxiliary utility exchanger are used for control in the heat integration system. Only bypass control is used in the current study; the reason is that in our case the heat load of the heating stream is greater than the required heat load of reboiler.

For all the heat integration units, the bypass streams are designed to control the outlet temperatures of FEHEs and the tray temperatures in the three columns. The bypass stream should be about 5 to 10 percent of the total flow to be able to handle disturbances (Jones and Wilson, 1997). In normal operation, a control valve should operate with an opening between 20 to 80 percent (Jones and Wilson, 1997). In our study, the bypass valves in the process-to-process-heat-exchangers are designed with the valve opening of 50%, i.e. this translates into the bypass flow rates of about 6% of the total flow. In practice we have to overdesign the process-to-process-heat-exchanger, in order to be able to handle the disturbances. In this work, it is not our intention to study the best overdesign policy. The oversize of the heat exchanger is related to the estimated maximum size of disturbance loads of both the cold and hot streams. The size of disturbance in this study is about 5 to 10% according to Luyben's recommendations.

So, there are 72 alternatives of the heat exchanger networks with control structures for HDA process include with 3 control structures, AL 1 to 4 with 3 control structures and HIP 1 to 4 with 3 control structures as show in Figures 6.17 to 6.40.

6.3.1 Design of Plantwide Control for the Typical of HDA Process Alternatives (AL1, AL2, AL3 and AL4)

The three new control structures are designed for the Typical of HDA process alternatives 1 to 4 that is propose in this research.

6.3.1.1 Control Structure 1 (CS1) for the Typical of HDA Process Alternative 1

This control structure is shown in Figure 6.17 and the controller parameters are given in table 6.1. This control structure, the all bypass of feed effluent heat exchangers (FEHE) is on cold side. Figure 6.17 shows the plantwide control structure of HDA process alternative 1. The major loops in HDA process alternative 1 are the same as those used in Luyben et al. (1999), except for the outlet temperature control in FEHE and the tray temperature control in the recycle column (C3) the furnace inlet temperature is controlled by manipulating the valve on the bypass line Based on the heat pathway heuristics for plantwide control

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. 6.17) 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 employed for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.3.1.2 Control Structure 2 (CS2) for the Typical of HDA Process Alternative 1

This control structure is shown in Figure 6.18 and the controller parameter is given in table 6.1. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass of FEHE will be on hot side.

6.3.1.3 Control Structure 3 (CS3) for the Typical of HDA Process Alternative 1

This control structure is shown in Figure 6.19 and the controller parameter is given in table 6.1. 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.

Table 6.1 Control Structure and Controller Parameter for the typical of HDA Process Alternative 1: Control Structure 1, Control Structure 2 and Control Structure 3

Controller	Controlled variable	Manipulated variable	Type	Kc	Ti	Td
Reaction section						
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 (V3)	PID	0.458	0.409	0.091
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.385	0.405	0.090
TCS	separator temperature	cooler duty (qcooler)	PID	0.489	0.198	0.044
TCVBP1*	FEHE cold outlet temperature	FEHE bypass cold stream valve (VBP1)	PID	1.237	0.349	0.078
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
Separation sections						
Stabilizer column						
PC1	column C1 pressure	column C1 gas valve (V6)	PI	2	10	-
TC1	column C1 tray-6 temperature	column C1 reboiler duty (qr1)	PI	2	10	-
LC11	column C1 base level	column C2 feed valve (V7)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	2	-	-
Product column						
PC2	column C2 pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	heat reboiler (qAR2) duty	PID	3.952	4.094	0.910
TC2-2**	column C2 tray-17 temperature	column C2 reflux flow rate	PI	1.8839	107.31	23.847
LC21	column C2 base level	column C3 feed valve (V9)	P	2	-	-
LC22	column C2 reflux drum level	column C2 product valve level (V8)	P	2	-	-
Recycle column						
PC3	column C3 pressure	CR bypass valve (VBP4)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 by-product valve (V8)	PID	0.511	14.778	3.284
LC31	column C3 base level	column C3 reboiler duty (qr3)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V11)	P	2	-	-

** TCVBP 1 or 3 manipulated variable: FEHE 1,2 bypass hot stream valve (VBP1 or 3) used in only CS2.*

*** TC2-2 controller used in only CS3.*

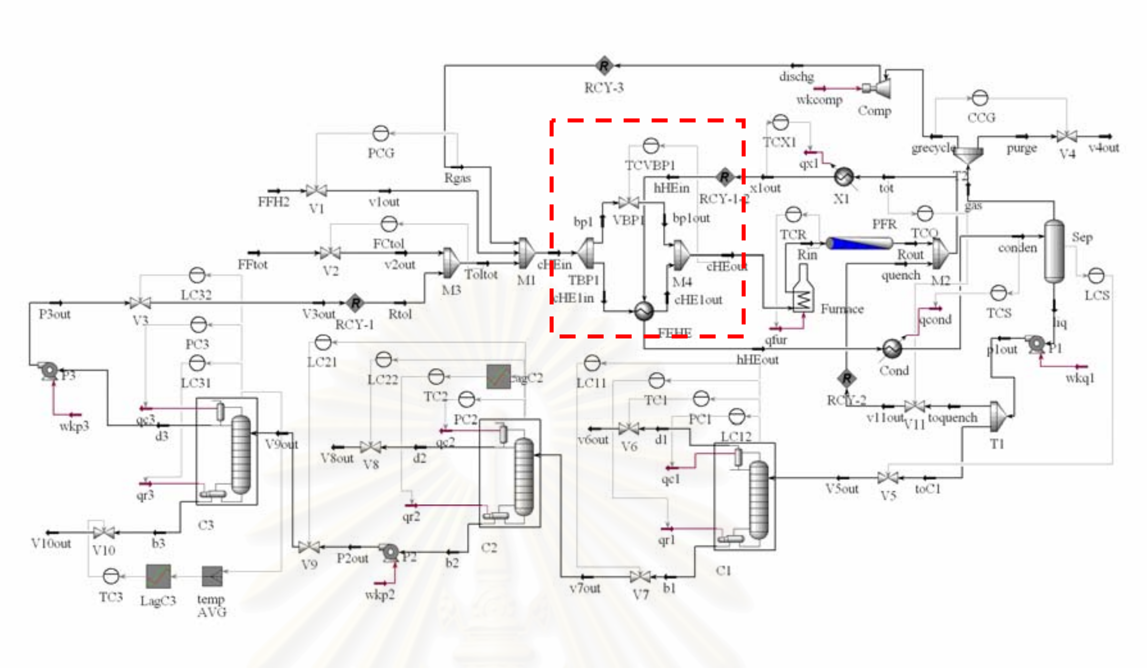


Figure 6.17 Application of control structure 1 (CS1) to the typical of HDA process Alternative 1 (AL1)

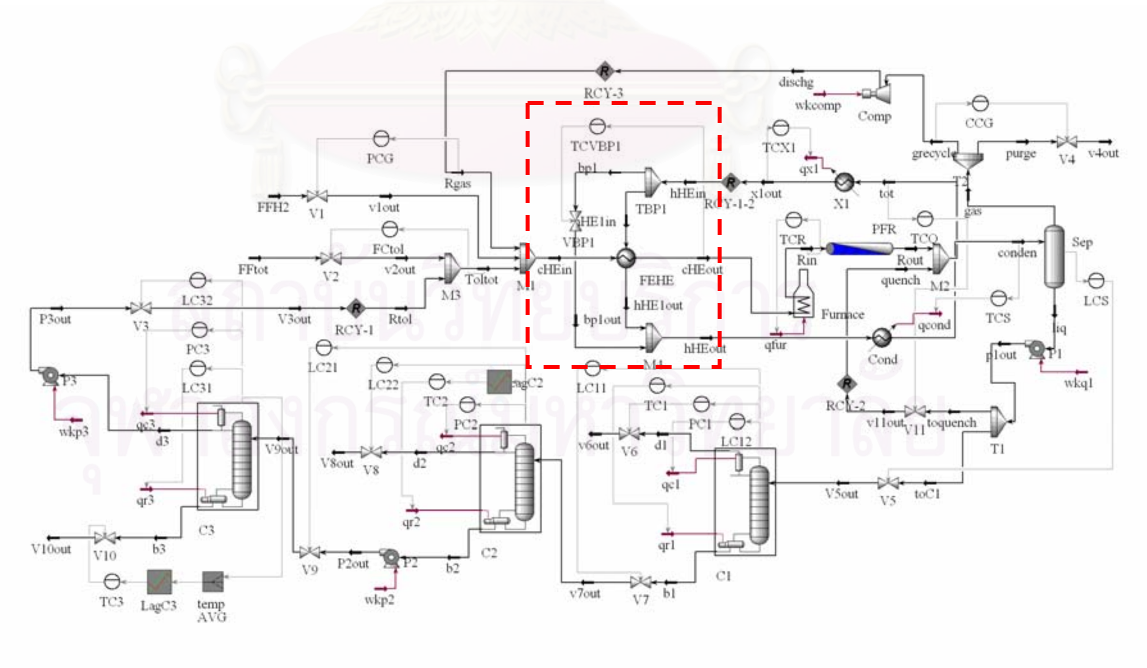


Figure 6.18 Application of control structure 2 (CS2) to the typical of HDA process Alternative 1 (AL1)

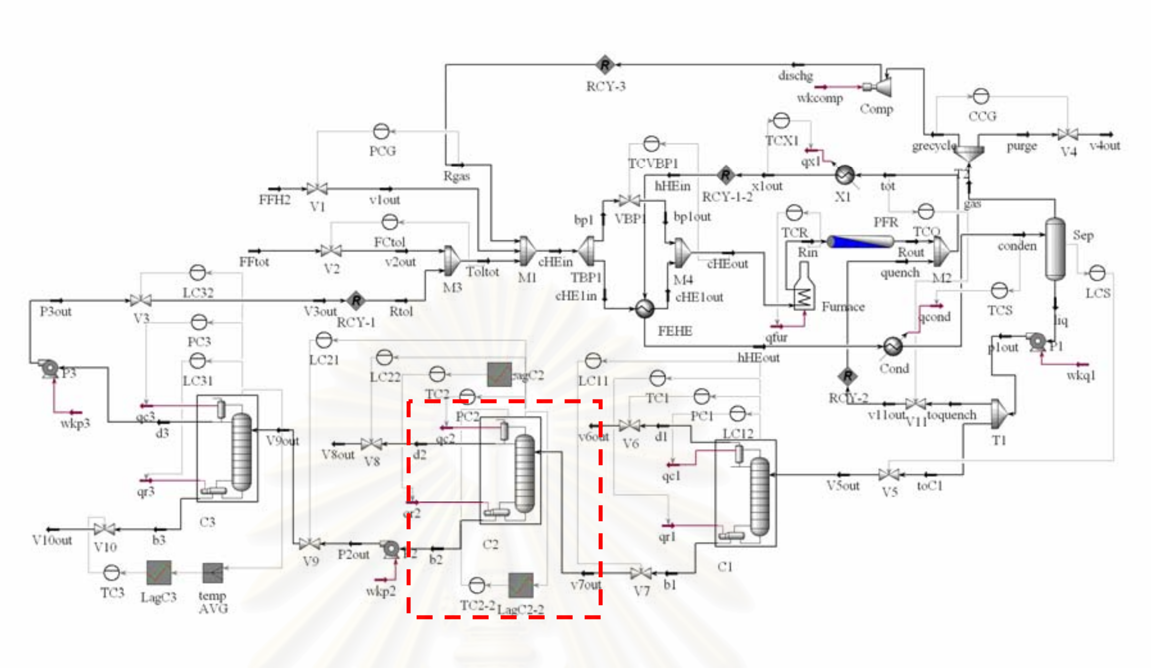


Figure 6.19 Application of control structure 3 (CS3) to the typical of HDA process Alternative 1 (AL1)

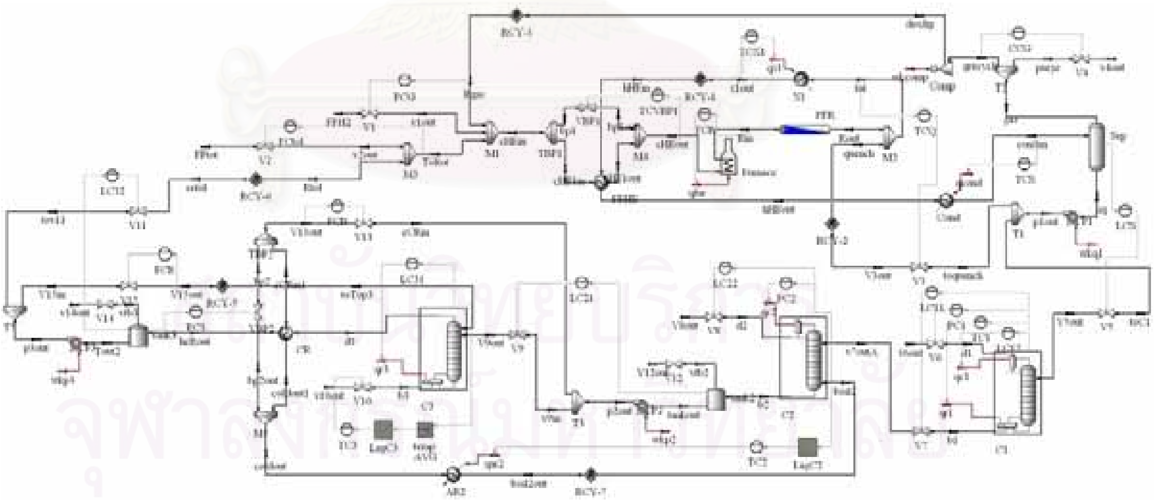


Figure 6.20 Application of control structure 1 (CS1) to the typical of HDA process Alternative 2 (AL2)

6.3.1.4 Control Structure 1 (CS1) for the typical of HDA Process Alternative 2

This control structure is shown in Figure 6.20 and the controller parameters are given in table 6.2. this control structure, the all bypass of a feed effluent heat exchangers (FEHE) is on cold side. Figure 6.20 shows the plantwide control structure of HDA process alternative 2. The major loops in HDA process alternative 2 are the same as those used in Alternative 1 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. 6.20) 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 6.2. P controllers are employed for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.3.1.5 Control Structure 2 (CS2) for the typical of HDA Process Alternative 2

This control structure is shown in Figure 6.21 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 for FEHE. The all bypass of FEHE will be on hot side.

6.3.1.5 Control Structure 3 (CS3) for the typical of HDA Process Alternative 2

This control structure is shown in Figure 6.22 and the controller parameter is given in table 6.2. 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.

Table 6.2 Control Structure and Controller Parameter for the typical of HDA Process Alternative 2: Control Structure 1, Control Structure 2 and Control Structure 3

Controller	Controlled variable	Manipulated variable	Type	Kc	Ti	Td
Reaction section						
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 (V3)	PID	0.828	0.373	0.083
TCR	reactor inlet temperature	furnace duty (qfur)	PID	1.533	0.334	0.074
TCS	separator temperature	cooler duty (qcooler)	PID	0.497	0.198	0.044
TCVBP1*	FEHE cold outlet temperature	FEHE bypass cold stream valve (VBP1)	PID	2.169	0.154	0.034
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
Separation sections						
Stabilizer column						
PC1	column C1 pressure	column C1 gas valve (V6)	PI	2	10	-
TC1	column C1 tray-6 temperature	column C1 reboiler duty (qr1)	PI	2	10	-
LC11	column C1 base level	column C2 feed valve (V7)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	2	-	-
Product column						
PC2	column C2 pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	heat reboiler (qAR2) duty	PID	5.510	4.134	0.900
TC2-2**	column C2 tray-17 temperature	column C2 reflux flow rate	PI	2	10	-
LC21	column C2 base level	column C3 feed valve (V9)	P	2	-	-
LC22	column C2 reflux drum level	column C2 product valve level (V8)	P	2	-	-
FCB	column C2 boil up flow rate	CR cold-inlet valve (V13)	PI	0.5	0.3	-
Recycle column						
PC3	column C3 pressure	CR bypass valve (VBP4)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 by-product valve (V8)	PID	0.616	22.640	5.031
LC31	column C3 base level	column C3 reboiler duty (qr3)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V11)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

** TCVBP 1 or 3 manipulated variable: FEHE 1,2 bypass hot stream valve (VBP1 or 3) used in only CS2.*

*** TC2-2 controller used in only CS3.*

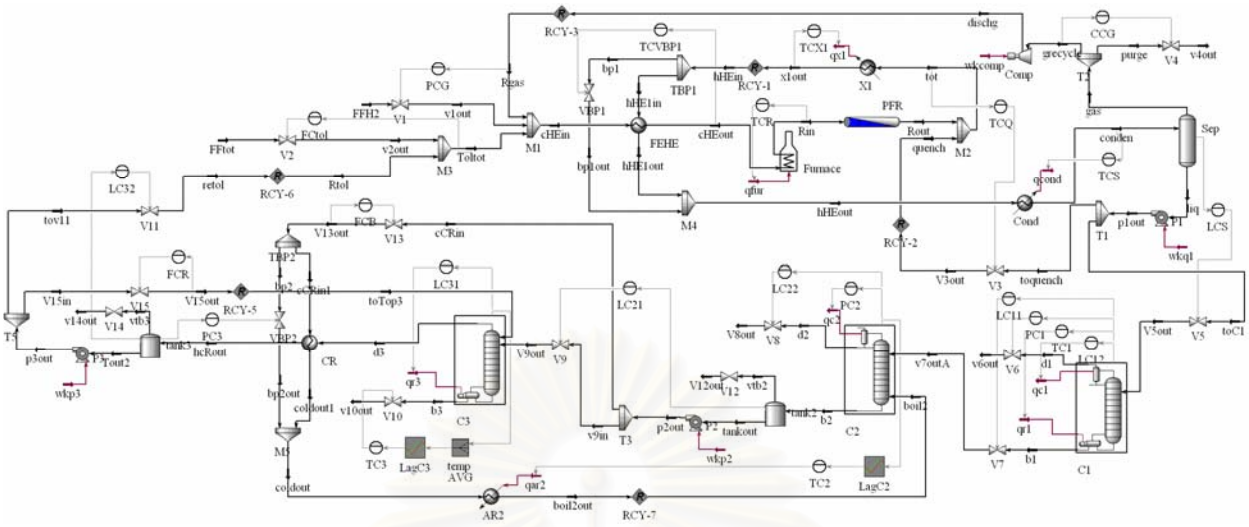


Figure 6.21 Application of control structure 2 (CS2) to the typical of HDA process Alternative 2 (AL2)

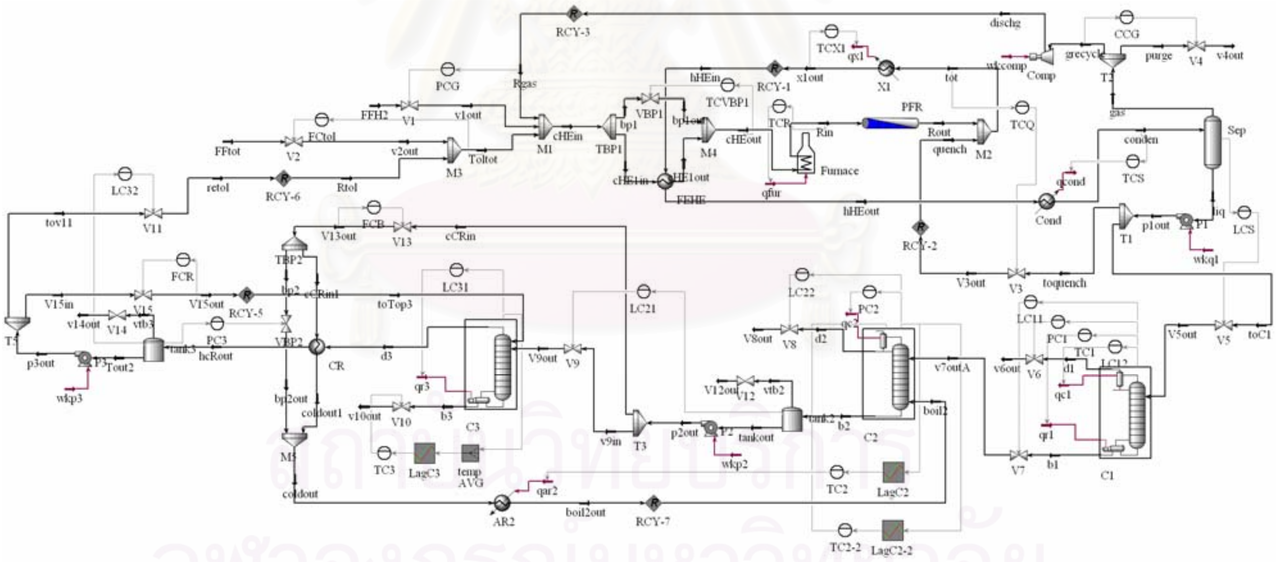


Figure 6.22 Application of control structure 3 (CS3) to the typical of HDA process Alternative 2 (AL2)

6.3.1.7 Control Structure 1 (CS1) for the typical of HDA Process Alternative 3

This control structure is shown in Figure 6.23 and the controller parameters are given in table 6.3. this control structure, the all bypass of a feed effluent heat exchangers (FEHE) is on cold side. Figure 6.23 shows the plantwide control structure of HDA process alternative 3. The major loops in HDA process alternative 3 are the same as those used in Alternative 1. 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. 6.23) 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 6.3. P controllers are employed for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.3.1.8 Control Structure 2 (CS2) for the typical of HDA Process Alternative 3

This control structure is shown in Figure 6.24 and the controller parameter is given in table 6.3. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass of FEHE will be on hot side.

6.3.1.9 Control Structure 3 (CS3) for the typical of HDA Process Alternative 3

This control structure is shown in Figure 6.25 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.

Table 6.3 Control Structure and Controller Parameter for the typical of HDA Process Alternative 3: Control Structure 1, Control Structure 2 and Control Structure 3

Controller	Controlled variable	Manipulated variable	Type	Kc	Ti	Td
Reaction section						
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 (V3)	PID	0.707	0.390	0.087
TCR	reactor inlet temperature	furnace duty (qfur)	PID	1.362	0.343	0.076
TCS	separator temperature	cooler duty (qcooler)	PID	0.887	0.175	0.039
TCVBP1*	FEHE1 cold outlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	5.666	0.221	0.049
TCVBP2*	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	8.414	0.493	0.110
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
Separation sections						
Stabilizer column						
PC1	column C1 pressure	column C1 gas valve (V6)	PI	2	10	-
TC1	column C1 tray-6 temperature	R1 bypass valve (VBP3) and auxiliary reboiler (Q-100) duty	PID	14.168	0.909	0.202
LC11	column C1 base level	column C2 feed valve (V7)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	2	-	-
FCB1	column C1 boil up flow rate	R1 cold-inlet valve (V19)	PI	0.5	0.3	-
Product column						
PC2	column C2 pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	heat reboiler (qAR2) duty	PID	5.629	4.100	0.911
TC2-2*	column C2 tray-17 temperature	column C2 reflux flow rate	PID	2.386	81.441	18.098
LC21	column C2 base level	column C3 feed valve (V9)	P	2	-	-
LC22	column C2 reflux drum level	column C2 product valve level (V8)	P	2	-	-
FCB	column C2 boil up flow rate	CR cold-inlet valve (V13)	PI	0.5	0.3	-
Recycle column						
PC3	column C3 pressure	CR bypass valve (VBP4)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 by-product valve (V8)	PID	0.693	20.914	4.648
LC31	column C3 base level	column C3 reboiler duty (qr3)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V11)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

** TCVBP 1 or 3 manipulated variable: FEHE 1,2 bypass hot stream valve (VBP1 or 3) used in only CS2.*

*** TC2-2 controller used in only CS3.*

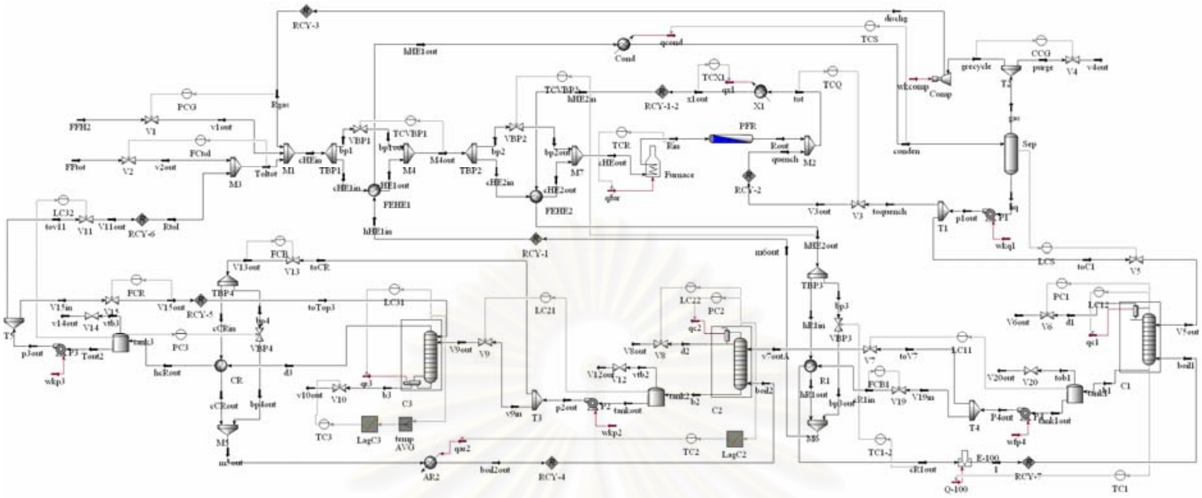


Figure 6.23 Application of control structure 1 (CS1) to the typical of HDA process Alternative 3 (AL3)

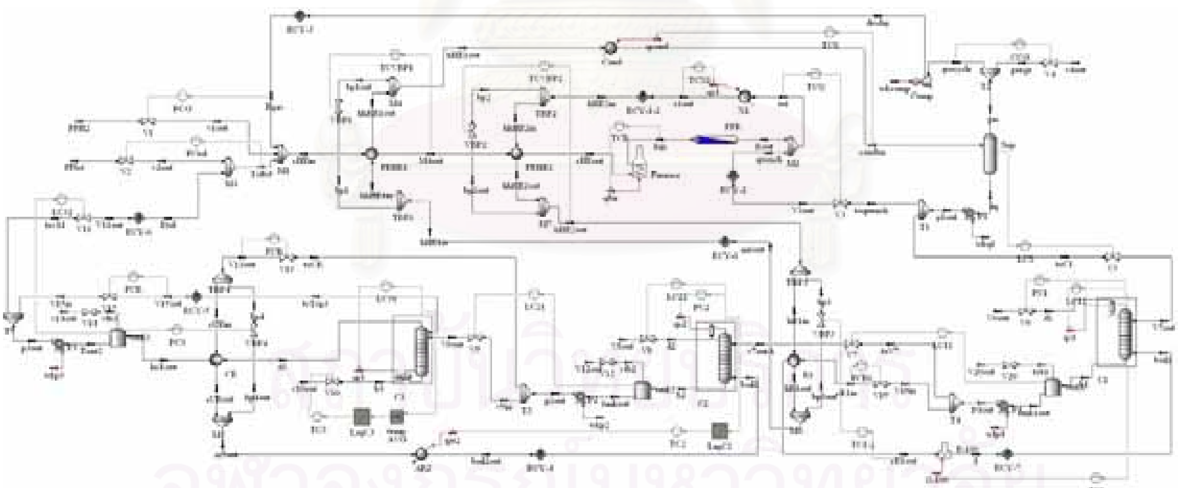


Figure 6.24 Application of control structure 2 (CS2) to the typical of HDA process Alternative 3 (AL3)

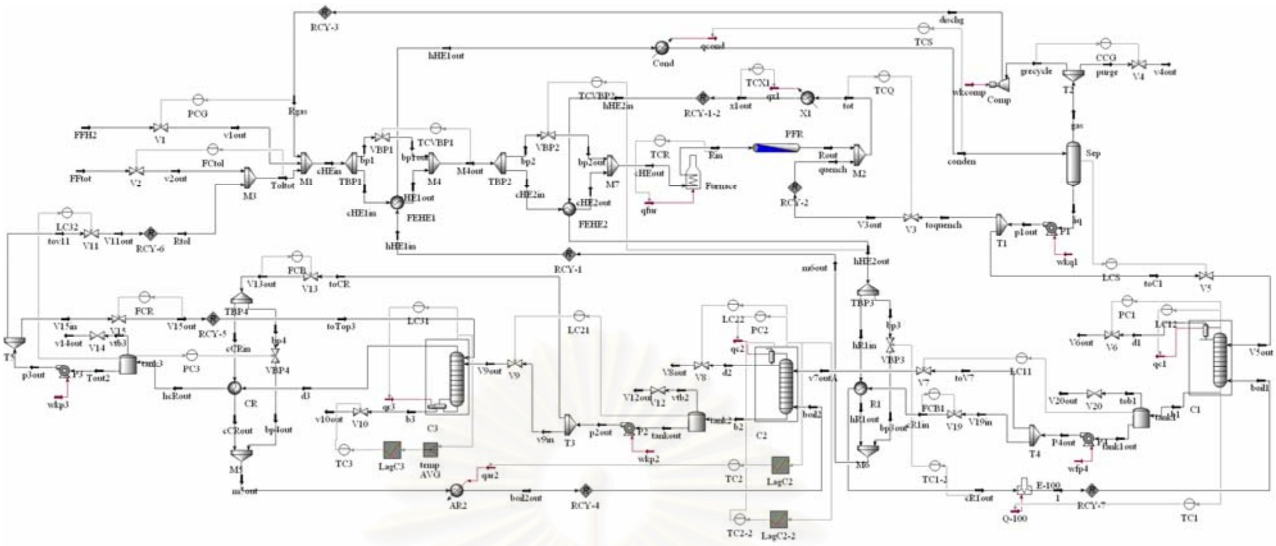


Figure 6.25 Application of control structure 3 (CS3) to the typical of HDA process Alternative 3 (AL3)

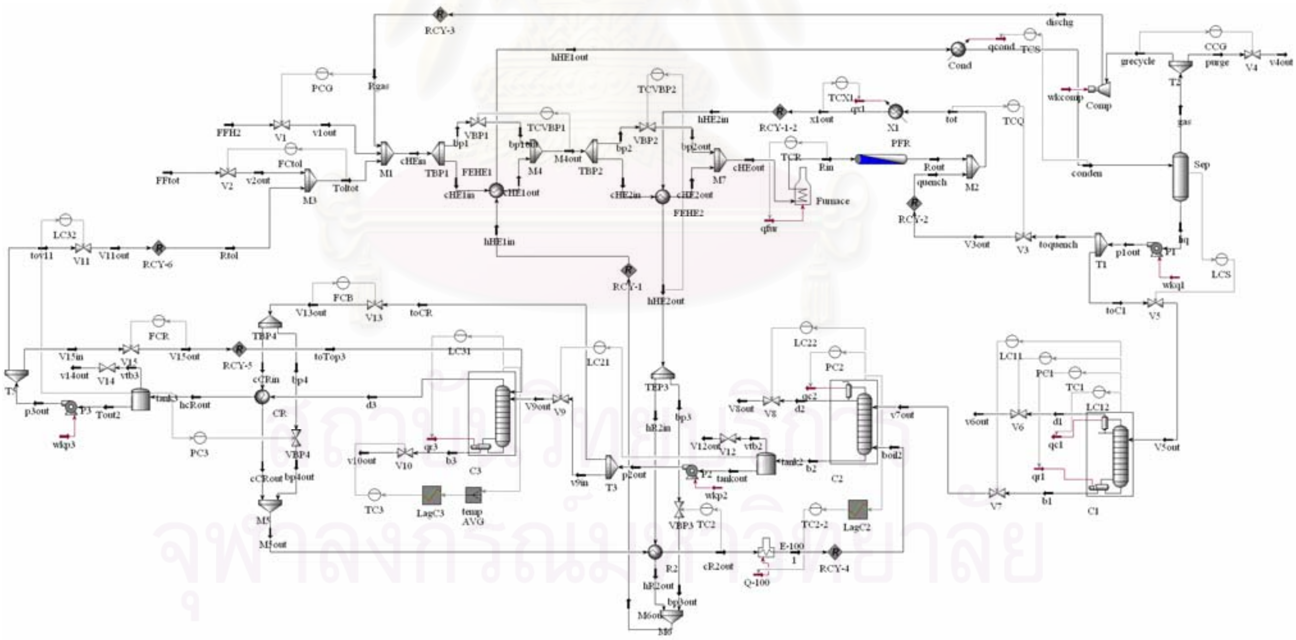


Figure 6.26 Application of control structure 1 (CS1) to the typical of HDA process Alternative 4 (AL4)

6.3.1.10 Control Structure 1 (CS1) for the typical of HDA Process Alternative 4

This control structure is shown in Figure 6.26 and the controller parameters are given in table 6.4. This control structure, the all bypass of feed effluent heat exchangers (FEHE) is on cold side. Figure 6.26 shows the plantwide control structure of HDA process alternative 4. The major loops in HDA process alternative 4 are the same as those used in Alternative 1. 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. 6.26) 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 6.4. P controllers are employed for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.3.1.11 Control Structure 2 (CS2) for the typical of HDA Process Alternative 2

This control structure is shown in Figure 6.27 and the controller parameter is given in table 6.4. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass of FEHE will be on hot side.

6.3.1.12 Control Structure 3 (CS3) for the typical of HDA Process Alternative 2

This control structure is shown in Figure 6.28 and the controller parameter is given in table 6.4. 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.

Table 6.4 Control Structure and Controller Parameter for the typical of HDA Process Alternative 4: Control Structure 1, Control Structure 2 and Control Structure 3

Controller	Controlled variable	Manipulated variable	Type	Kc	Ti	Td
Reaction section						
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 (V3)	PID	0.717	0.380	0.084
TCR	reactor inlet temperature	furnace duty (qfur)	PID	1.044	0.360	0.080
TCS	separator temperature	cooler duty (qcooler)	PID	1.762	0.154	0.034
TCVBP1*	FEHE1 cold outlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	5.869	0.217	0.048
TCVBP2*	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP2)	PID	8.332	0.514	0.114
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
Separation sections						
Stabilizer column						
PC1	column C1 pressure	column C1 gas valve (V6)	PI	2	10	-
TC1	column C1 tray-6 temperature	column C1 reboiler duty (qr1)	PI	2	10	-
LC11	column C1 base level	column C2 feed valve (V7)	P	2	-	-
LC12	column C1 reflux drum	column C1 condenser duty (qc1)	P	2	-	-
Product column						
PC2	column C2 pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 tray-12 temperature	R2 bypass valve (VBP3) and auxiliary reboiler (Q-100) duty	PID	4.372	1.195	0.265
TC2-2**	column C2 tray-17 temperature	column C2 reflux flow rate	PID	5.209	11.266	2.504
LC21	column C2 base level	column C3 feed valve (V9)	P	2	-	-
LC22	column C2 reflux drum level	column C2 product valve level (V8)	P	2	-	-
FCB	column C2 boil up flow rate	CR cold-inlet valve (V13)	PI	0.5	0.3	-
Recycle column						
PC3	column C3 pressure	CR bypass valve (VBP4)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 by-product valve (V8)	PID	0.688	21.123	4.694
LC31	column C3 base level	column C3 reboiler duty (qr3)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V11)	P	2	-	-
FCR	column C3 reflux flow rate	reflux valve (V13)	PI	0.5	0.3	-

** TCVBP 1 or 3 manipulated variable: FEHE 1,2 bypass hot stream valve (VBP1 or 3) used in only CS2.*

*** TC2-2 controller used in only CS3.*

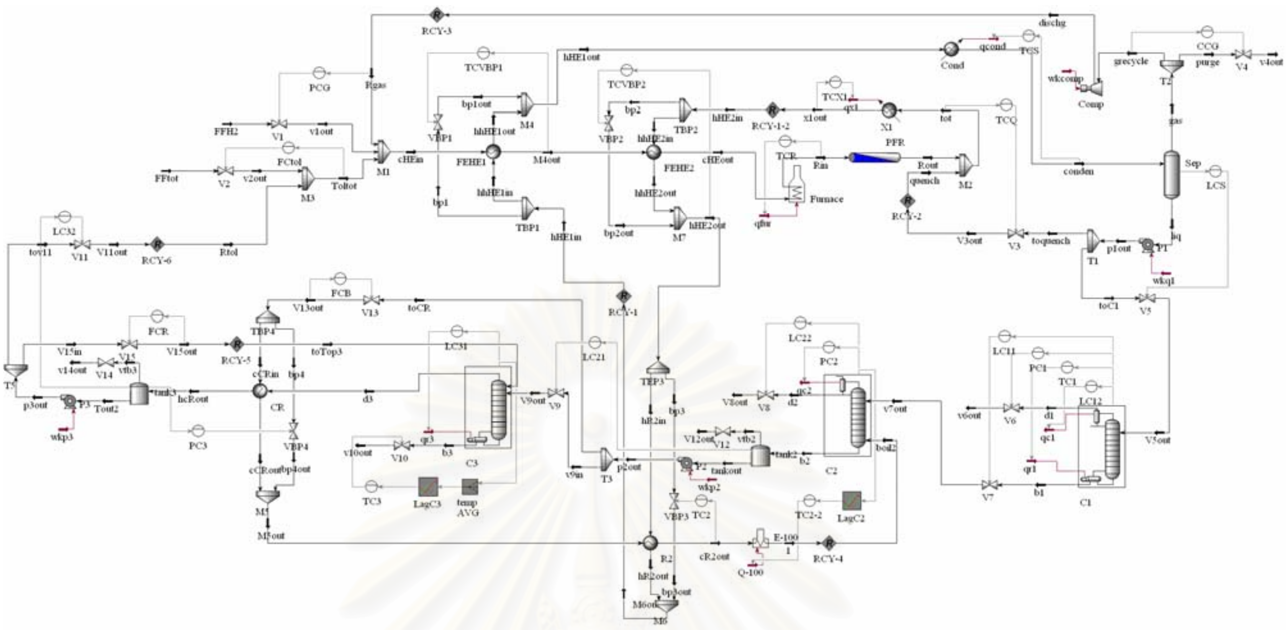


Figure 6.27 Application of control structure 2 (CS2) to the typical of HDA process Alternative 4 (AL4)

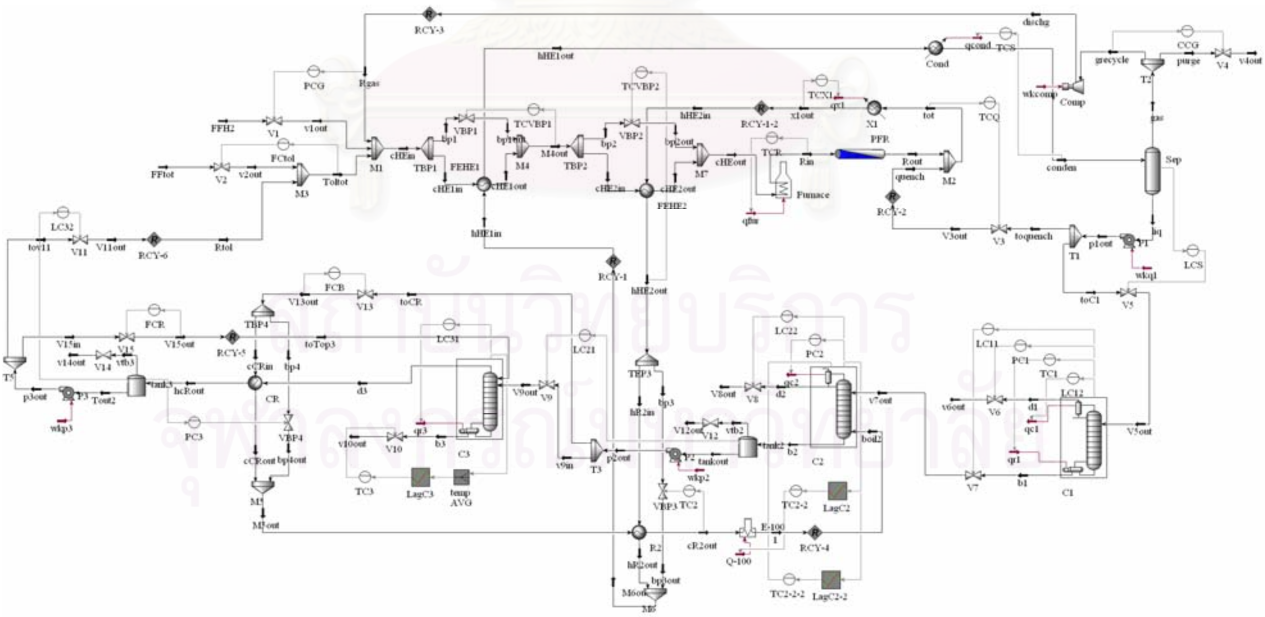


Figure 6.28 Application of control structure 3 (CS3) to the typical of HDA process Alternative 4 (AL4)

6.3.2 Design of Plantwide Control for the Heat-Integrated Plant of HDA Process (HIP1, HIP2, HIP3 and HIP4)

The three new control structures are designed for the heat-integrated plant of HDA process HIP1 to 4 that is propose in this research.

6.3.2.1 Control Structure 1 for the heat-integrated plant of HDA Process HIP1

This control structure is shown in Figure 6.29 and the controller parameters are given in table 6.5. this control structure, the all bypass of a feed effluent heat exchangers (FEHE) is on cold side. Figure 6.29 shows the plantwide control structure for the heat-integrated plant of HDA process HIP1. The major loops in HDA process HIP1 are the same as those used alternative 1, for the outlet temperature control in FEHE and the tray temperature control in the recycle column (C3). The furnace inlet temperature is controlled by manipulating the valve on the bypass line Based on the heat pathway heuristics for plantwide control. The control structure and controller parameter are given in table 6.5. P controllers are employed for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.3.2.2 Control Structure 2 for the heat-integrated plant of HDA Process HIP1

This control structure is shown in Figure 6.30 and the controller parameter is given in table 6.5. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass of FEHE will be on hot side.

6.3.2.3 Control Structure 3 for the heat-integrated plant of HDA Process HIP1

This control structure is shown in Figure 6.31 and the controller parameter is given in table 6.5. 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 bottom stage temperature and tray 2 temperature controls.

Table 6.5 Control Structure and Controller Parameter for heat-integrated plant of HDA Process HIP1: Control Structure 1, Control Structure 2 and Control Structure 3

Controller	Controlled variable	Manipulated variable	Type	Kc	Ti	Td
Reaction section						
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 (V3)	PID	0.425	0.402	0.089
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.376	0.406	0.090
TCS	separator temperature	cooler duty (qcooler)	PID	0.526	0.197	0.044
TCVBP1*	FEHE cold outlet temperature	FEHE bypass cold stream valve (VBP1)	PID	1.609	0.331	0.074
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
Separation sections						
Stabilizer column						
PC1	column C1 pressure	column C1 condenser duty (qc1)	PI	2	10	-
TC1	column C1 tray-14 temperature	column C1 reboiler duty (qr1)	PI	2	10	-
LC11	column C1 base level	column C3 feed valve (V9)	P	2	-	-
LC12	column C1 reflux drum	column C2 gas feed valve (V6)	P	2	-	-
Product column						
PC2	column C2 pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 bottom stage temperature	column C2 reboiler duty (qr2)	P	2	-	-
TC2-2**	column C2 tray-2 temperature	column C2 reflux flow rate	PID	13.818	1.918	0.426
LC21	column C2 base level	column C2 product valve level (V8)	P	2	-	-
LC22	column C2 reflux drum level	column C2 gas valve (V7)	P	2	-	-
Recycle column						
PC3	column C3 pressure	column C3 condenser duty (qc3)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 by-product valve (V10)	PID	0.440	16.228	3.606
LC31	column C3 base level	column C3 reboiler duty (qr3)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-

** TCVBP 1 or 3 manipulated variable: FEHE 1,2 bypass hot stream valve (VBP1 or 3) used in only CS2.*

*** TC2-2 controller used in only CS3.*

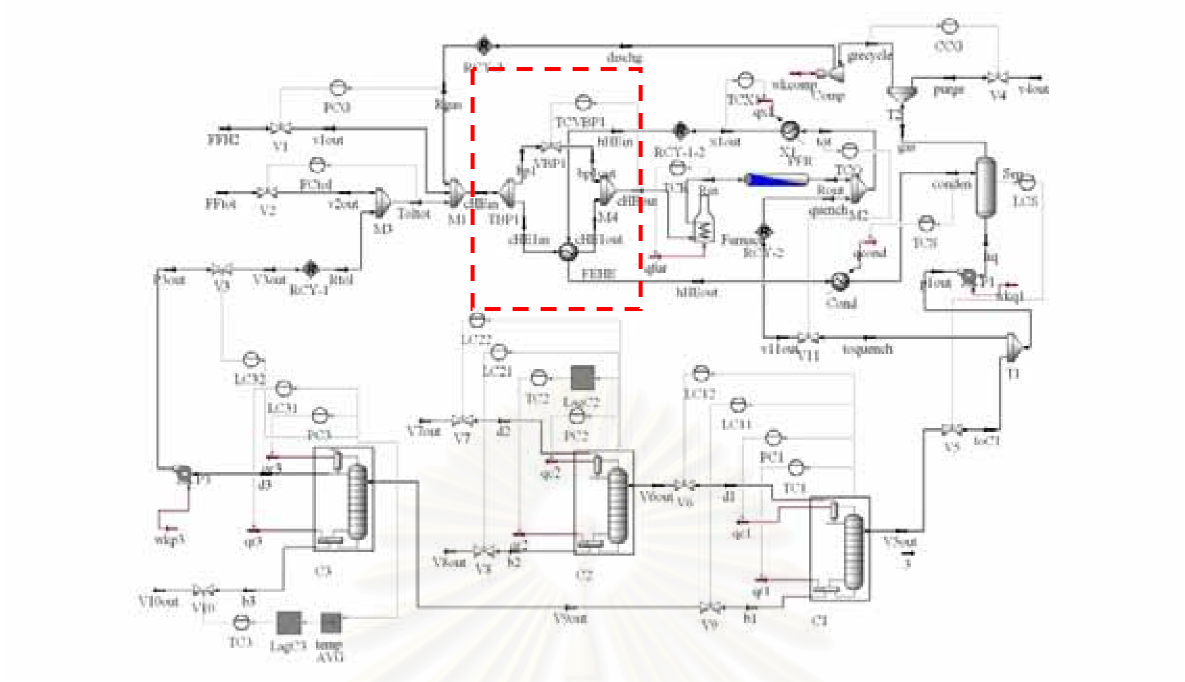


Figure 6.29 Application of control structure 1 (CS1) to the heat-integrated plant of HDA process HIP1

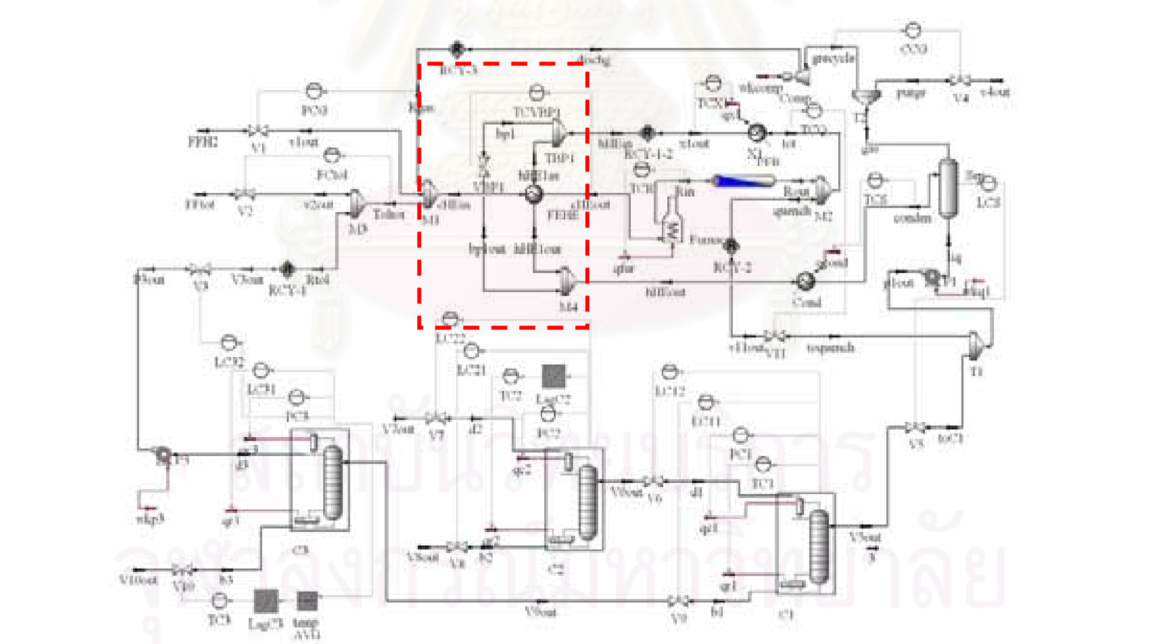


Figure 6.30 Application of control structure 2 (CS2) to the heat-integrated plant of HDA process HIP1

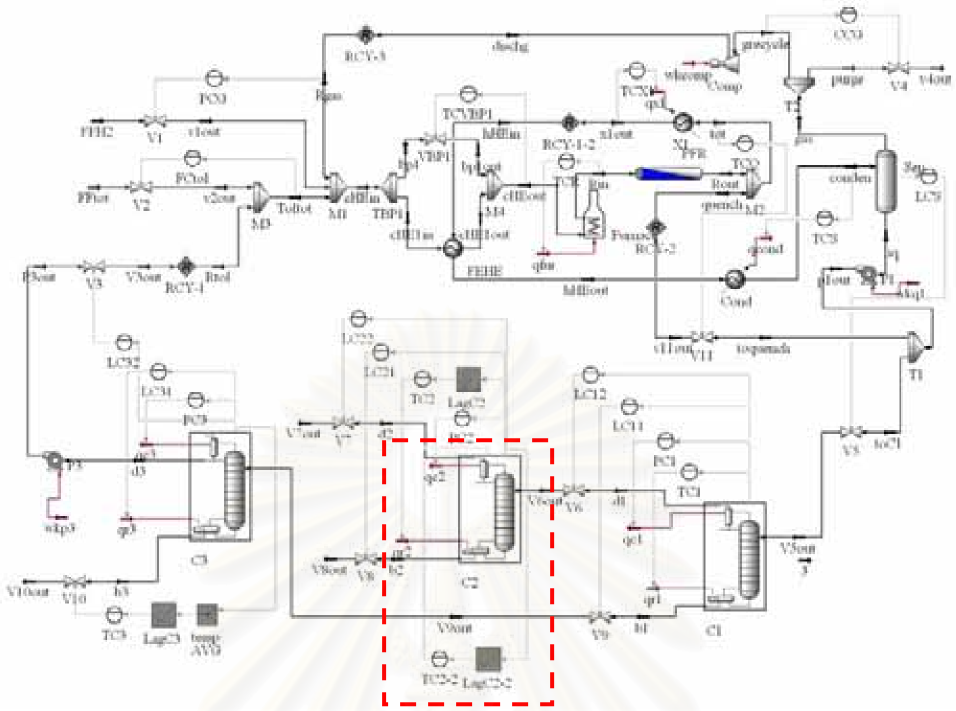


Figure 6.31 Application of control structure 3 (CS3) to the heat-integrated plant of HDA process HIP1

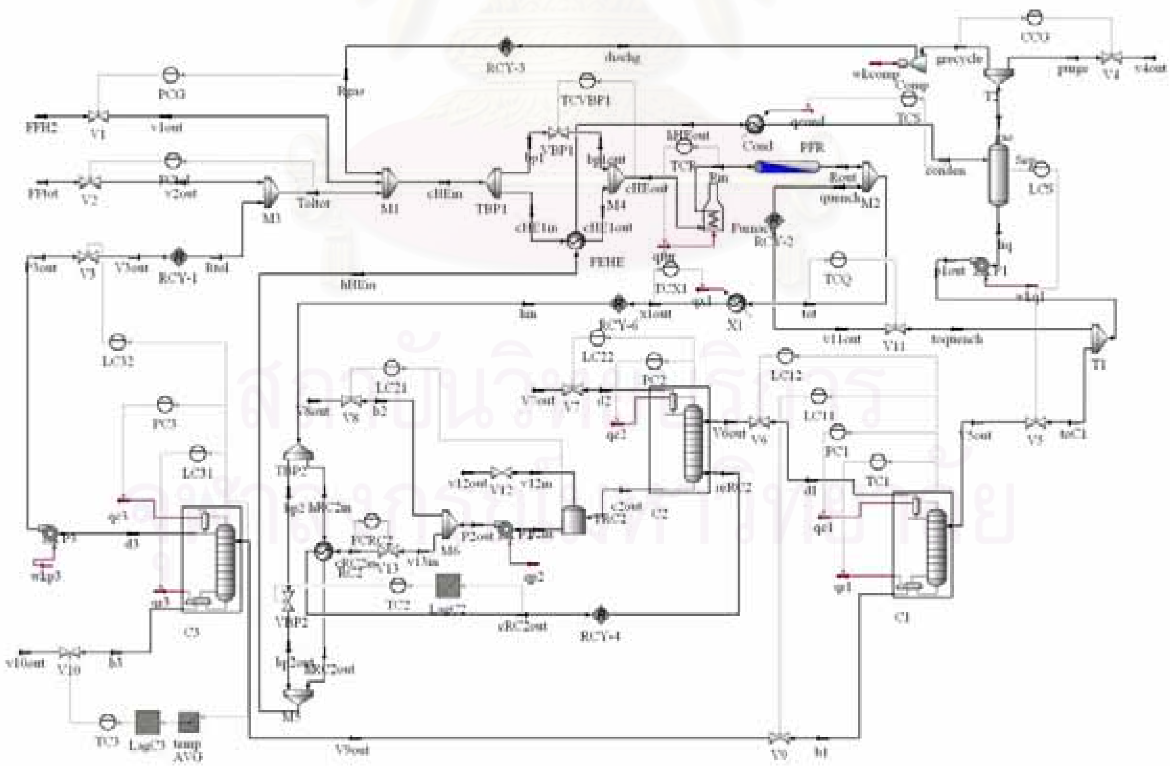


Figure 6.32 Application of control structure 1 (CS1) to the heat-integrated plant of HDA process HIP2

6.3.2.4 Control Structure 1 for the heat-integrated plant of HDA Process HIP2

This control structure is shown in Figure 6.32 and the controller parameters are given in table 6.6. this control structure, the all bypass of a feed effluent heat exchangers (FEHE) is on cold side. Figure 6.32 shows the plantwide control structure for the heat-integrated plant of HDA process HIP2. The major loops in HDA process HIP2 are the same as those used HIP1, for the outlet temperature control in FEHE and the tray temperature control in the recycle column (C3).The furnace inlet temperature is controlled by manipulating the valve on the bypass line Based on the heat pathway heuristics for plantwide control. The control structure and controller parameter are given in table 6.6. P controllers are employed for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.3.2.5 Control Structure 2 for the heat-integrated plant of HDA Process HIP2

This control structure is shown in Figure 6.33 and the controller parameter is given in table 6.6. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass of FEHE will be on hot side.

6.3.2.6 Control Structure 3 for the heat-integrated plant of HDA Process HIP2

This control structure is shown in Figure 6.34 and the controller parameter is given in table 6.6. 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 bottom stage temperature and tray 2 temperature controls.

Table 6.6 Control Structure and Controller Parameter for heat-integrated plant of HDA Process HIP2: Control Structure 1, Control Structure 2 and Control Structure 3

Controller	Controlled variable	Manipulated variable	Type	Kc	Ti	Td
Reaction section						
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 (V3)	PID	0.436	0.401	0.089
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.376	0.406	0.090
TCS	separator temperature	cooler duty (qcooler)	PID	0.538	0.196	0.044
TCVBP1*	FEHE cold outlet temperature	FEHE bypass cold stream valve (VBP1)	PID	1.498	0.336	0.075
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
Separation sections						
Stabilizer column						
PC1	column C1 pressure	column C1 condenser duty (qc1)	PI	2	10	-
TC1	column C1 tray-14 temperature	column C1 reboiler duty (qr1)	PI	2	10	-
LC11	column C1 base level	column C3 feed valve (V9)	P	2	-	-
LC12	column C1 reflux drum	column C2 gas feed valve (V6)	P	2	-	-
Product column						
PC2	column C2 pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 bottom stage temperature	column C2 reboiler duty (qr2)	P	2	-	-
TC2-2*	column C2 tray-2 temperature	column C2 reflux flow rate	PID	14.287	1.771	0.394
LC21	column C2 base level	column C2 product valve level (V8)	P	2	-	-
LC22	column C2 reflux drum level	column C2 gas valve (V7)	P	2	-	-
FCRC2	column C2 boil up flow rate	RC2 cold-inlet valve (V13)	PI	0.5	0.3	-
Recycle column						
PC3	column C3 pressure	column C3 condenser duty (qc3)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 by-product valve (V10)	PID	0.446	16.107	3.579
LC31	column C3 base level	column C3 reboiler duty (qr3)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-

** TCVBP 1 or 3 manipulated variable: FEHE 1,2 bypass hot stream valve (VBP1 or 3) used in only CS2.*

*** TC2-2 controller used in only CS3.*

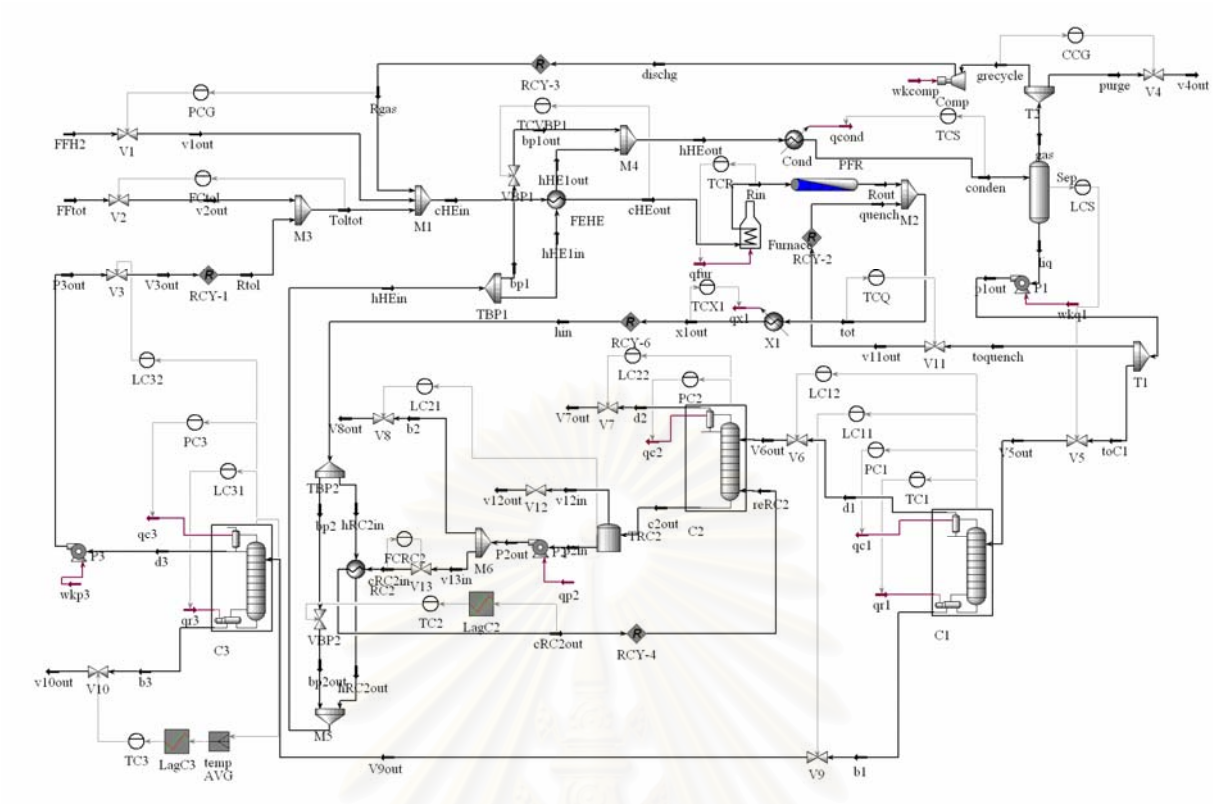


Figure 6.33 Application of control structure 2 (CS2) to the heat-integrated plant of HDA process HIP2

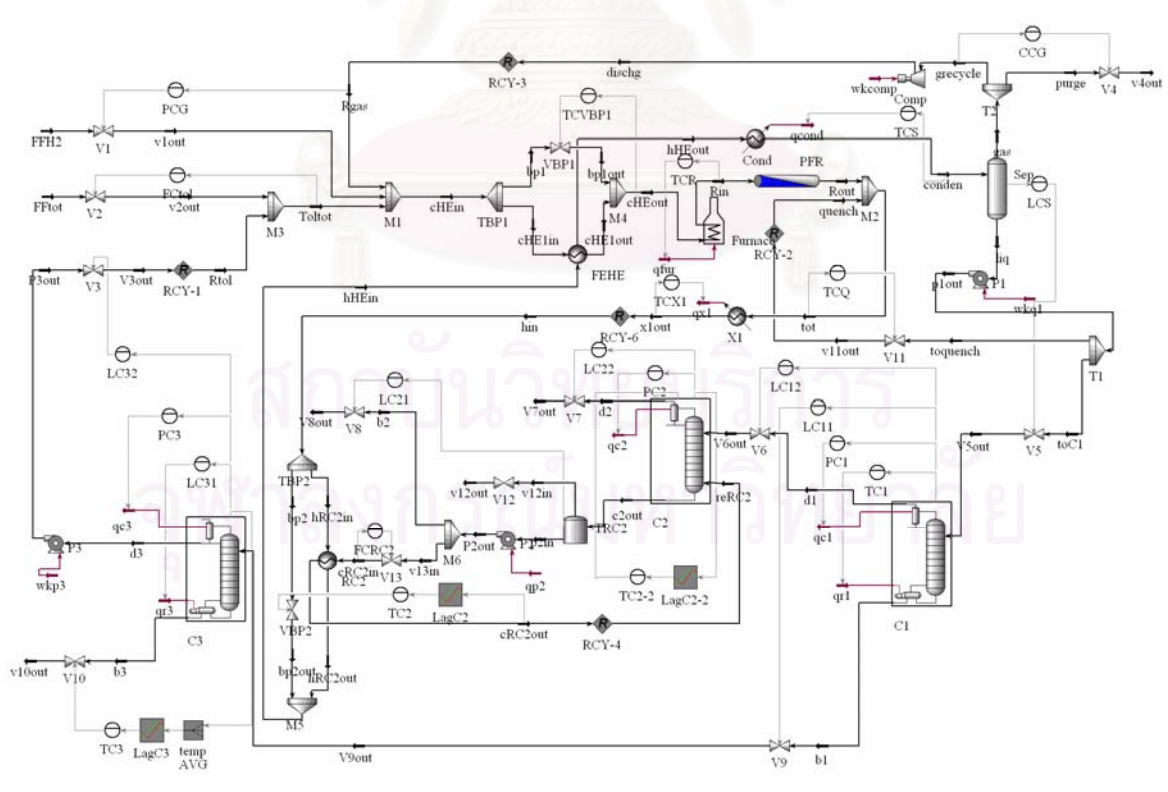


Figure 6.34 Application of control structure 3 (CS3) to the heat-integrated plant of HDA process HIP2

6.3.2.7 Control Structure 1 for the heat-integrated plant of HDA Process HIP3

This control structure is shown in Figure 6.35 and the controller parameters are given in table 6.7. this control structure, the all bypass of a feed effluent heat exchangers (FEHE) is on cold side. Figure 6.35 shows the plantwide control structure for the heat-integrated plant of HDA process HIP3. The major loops in HDA process HIP3 are the same as those used HIP1, for the outlet temperature control in FEHE and the tray temperature control in the recycle column (C3).The furnace inlet temperature is controlled by manipulating the valve on the bypass line Based on the heat pathway heuristics for plantwide control. 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.

6.3.2.8 Control Structure 2 for the heat-integrated plant of HDA Process HIP3

This control structure is shown in Figure 6.36 and the controller parameter is given in table 6.7. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass of FEHE will be on hot side.

6.3.2.9 Control Structure 3 for the heat-integrated plant of HDA Process HIP3

This control structure is shown in Figure 6.37 and the controller parameter is given in table 6.7. 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 bottom stage temperature and tray 2 temperature controls.

Table 6.7 Control Structure and Controller Parameter for heat-integrated plant of HDA Process HIP3: Control Structure 1, Control Structure 2 and Control Structure 3

Controller	Controlled variable	Manipulated variable	Type	Kc	Ti	Td
Reaction section						
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 (V3)	PID	0.508	0.316	0.070
TCR	reactor inlet temperature	furnace duty (qfur)	PID	1.269	0.347	0.077
TCS	separator temperature	cooler duty (qcooler)	PID	0.743	0.180	0.040
TCVBP1*	FEHE1 cold outlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	3.347	0.271	0.060
TCVBP2*	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP3)	PID	9.414	0.421	0.093
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
Separation sections						
Stabilizer column						
PC1	column C1 pressure	column C1 condenser duty (qc1)	PI	2	10	-
TC1	column C1 tray-14 temperature	column C1 reboiler duty (qr1)	PI	2	10	-
LC11	column C1 base level	column C3 feed valve (V9)	P	2	-	-
LC12	column C1 reflux drum	column C2 gas feed valve (V6)	P	2	-	-
Product column						
PC2	column C2 pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 bottom stage temperature	column C2 reboiler duty (qr2)	P	2	-	-
TC2-2*	column C2 tray-2 temperature	column C2 reflux flow rate	PID	13.893	1.965	0.437
LC21	column C2 base level	column C2 product valve level (V8)	P	2	-	-
LC22	column C2 reflux drum level	column C2 gas valve (V7)	P	2	-	-
Recycle column						
PC3	column C3 pressure	column C3 condenser duty (qc3)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 by-product valve (V10)	PI	0.1	0.1	-
LC31	column C3 base level	column C3 reboiler duty (qr3)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-
FCRC3	column C3 boil up flow rate	RC3 cold-inlet valve (V14)	PI	0.5	0.3	-

** TCVBP 1 or 3 manipulated variable: FEHE 1,2 bypass hot stream valve (VBP1 or 3) used in only CS2.*

*** TC2-2 controller used in only CS3.*

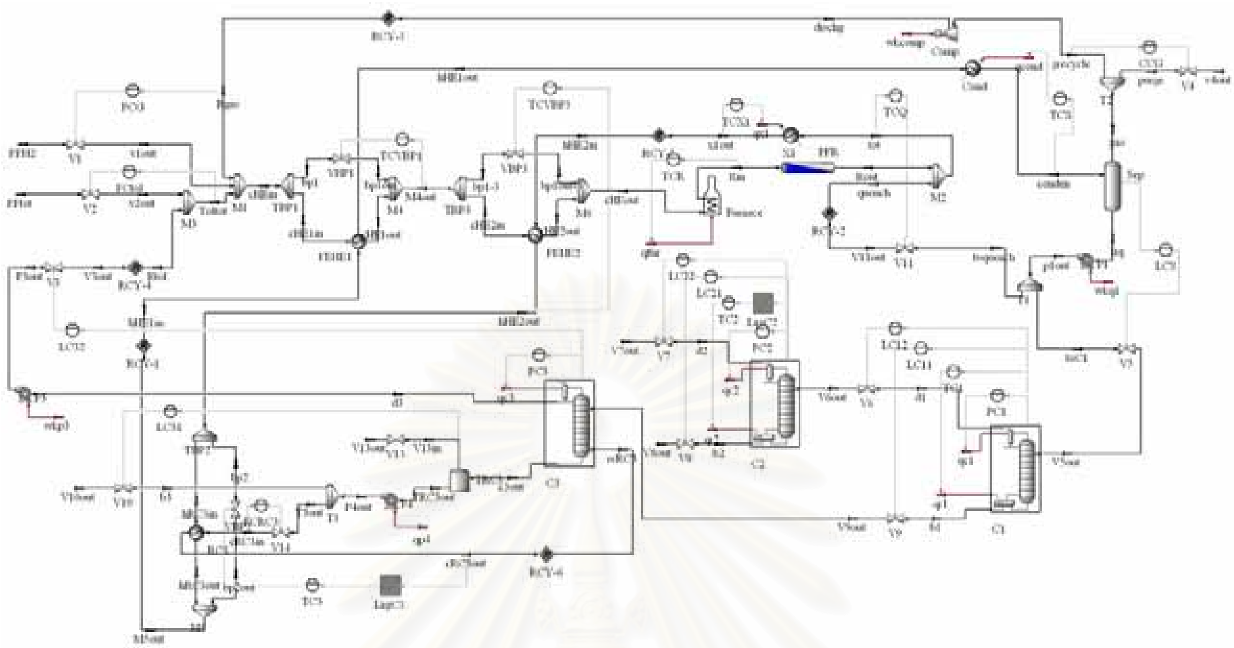


Figure 6.35 Application of control structure 1 (CS1) to the heat-integrated plant of HDA process HIP3

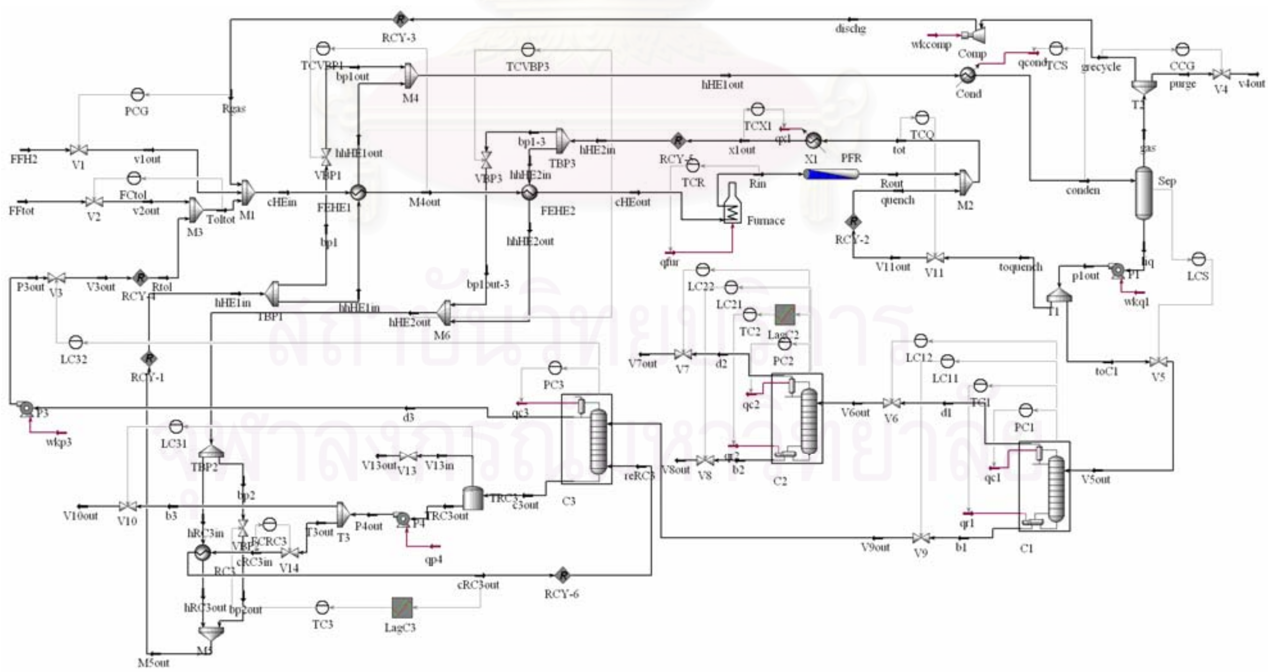


Figure 6.36 Application of control structure 2 (CS2) to the heat-integrated plant of HDA process HIP3

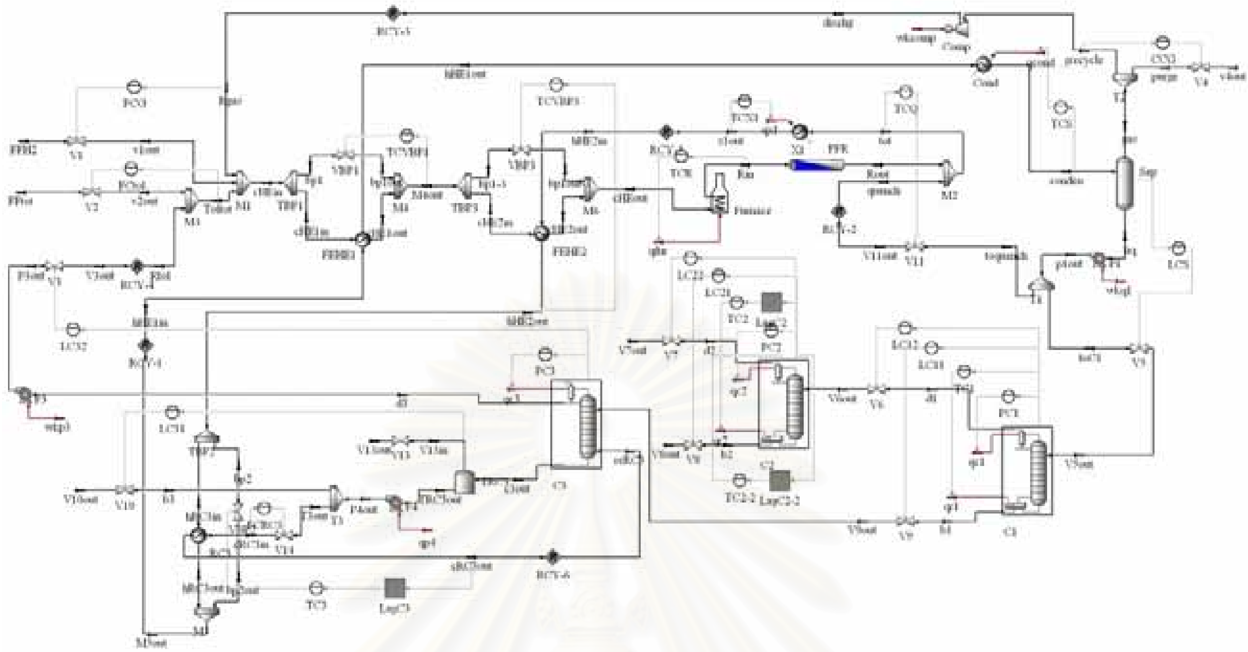


Figure 6.37 Application of control structure 3 (CS3) to the heat-integrated plant of HDA process HIP3

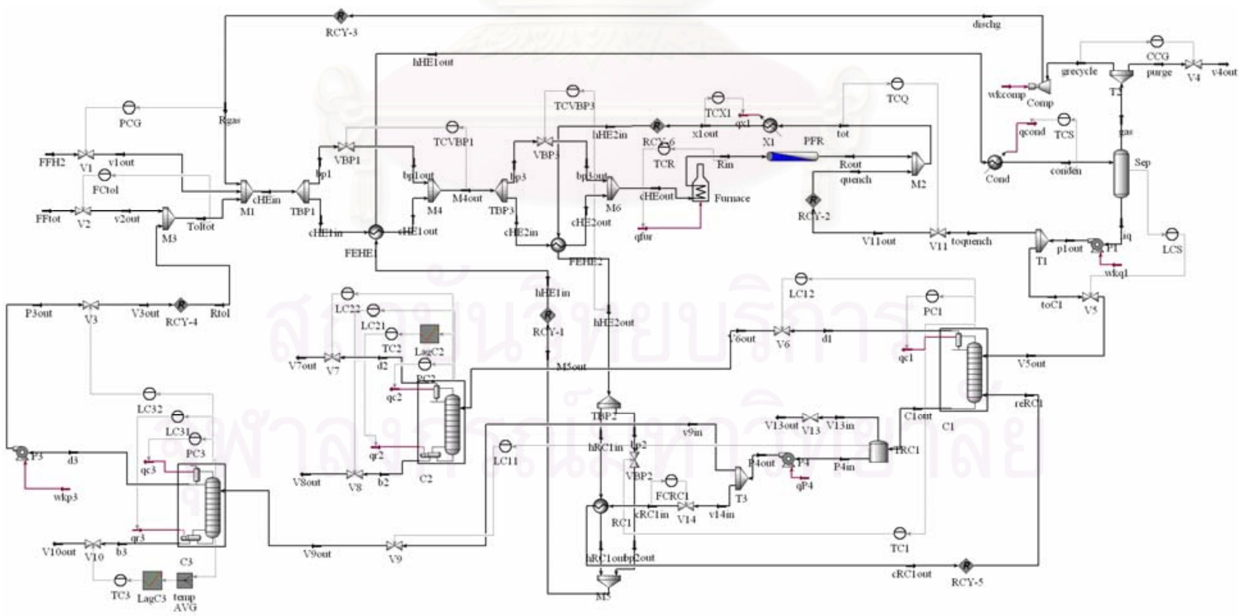


Figure 6.38 Application of control structure 1 (CS1) to the heat-integrated plant of HDA process HIP4

6.3.2.10 Control Structure 1 for the heat-integrated plant of HDA Process HIP4

This control structure is shown in Figure 6.38 and the controller parameters are given in table 6.8. this control structure, the all bypass of a feed effluent heat exchangers (FEHE) is on cold side. Figure 6.38 shows the plantwide control structure for the heat-integrated plant of HDA process HIP3. The major loops in HDA process HIP3 are the same as those used alternative 1, for the outlet temperature control in FEHE and the tray temperature control in the recycle column (C3).Tthe furnace inlet temperature is controlled by manipulating the valve on the bypass line Based on the heat pathway heuristics for plantwide control. The control structure and controller parameter are given in table 6.22. P controllers are employed for level loops, PI controllers for the pressure and flow loops and PID controllers for temperature loop.

6.3.2.11 Control Structure 2 for the heat-integrated plant of HDA Process HIP4

This control structure is shown in Figure 6.39 and the controller parameter is given in table 6.8. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass of FEHE will be on hot side.

6.3.2.12 Control Structure 3 for the heat-integrated plant of HDA Process HIP4

This control structure is shown in Figure 6.40 and the controller parameter is given in table 6.8. 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 bottom stage temperature and tray 2 temperature controls.

Table 6.8 Control Structure and Controller Parameter for heat-integrated plant of HDA Process HIP4: Control Structure 1, Control Structure 2 and Control Structure 3

Controller	Controlled variable	Manipulated variable	Type	Kc	Ti	Td
Reaction section						
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 (V3)	PID	0.387	0.405	0.090
TCR	reactor inlet temperature	furnace duty (qfur)	PID	0.219	0.420	0.093
TCS	separator temperature	cooler duty (qcooler)	PID	0.745	0.180	0.040
TCVBP1*	FEHE1 cold outlet temperature	FEHE1 bypass cold stream valve (VBP1)	PID	4.649	0.234	0.052
TCVBP2*	FEHE2 hot-outlet temperature	FEHE2 bypass cold stream valve (VBP3)	PID	11.275	0.498	0.111
LCS	separator liquid level	column C1 feed valve (V5)	P	2	-	-
Separation sections						
Stabilizer column						
PC1	column C1 pressure	column C1 condenser duty (qc1)	PI	2	10	-
TC1	column C1 tray-14 temperature	RC1 bypass valve (VBP2)	PI	2	10	-
LC11	column C1 base level	column C3 feed valve (V9)	P	2	-	-
LC12	column C1 reflux drum	column C2 gas feed valve (V6)	P	2	-	-
FCRC1	column C1 boil up flow rate	RC1 cold-inlet valve (V14)	PI	0.5	0.3	-
Product column						
PC2	column C2 pressure	column C2 condenser duty (qc2)	PI	2	10	-
TC2	column C2 bottom stage temperature	column C2 reboiler duty (qr2)	P	2	-	-
TC2-2**	column C2 tray-2 temperature	column C2 reflux flow rate	PID	13.263	1.930	0.429
LC21	column C2 base level	column C2 product valve level (V8)	P	2	-	-
LC22	column C2 reflux drum level	column C2 gas valve (V7)	P	2	-	-
Recycle column						
PC3	column C3 pressure	column C3 condenser duty (qc3)	PI	2	10	-
TC3	AVG avg. temp. of C3-tray 1,2, 3, and 4	column C3 by-product valve (V10)	PID	0.468	13.200	2.933
LC31	column C3 base level	column C3 reboiler duty (qr3)	P	2	-	-
LC32	column C3 reflux drum level	toluene recycle valve (V3)	P	2	-	-

** TCVBP 1 or 3 manipulated variable: FEHE 1,2 bypass hot stream valve (VBP1 or 3) used in only CS2.*

*** TC2-2 controller used in only CS3.*

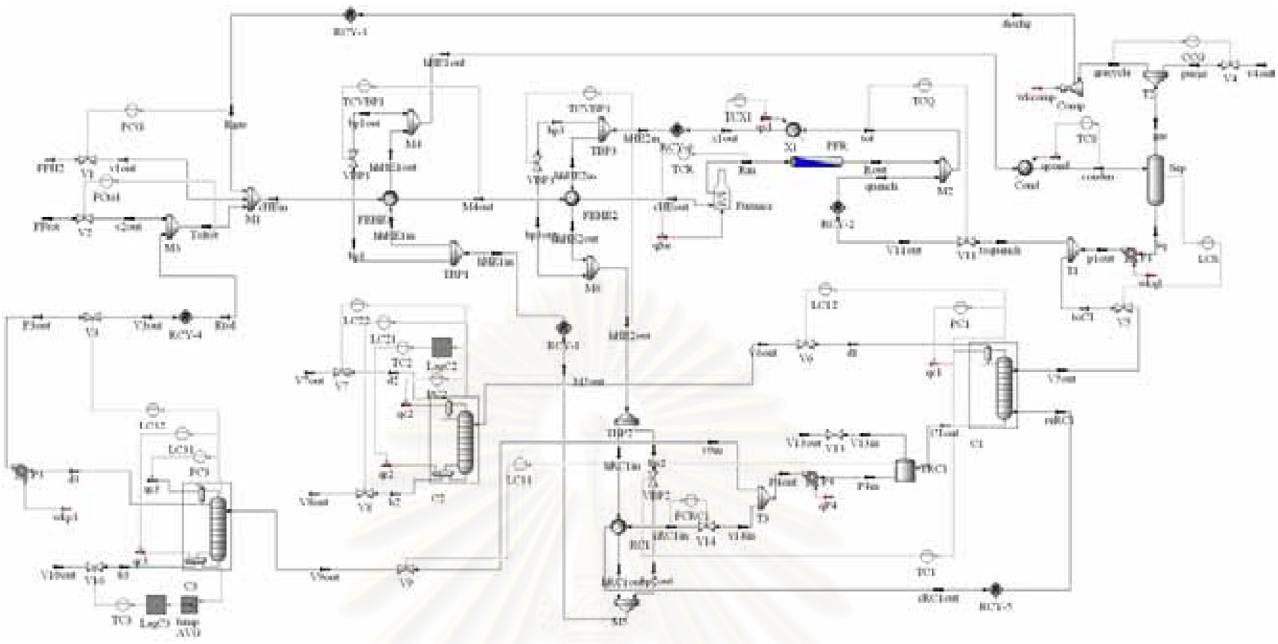


Figure 6.39 Application of control structure 2 (CS2) to the heat-integrated plant of HDA process HIP4

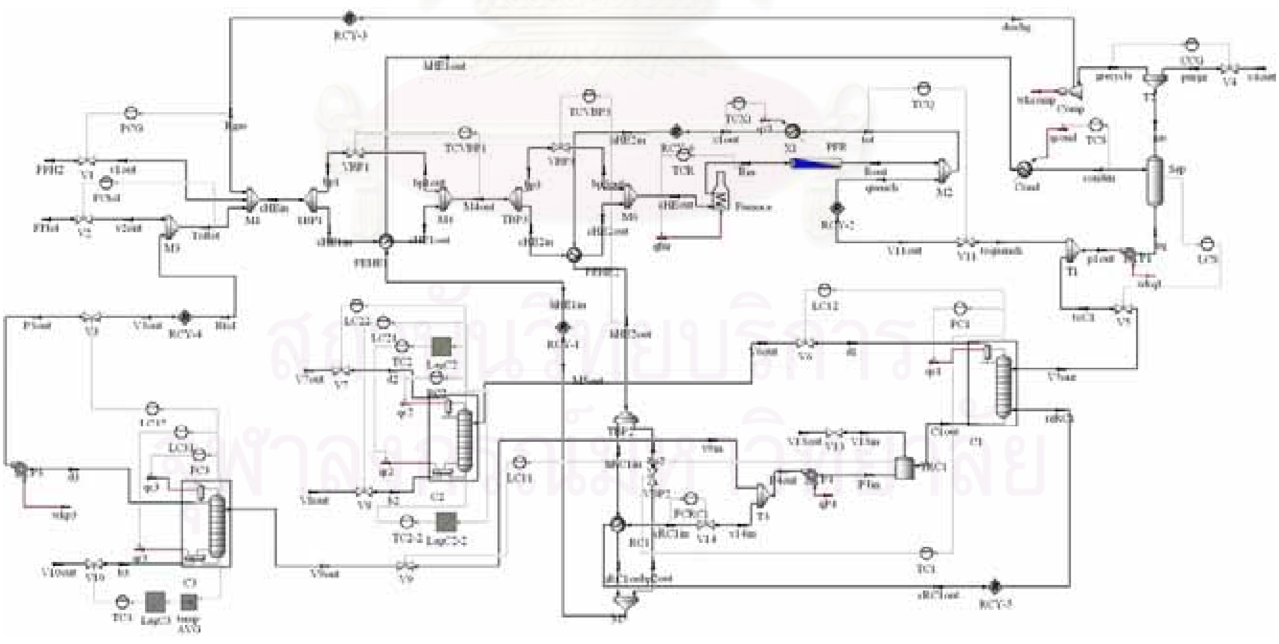


Figure 6.40 Application of control structure 3 (CS3) to the heat-integrated plant of HDA process HIP4

6.4 Dynamic Simulation Results

In order to illustrate the dynamic behaviors of new control structures, two kinds of disturbances: thermal and material disturbances are used in evaluation of the plantwide control structures. Three types of disturbance are used to test response of the system: (1) Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream), (2) Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream) and (3) Change in the Total Toluene Feed Flowrates Three disturbance loads are used to evaluate the dynamic performance of the new control structures (CS1, CS2, and CS3) for the typical and heat-integrated plant of HDA process.

Temperature controllers are PIDs which are tuned using relay feedback. Two temperature measurement lags of 0.5 minute are included in the two temperature loops (tray temperature of product column and tray temperature of recycle column). Flow and pressure controller are PIs and their parameters are heuristics values. Proportional-only level controllers are used and their parameters are heuristics values. Methane composition is measured and controlled using PI controller. All control valves are half-open at nominal operating condition.

Three control structures (CS1, CS2, and CS3) are implemented on 8 heat-integrated processes which are AL1, AL2, AL3, AL4 (typical of HDA process) and HIP1, HIP2, HIP3, HIP4 (heat-integrated plant of HDA process).

6.5 Dynamic Simulation Results for the Typical of HDA Process

In order to illustrate the dynamic behavior of the control structure in the typical of HDA process alternatives (AL1, AL2, AL3, and AL4) several disturbance loads are made. The dynamic results are explained in this part.

6.5.1 Dynamic Simulation Results for typical of HDA Process Alternative 1 (AL1)

Three disturbance loads are used to evaluate the dynamic performance of the new control structure (CS1, CS2 and CS3) for HDA process alternative 1.

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

Figure 6.40, 6.43 and 6.46 show the dynamic responses of the control systems of HDA process alternative 1 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 (Figure 6.40.a, 6.43.a, and 6.46.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 heater or cooler as follows. In the first the cold inlet temperature of FEHE is decreased and then both the cold and hot outlet temperatures of FEHE decrease suddenly. The hot outlet temperature decreasing is a desired condition, hence the control action to control the cold outlet temperature of FEHE As a result, the cold and hot outlet temperature of FEHE rapidly drops to a new steady state value (Figure 6.40.c, 6.43.c, and 6.46.c) and (Figure 6.40.e, 6.43.e, and 6.46.e) and the cooler duty decreases (Figure 6.40.l, 6.43.l, and 6.46.l) When the cold inlet temperature of FEHE increases, both the cold and hot outlet temperatures of FEHE increase. In order to the increasing cold outlet temperature is a desired condition, the control action to control the hot outlet temperature of FEHE. As a result, the cold outlet temperature of FEHE temperature quickly increases a new steady state value and the furnace duty decreases (Figure 6.40.k, 6.43.k, and 6.46.k).

The hot outlet temperature of FEHE is slightly well controlled to prevent the thermal disturbance load propagation to furnace utility. This disturbance load is shifted to the cooler utility. The cooler duty will be increased in this case

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

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

Figure 6.41, 6.44 and 6.47 shows the dynamic responses of the control systems of HDA process alternative1 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 FEHE-hot-inlet temperature controller (i.e. TCX1) is decreased from 621.1 to 611.1^{o}C at time equals 10 minutes and the temperature is increased from 611.1 to 631.1^{o}C at time equals 100 minutes, then its temperature is returned to its nominal value of 621.1^{o}C at time equals 200 minutes (Figure 6.41.a, 6.44.a and 6.47.a). As can be seen, this temperature response is very fast, the new steady state is reached quickly

Since the hot outlet temperature of FEHE is controlled (Figure 6.41.b, 6.44.b and 6.47.b) 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, this disturbance load is shifted to the cooler utility. The cooler duty will be decreased in this case (Figure 6.41.l, 6.44.l and 6.47.l). Consider the case when the hot inlet temperature of FEHE increases, this is a desired condition to shift the disturbance load to the cold stream. Therefore, the cooler duty increases to a new steady state value.

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 CS1 is smoother than CS2 and CS3.

6.5.1.3 Change in the Total Toluene Feed Flowrates

Figure 6.42, 6.45 and 6.48 shows the dynamic responses of the control systems of HDA process alternative 1 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 minutes (Figure 6.42.a, 6.45.a and 6.48.a).

The dynamic result can be seen that the drop in total toluene feed flowrates reduces the reaction rate, so fresh feed hydrogen flowrates (Figure 6.42.h, 6.45.h and 6.48.h). and the benzene product flowrates drops (Figure 6.42.j, 6.45.j and 6.48.j).and the benzene product quality increases (Figure 6.42.i, 6.45.i and 6.48.i). Consider the case when the total toluene feed flowrates increase and the benzene product flowrates increase because of the reaction rate enlargement. The benzene product quality will increase in this case. The deviation of benzene product quality from nominal value in CS1, CS2 and CS3 are slightly similar when total toluene feed flowrates change.

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

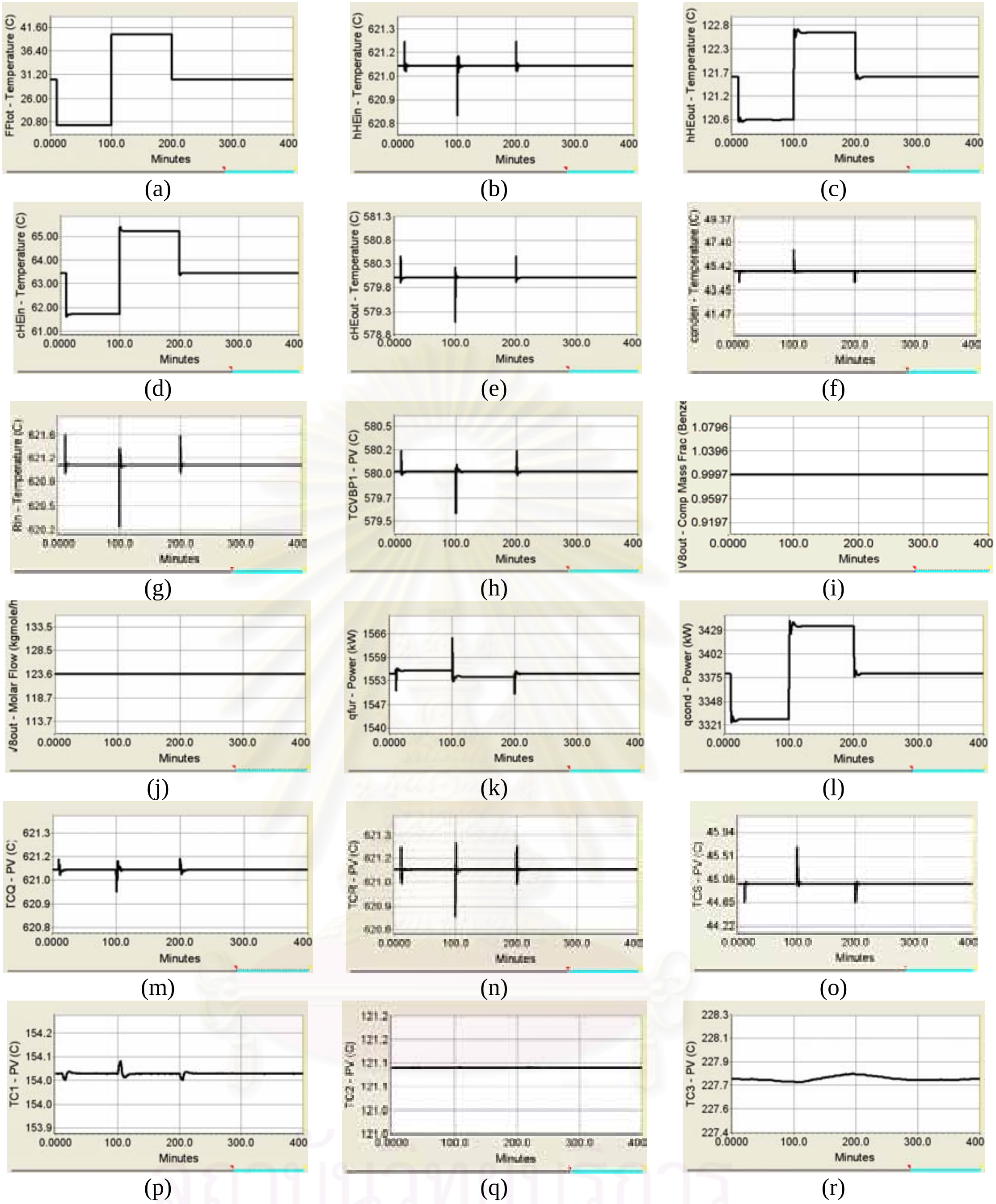


Figure 6.40 Dynamic Responses of the HDA Process Alternative 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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

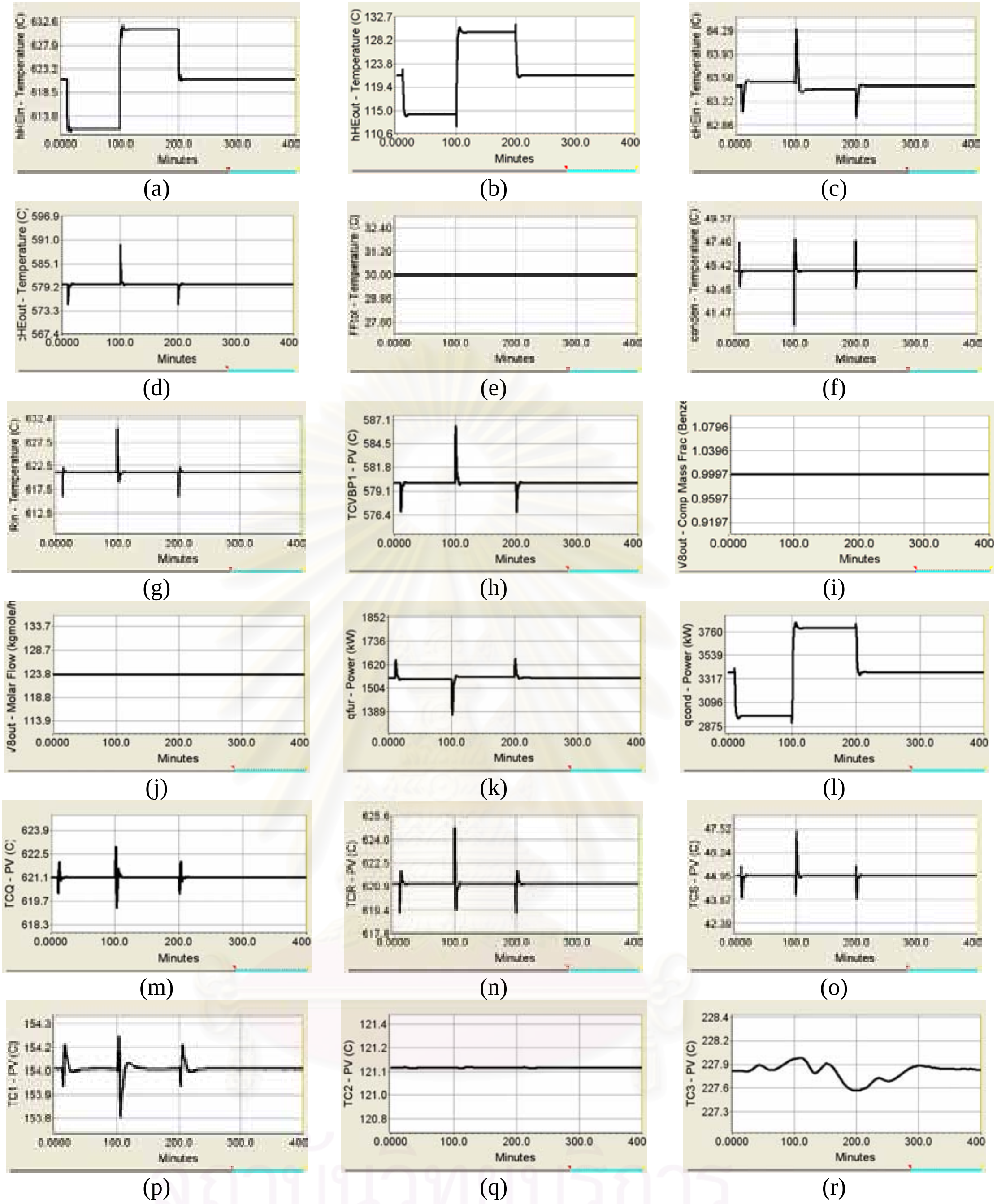


Figure 6.41 Dynamic Responses of the HDA Process Alternative 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

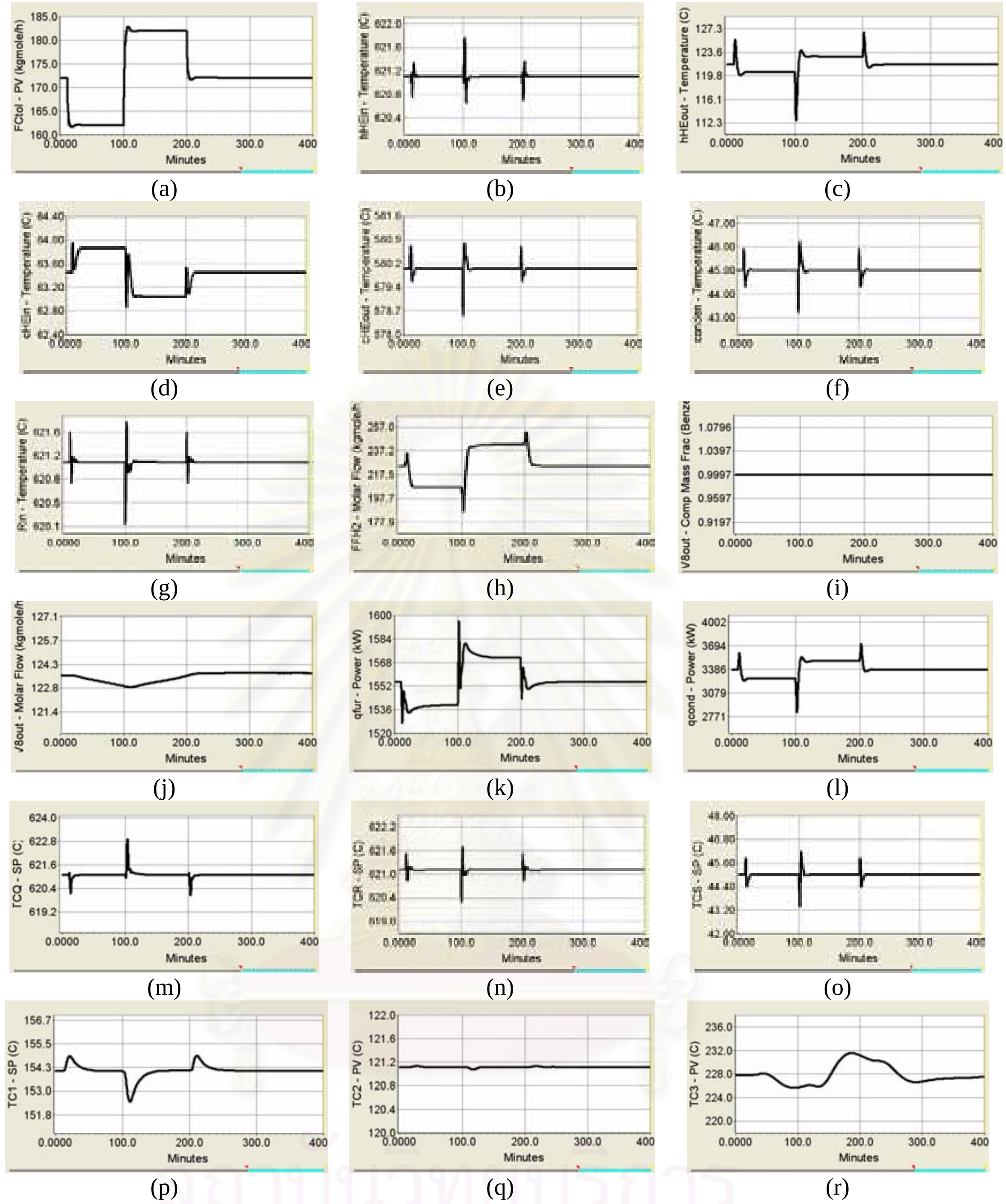
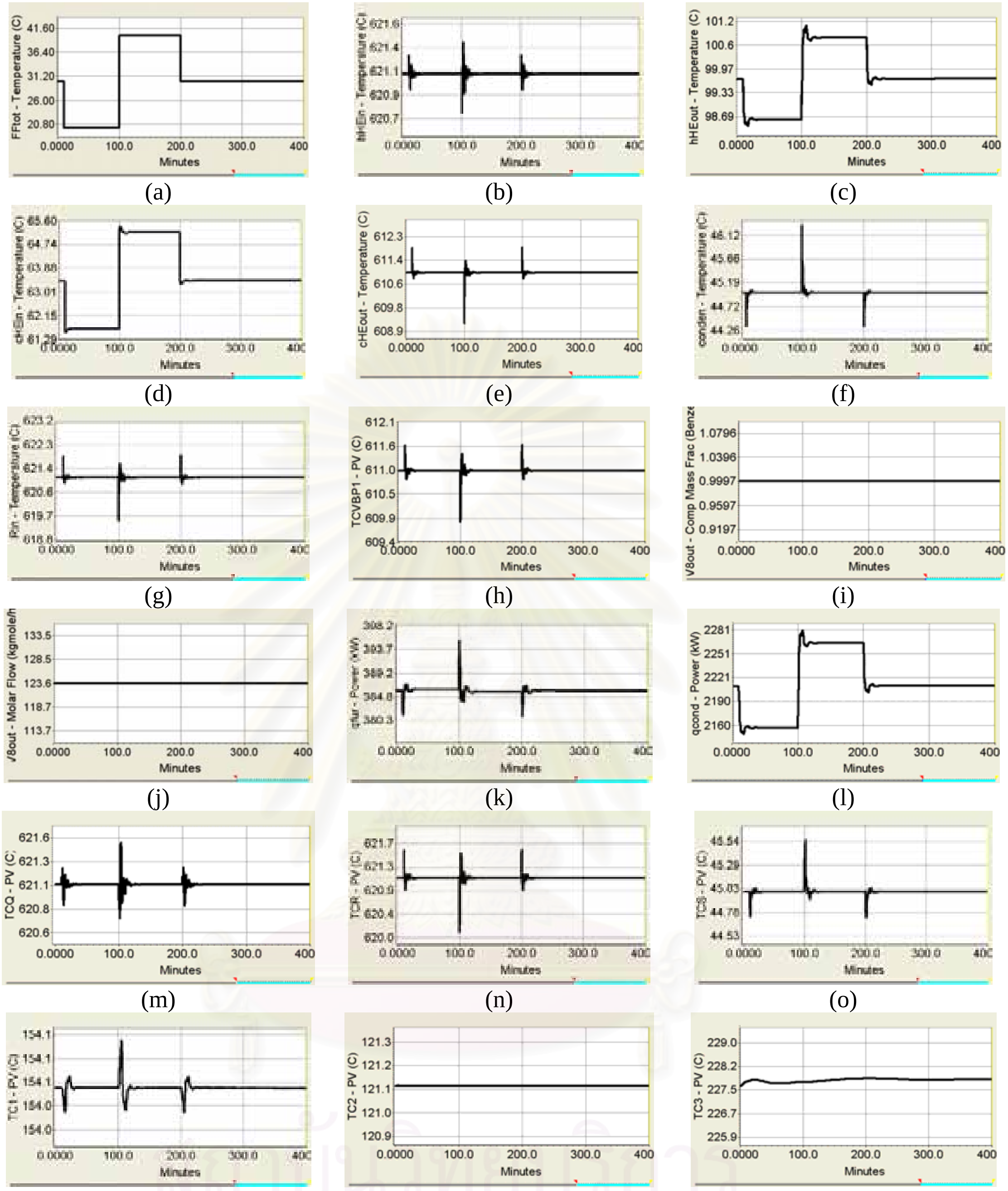


Figure 6.42 Dynamic Responses of the HDA Process Alternative 1 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.



(a) (b) (c) (d) (e) (f) (g) (h) (i) (j) (k) (l) (m) (n) (o) (p) (q) (r)

Figure 6.43 Dynamic Responses of the HDA Process Alternative 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

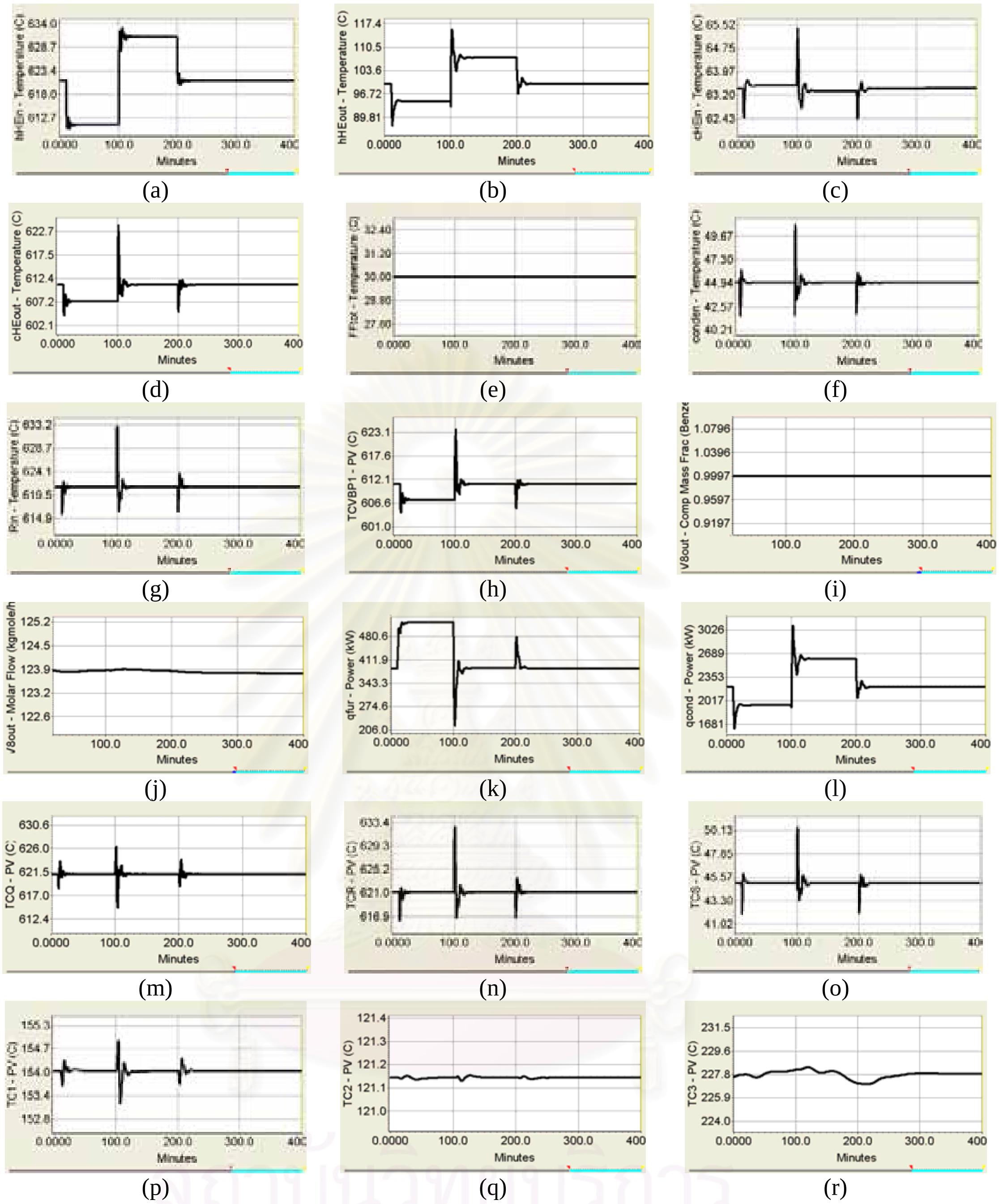


Figure 6.44 Dynamic Responses of the HDA Process Alternative 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

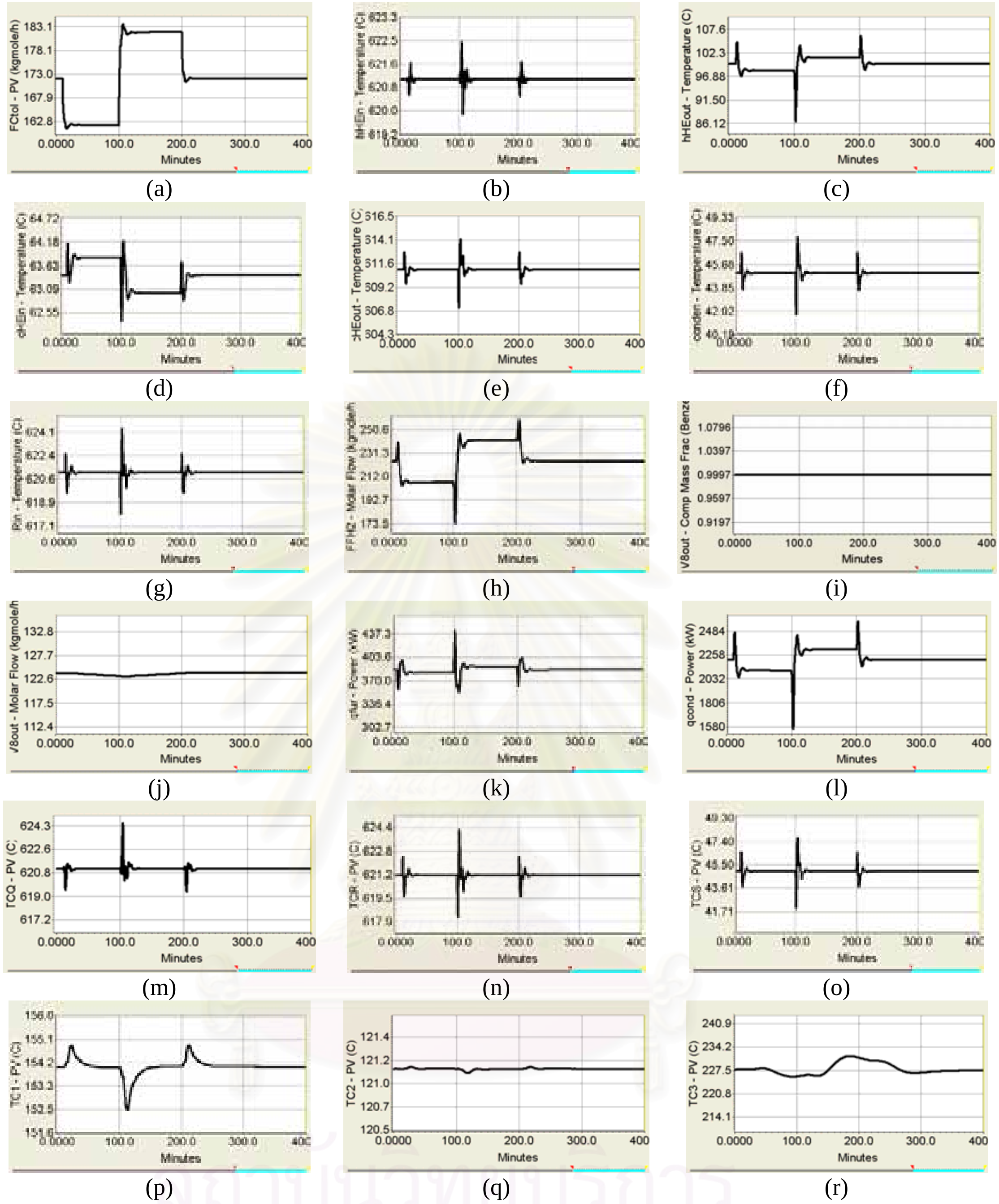


Figure 6.45 Dynamic Responses of the HDA Process Alternative 1 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

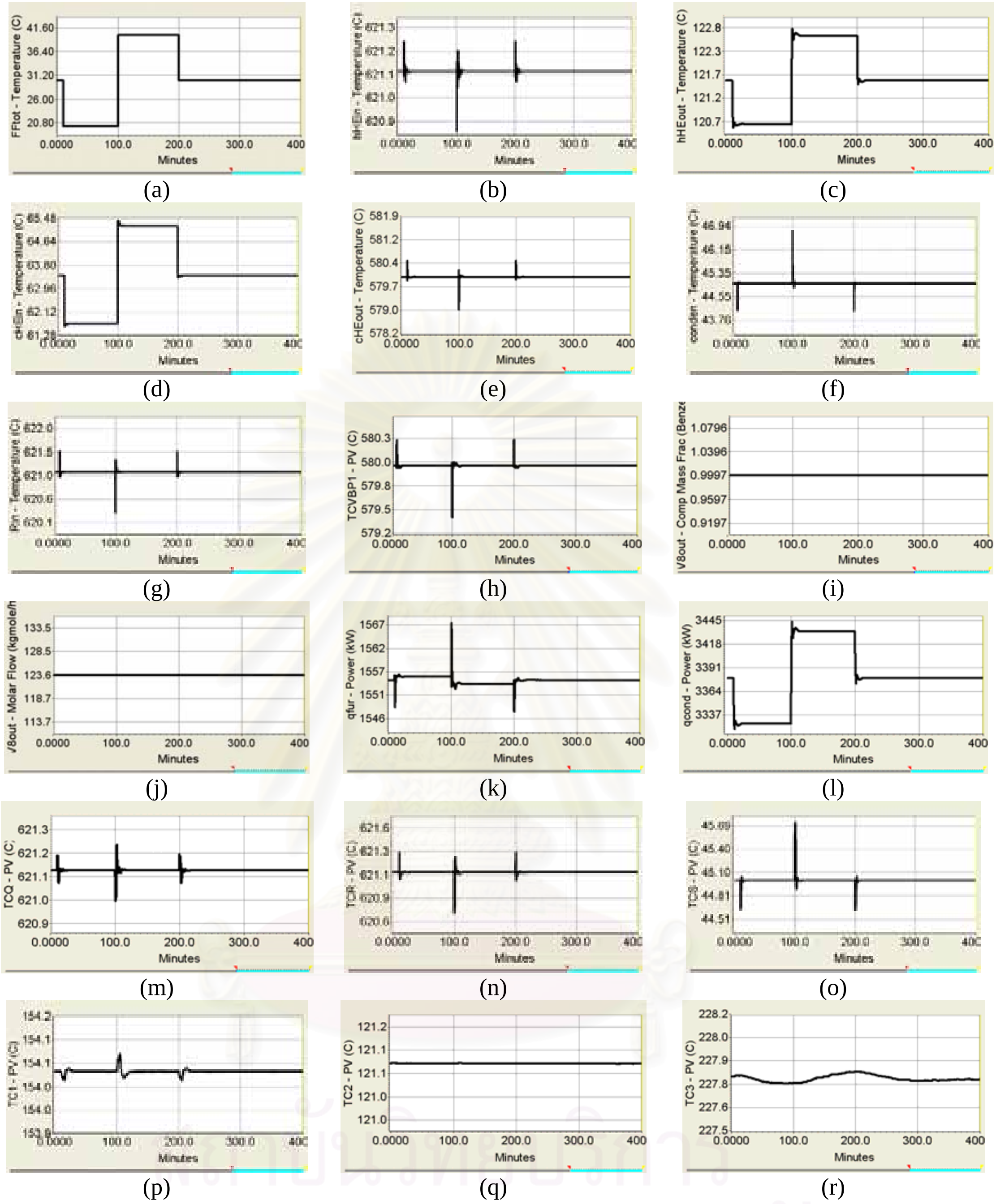


Figure 6.46 Dynamic Responses of the HDA Process Alternative 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

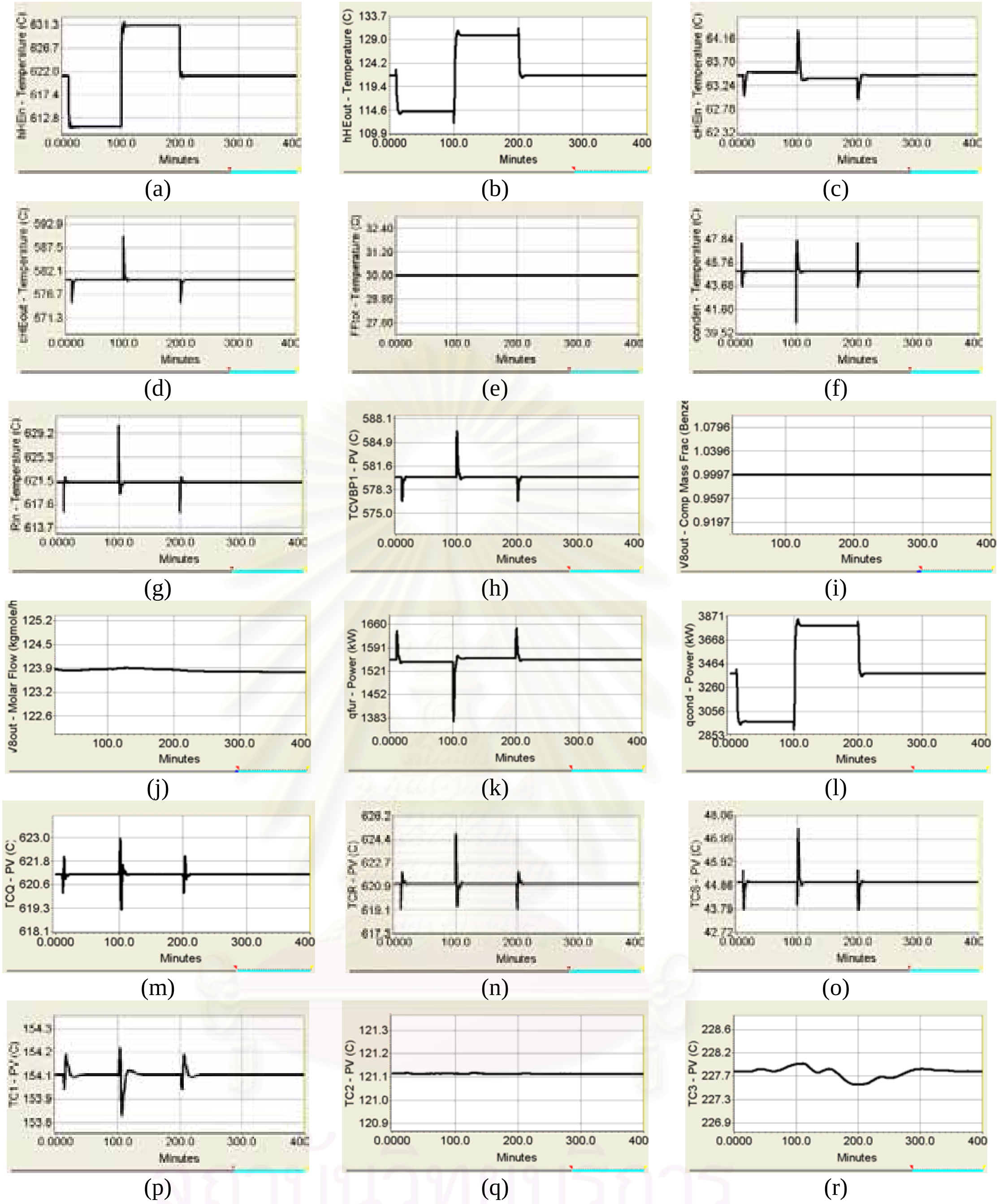


Figure 6.47 Dynamic Responses of the HDA Process Alternative 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

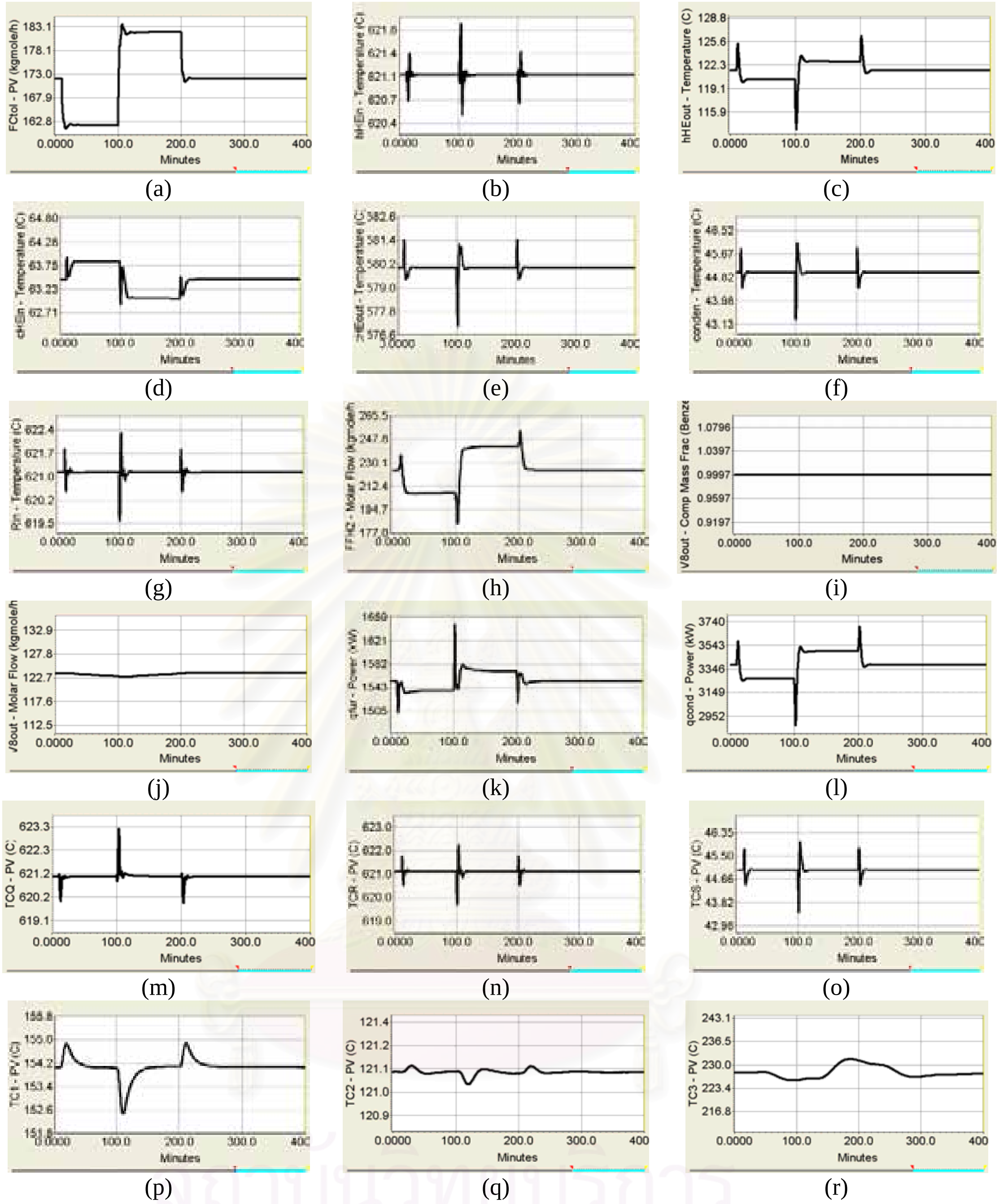


Figure 6.48 Dynamic Responses of the HDA Process Alternative 1 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

6.5.2 Dynamic Simulation Results for Typical of HDA Process Alternative 2 (AL2)

Three disturbance loads are also used to evaluate the dynamic performance of the new control structure (CS1, CS2 and CS3) for HDA process alternative 2, (1) Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream) show the dynamic responses on Figure 6.49, 6.52 and 6.55, (2) Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream) show the dynamic responses on Figure 6.50, 6.53 and 6.56, (3) Change in the Total Toluene Feed Flowrates show the dynamic responses on Figure 6.51, 6.54 and 6.57, As can be see, the dynamic response for HDA process alternative 2 are same in CS1, CS2 and CS3.

6.5.3 Dynamic Simulation Results for Typical of HDA Process Alternative 3 (AL3)

Three disturbance loads are also used to evaluate the dynamic performance of the new control structure (CS1, CS2 and CS3) for HDA process alternative 3, (1) Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream) show the dynamic responses on Figure 6.58, 6.61 and 6.64, (2) Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream) show the dynamic responses on Figure 6.59, 6.62 and 6.65, (3) Change in the Total Toluene Feed Flowrates show the dynamic responses on Figure 6.60, 6.63 and 6.66, As can be see, the dynamic response for HDA process alternative 3 are same in CS1, CS2 and CS3.

6.5.4 Dynamic Simulation Results for Typical of HDA Process Alternative 4 (AL4)

Three disturbance loads are also used to evaluate the dynamic performance of the new control structure (CS1, CS2 and CS3) for HDA process alternative 4, (1) Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream) show the dynamic responses on Figure 6.67, 6.70 and 6.73, (2) Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream) show the dynamic responses on Figure 6.68, 6.71 and 6.74, (3) Change in the Total Toluene Feed Flowrates show the dynamic responses on Figure 6.69, 6.72 and 6.75, As can be see, the dynamic response for HDA process alternative 4 are same in CS1, CS2 and CS3.

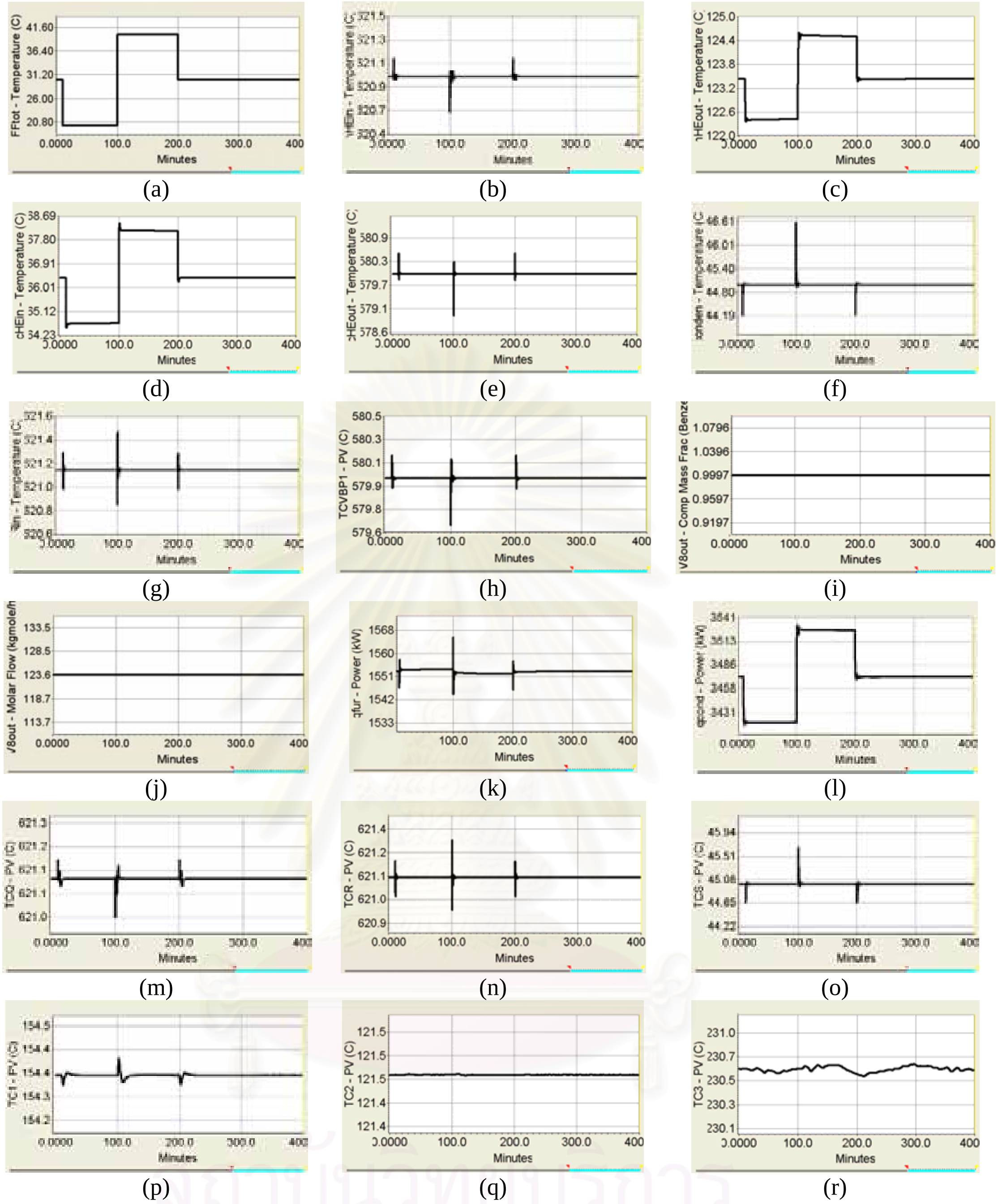


Figure 6.49 Dynamic Responses of the HDA Process Alternative 2 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

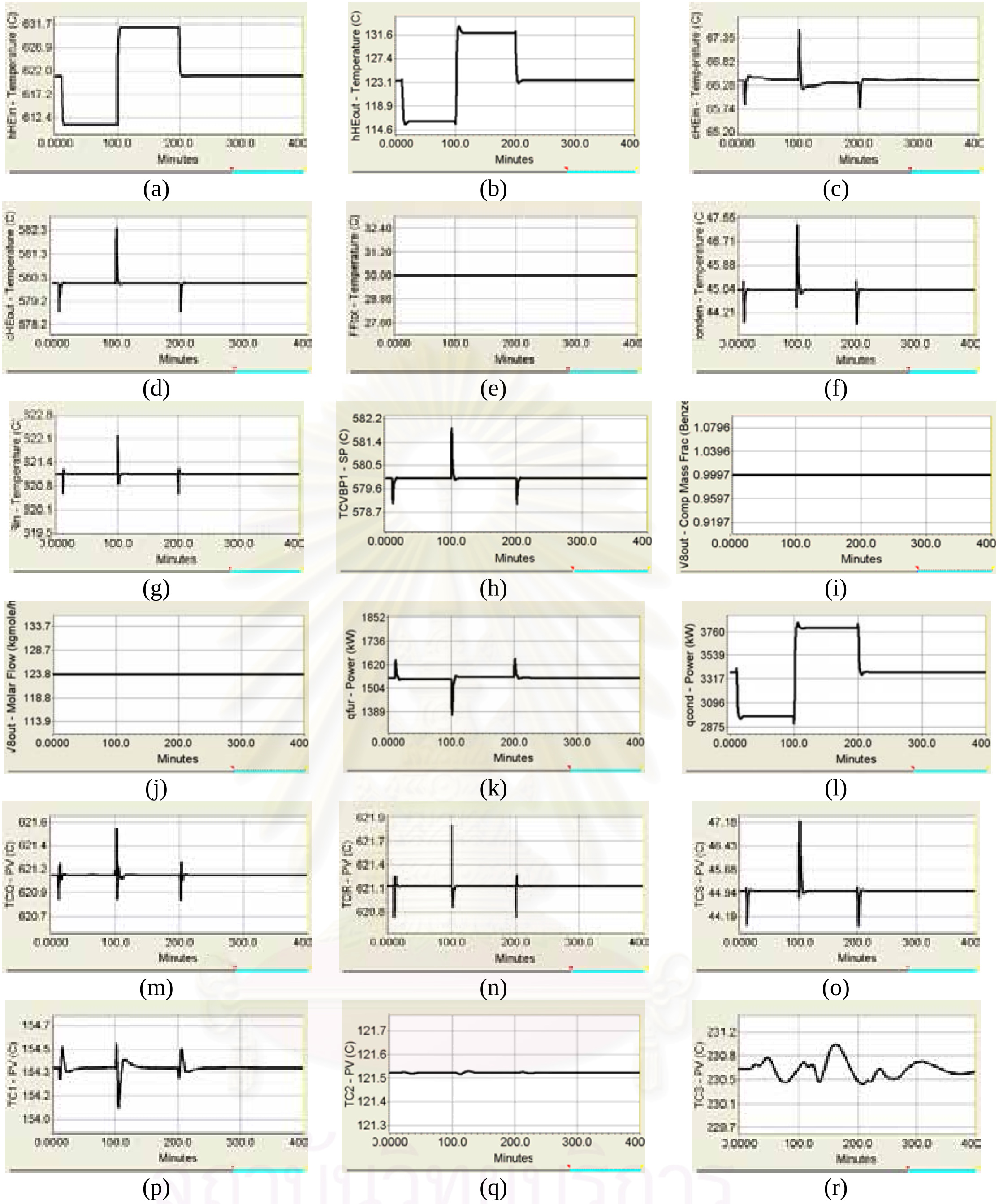


Figure 6.50 Dynamic Responses of the HDA Process Alternative 2 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

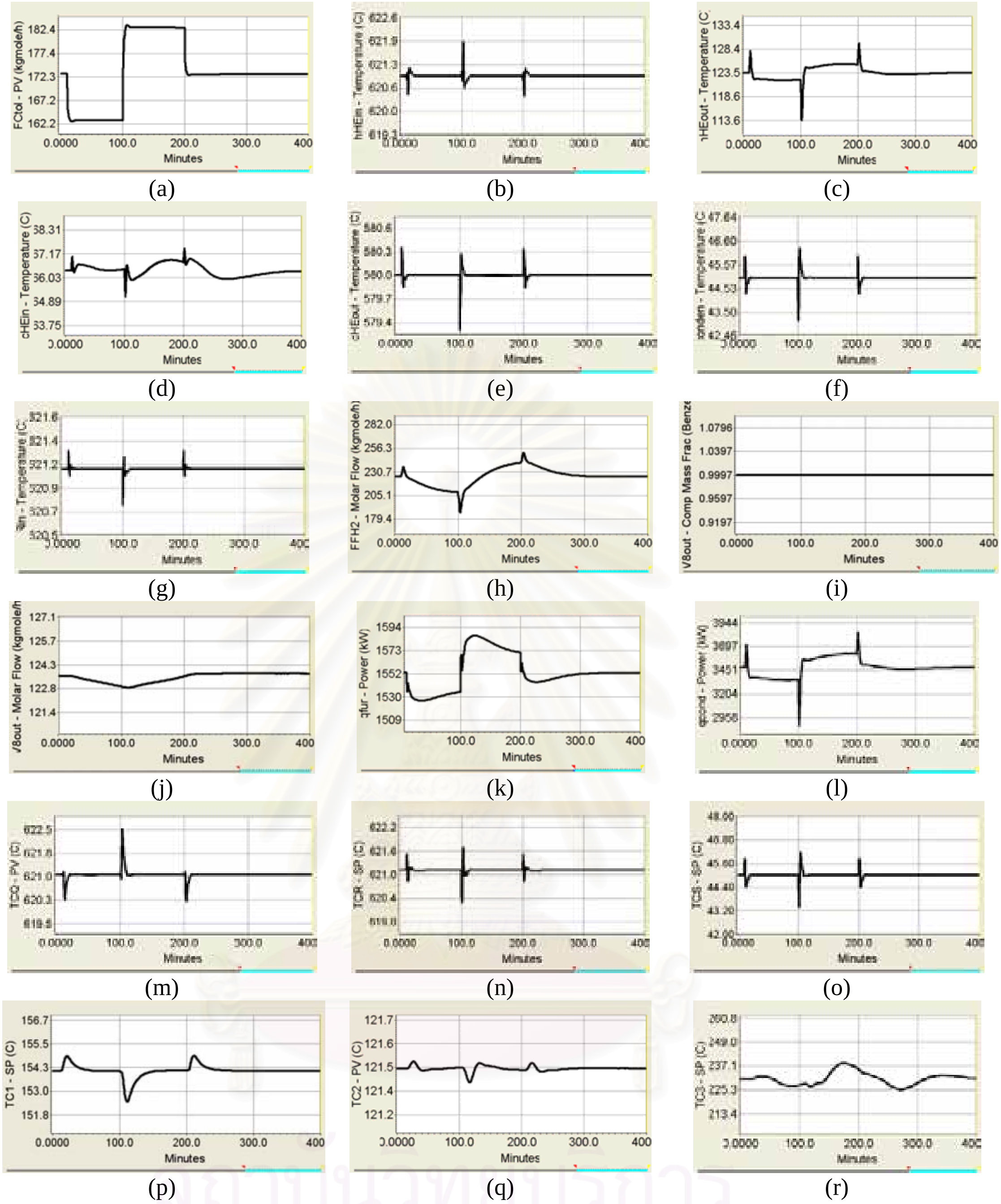


Figure 6.51 Dynamic Responses of the HDA Process Alternative 2 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

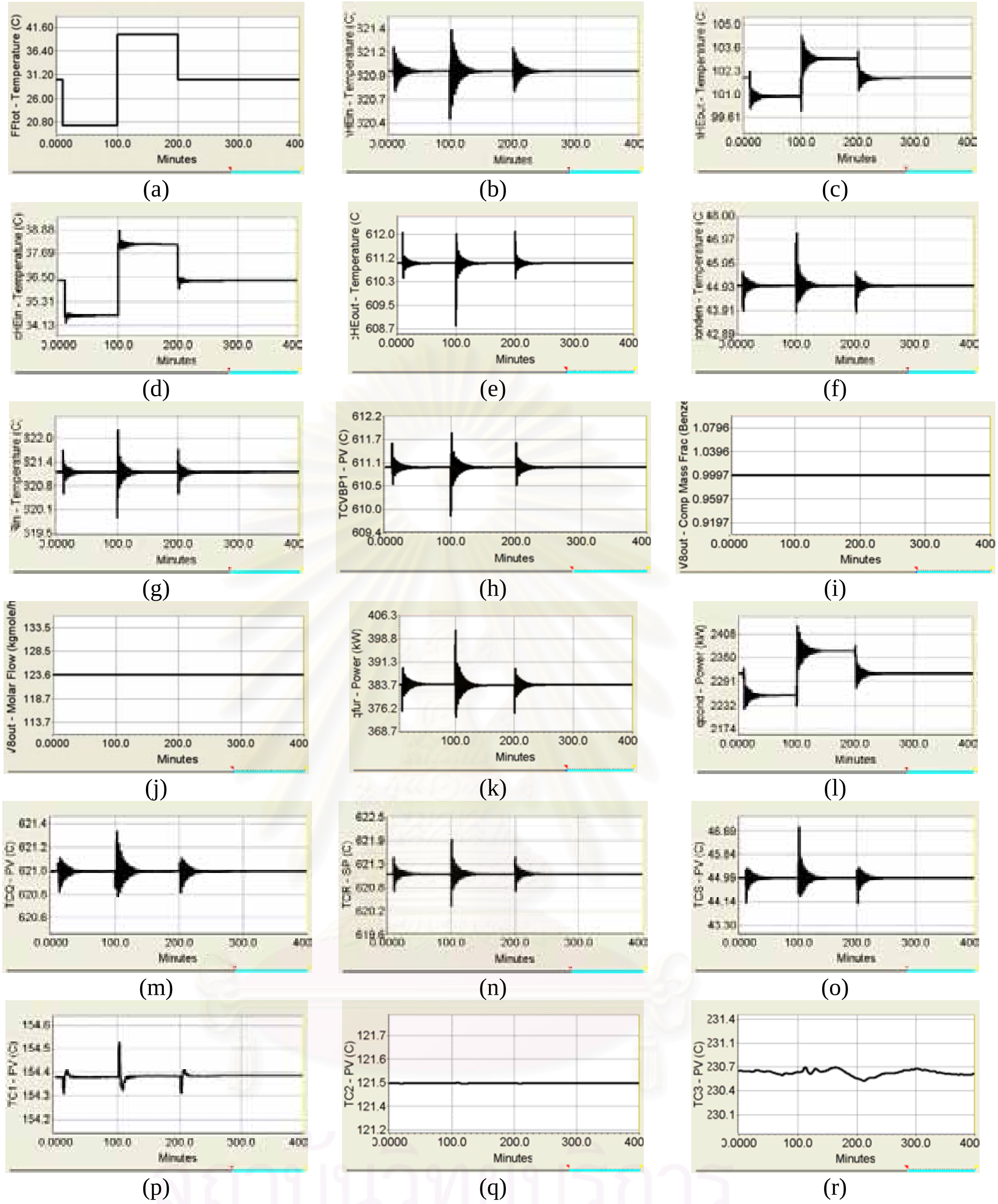


Figure 6.52 Dynamic Responses of the HDA Process Alternative 2 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

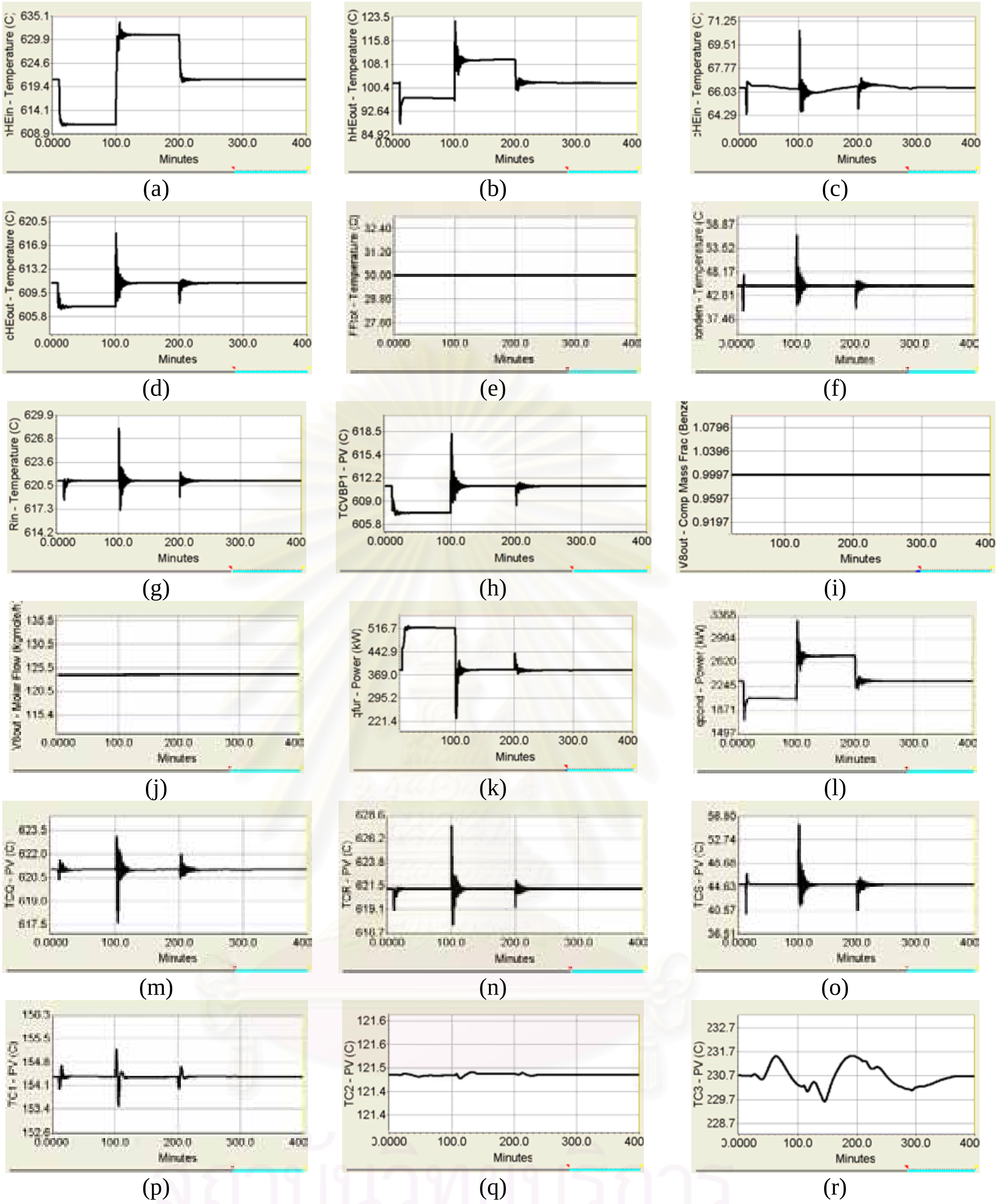


Figure 6.53 Dynamic Responses of the HDA Process Alternative 2 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

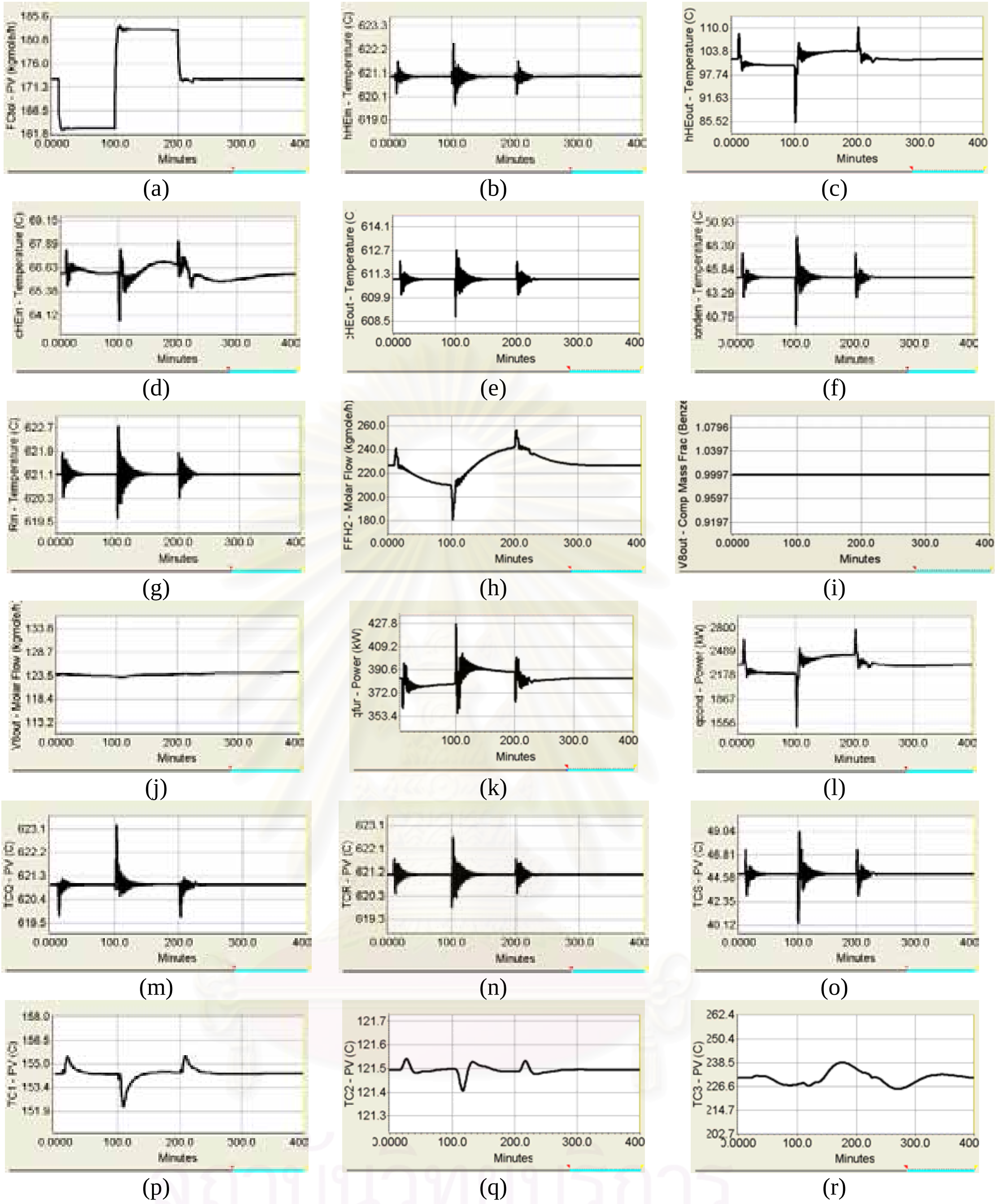


Figure 6.54 Dynamic Responses of the HDA Process Alternative 2 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

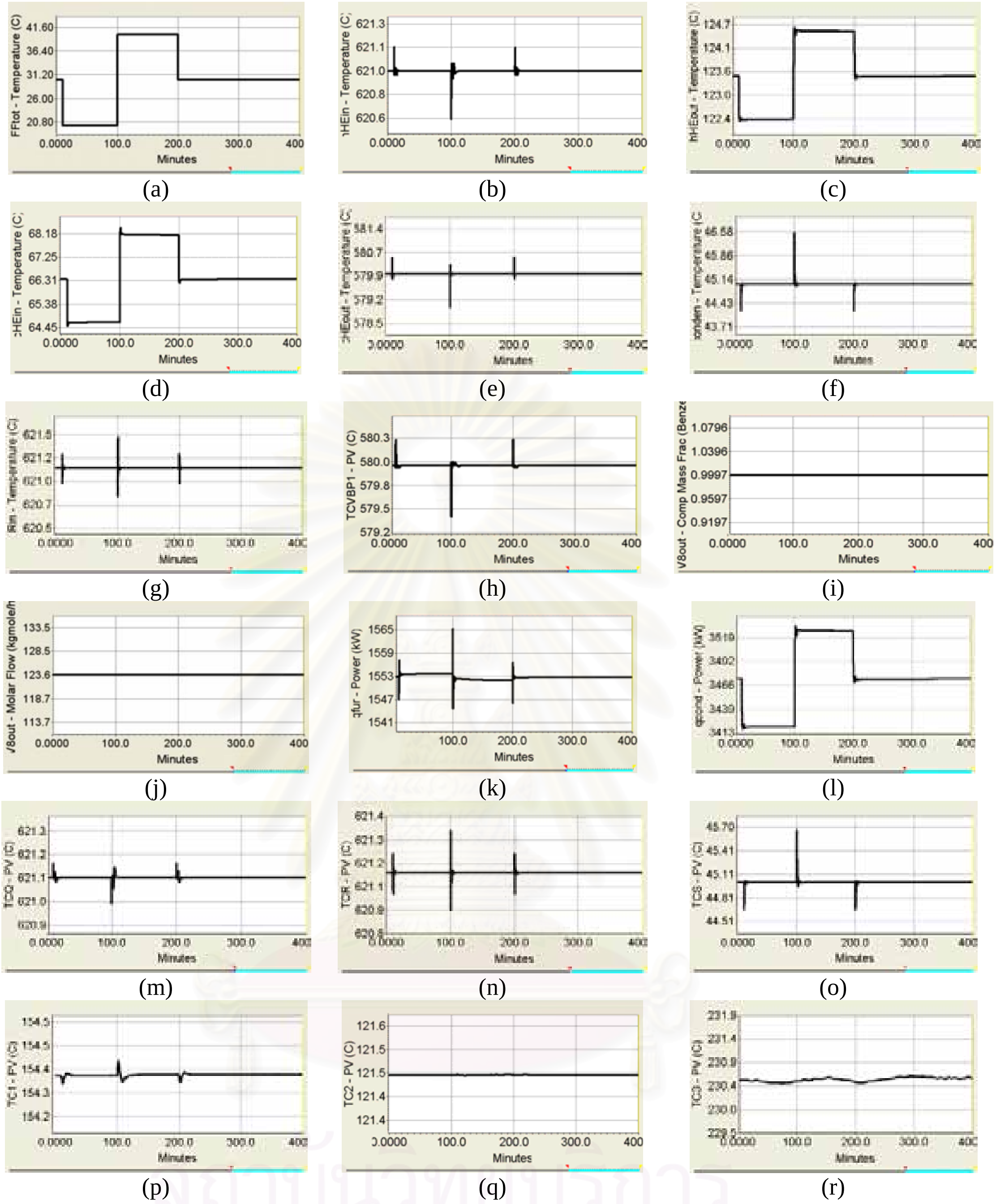


Figure 6.55 Dynamic Responses of the HDA Process Alternative 2 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

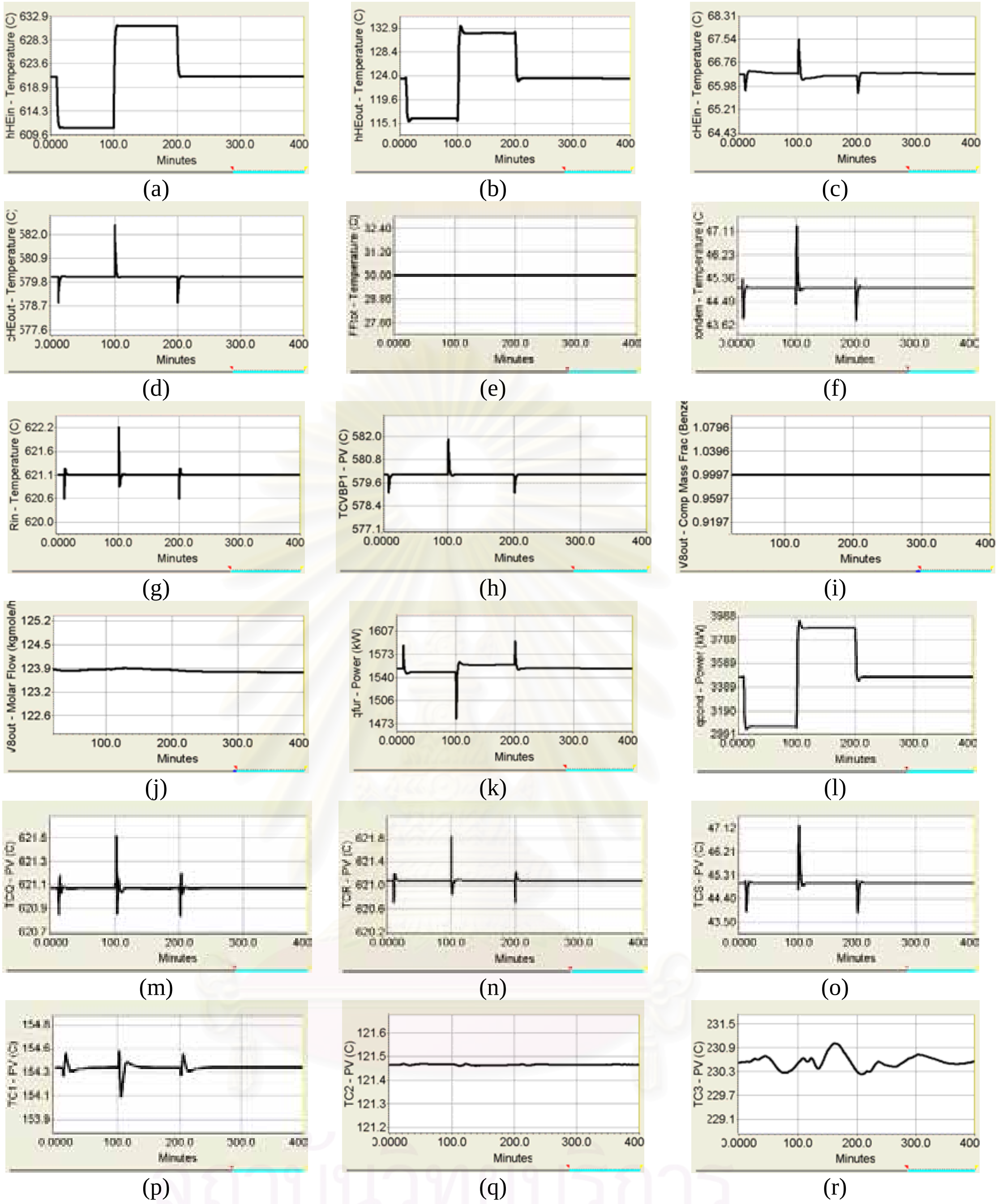


Figure 6.56 Dynamic Responses of the HDA Process Alternative 2 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

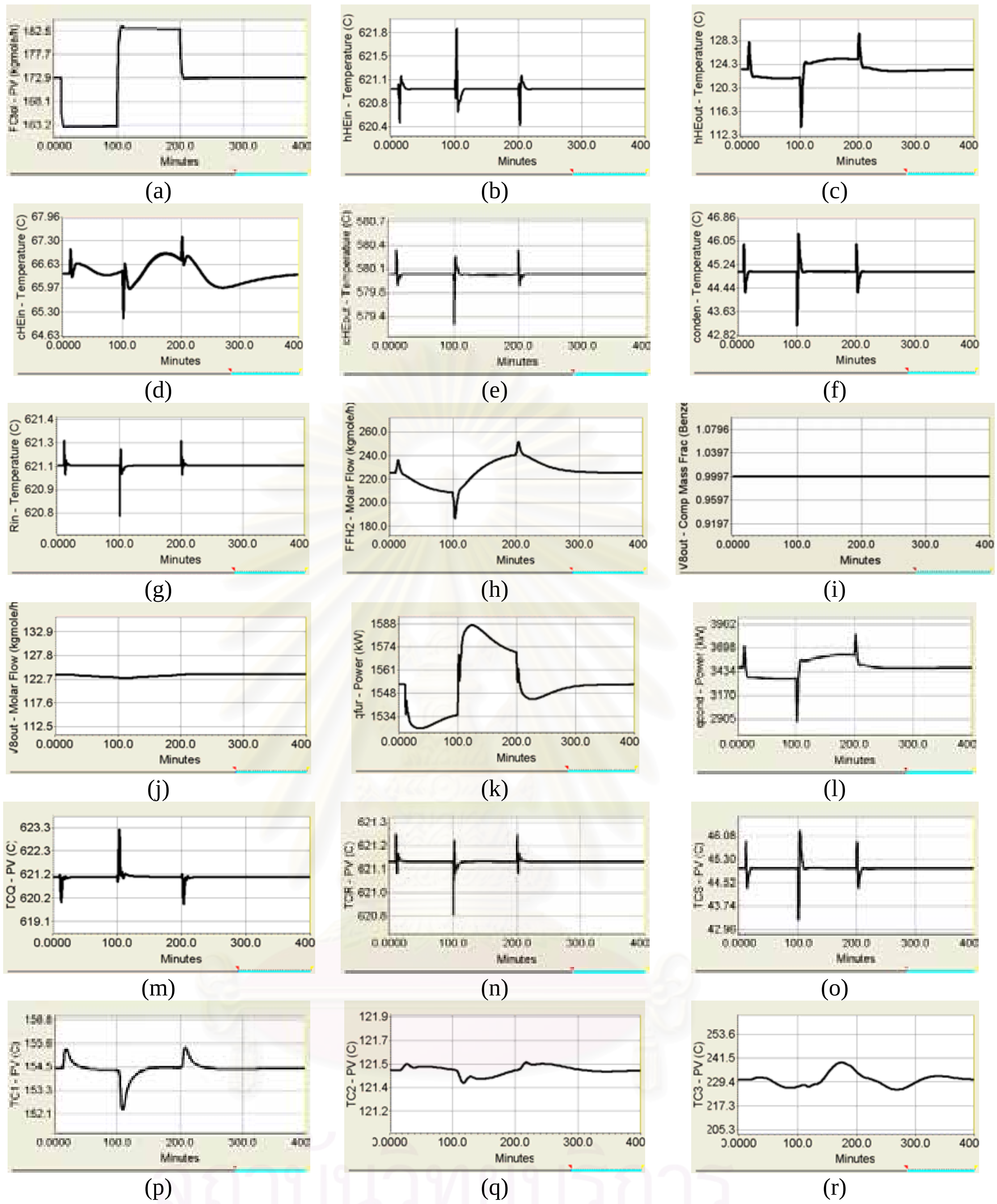


Figure 6.57 Dynamic Responses of the HDA Process Alternative 2 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

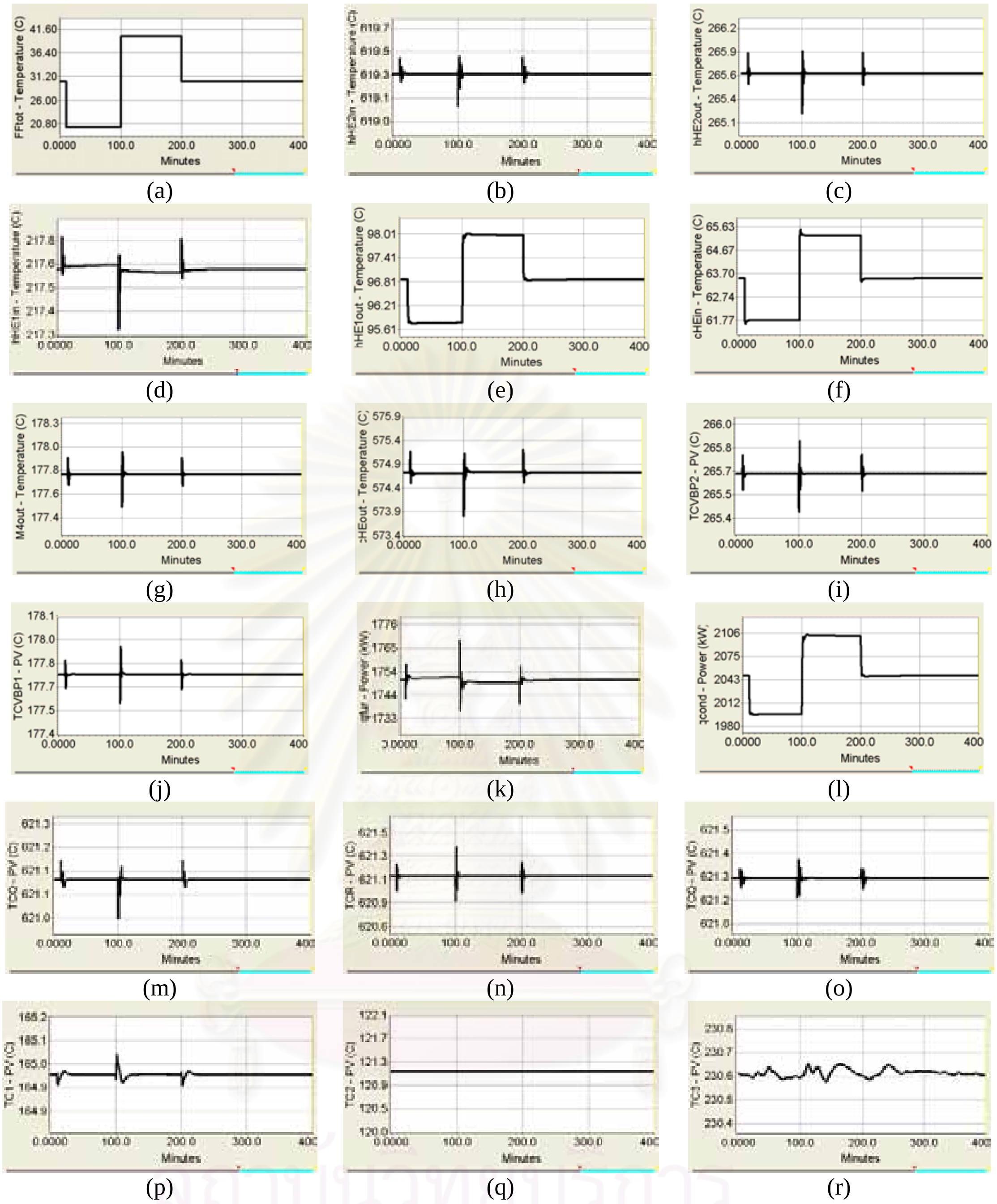


Figure 6.58 Dynamic Responses of the HDA Process Alternative 3 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

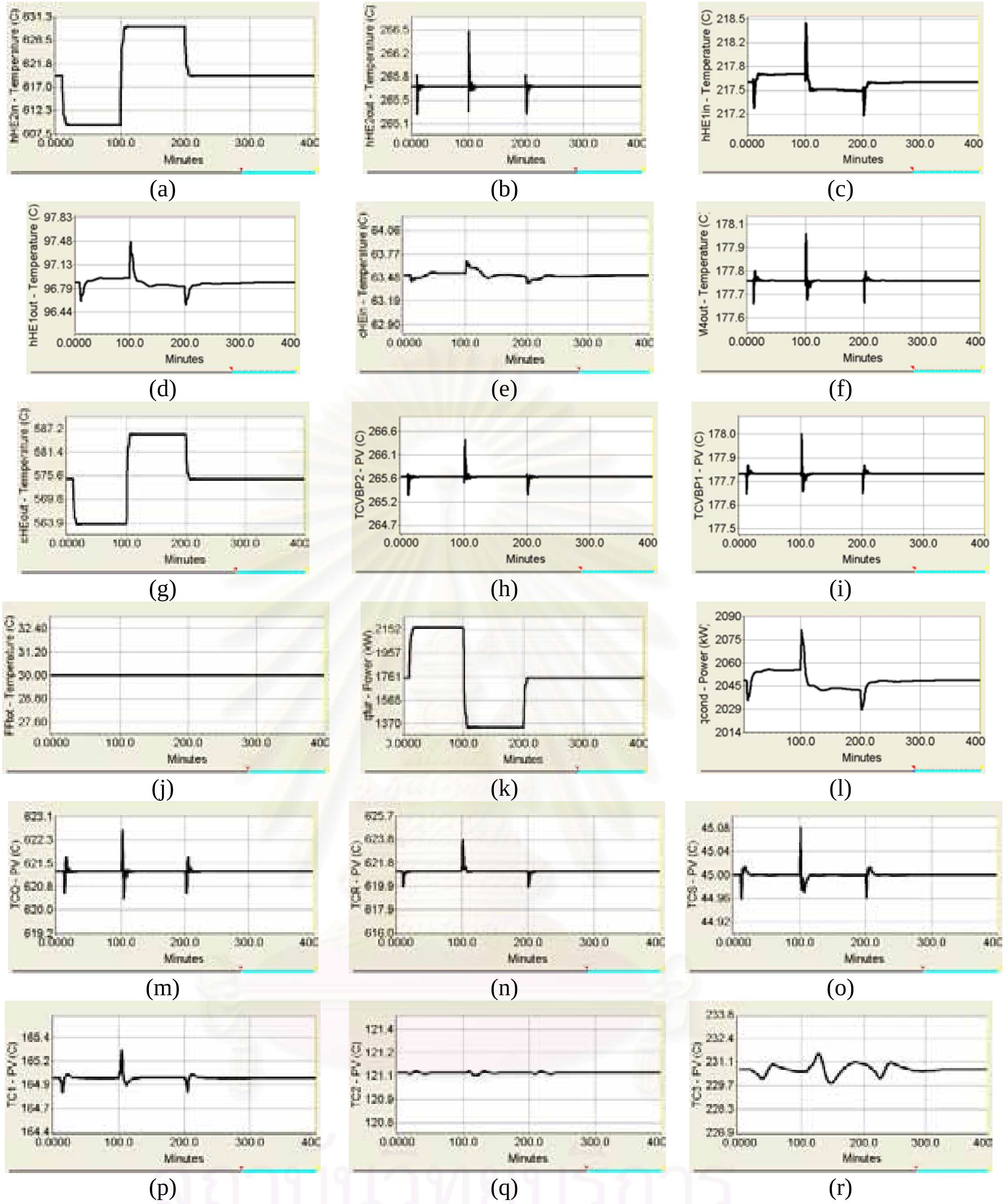


Figure 6.59 Dynamic Responses of the HDA Process Alternative 3 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

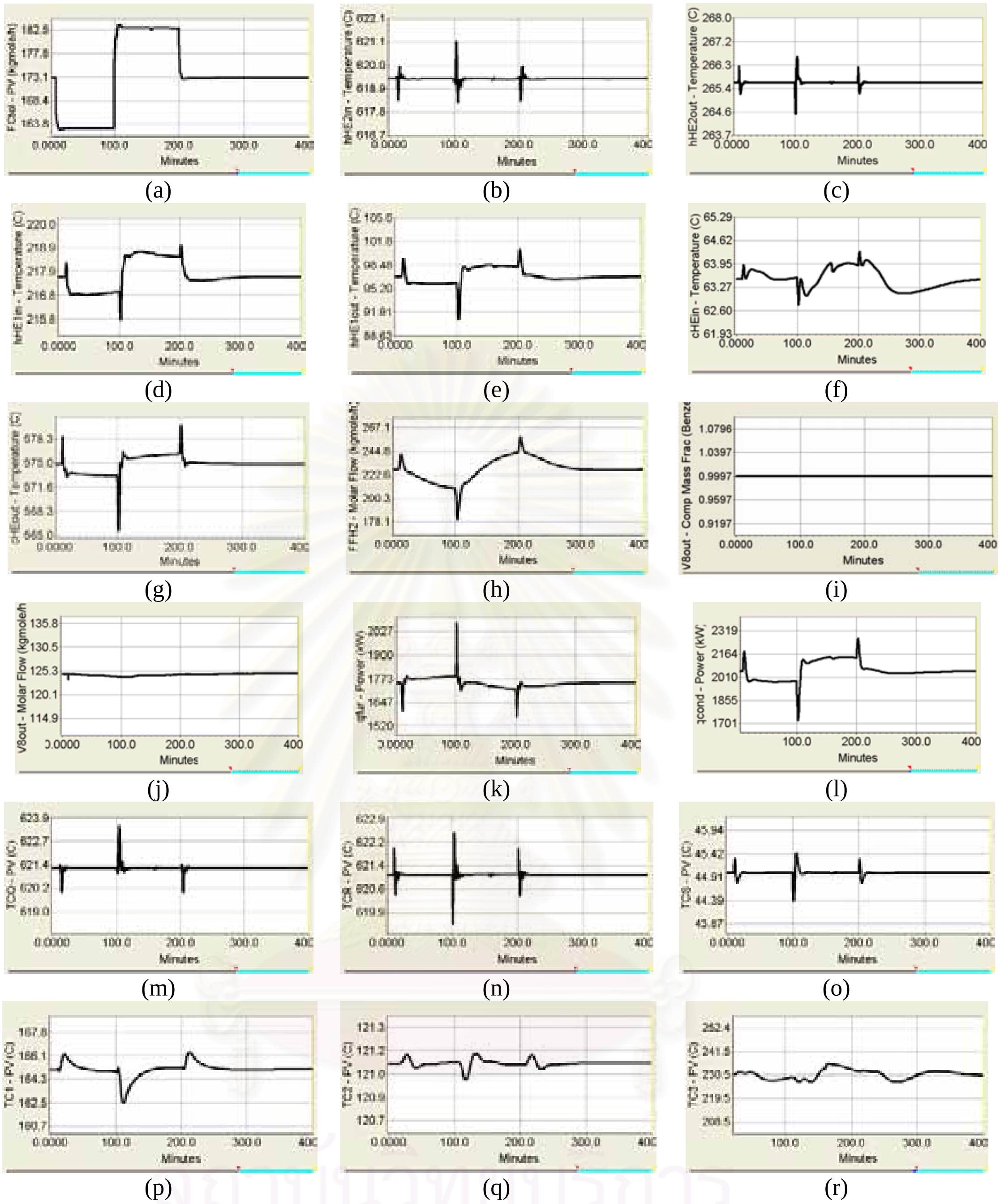


Figure 6.60 Dynamic Responses of the HDA Process Alternative 3 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

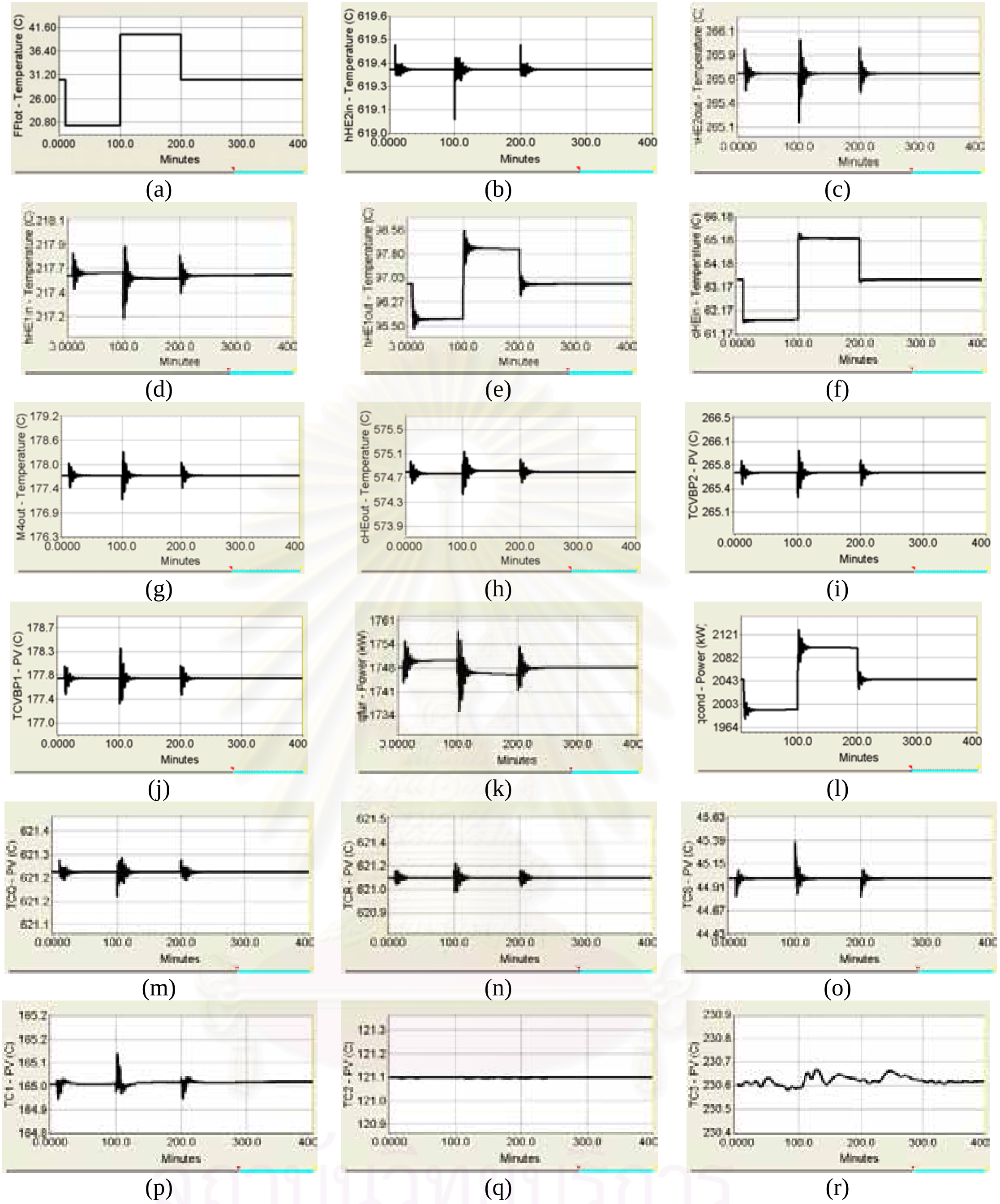


Figure 6.61 Dynamic Responses of the HDA Process Alternative 3 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

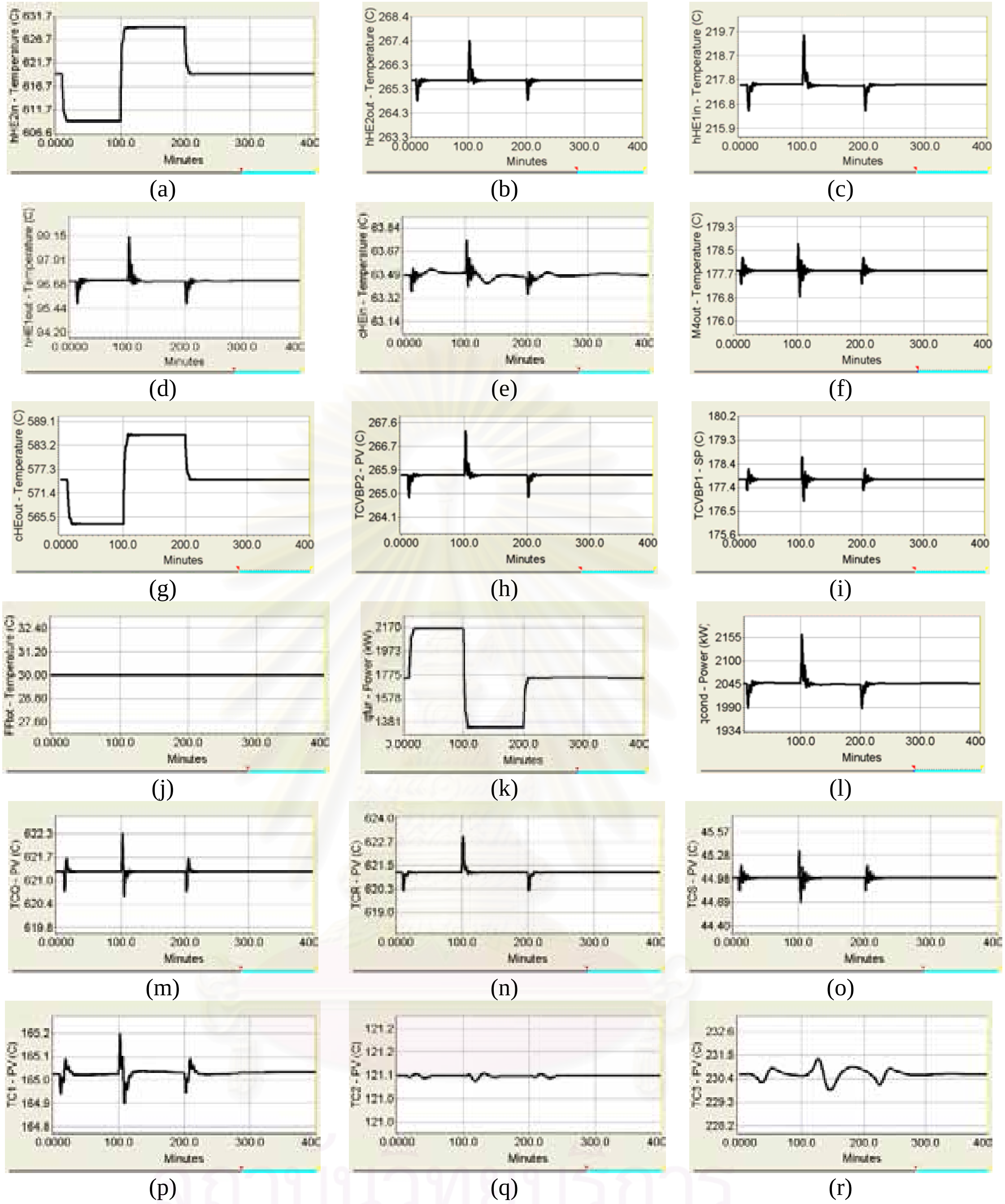


Figure 6.62 Dynamic Responses of the HDA Process Alternative 3 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

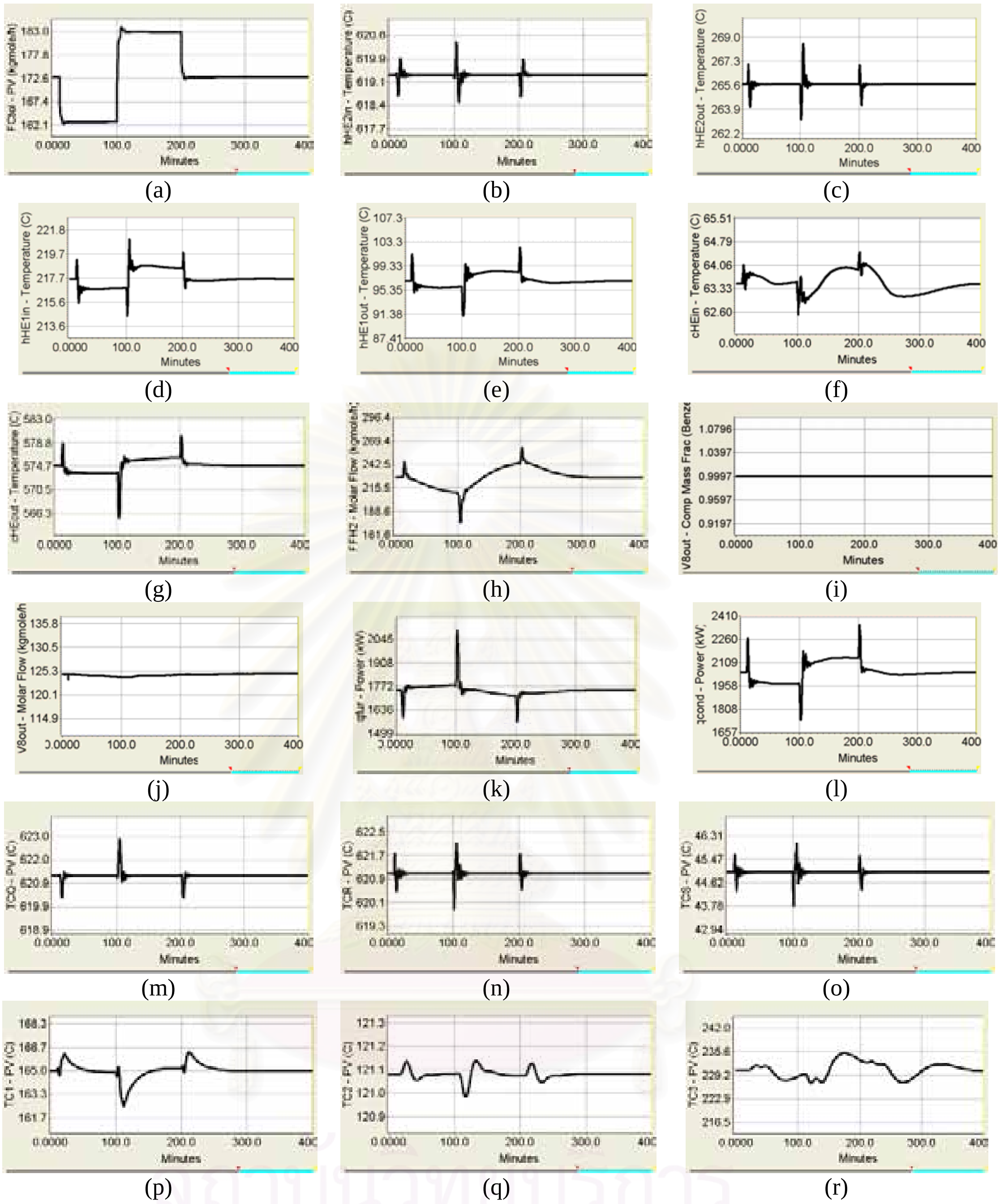


Figure 6.63 Dynamic Responses of the HDA Process Alternative 3 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

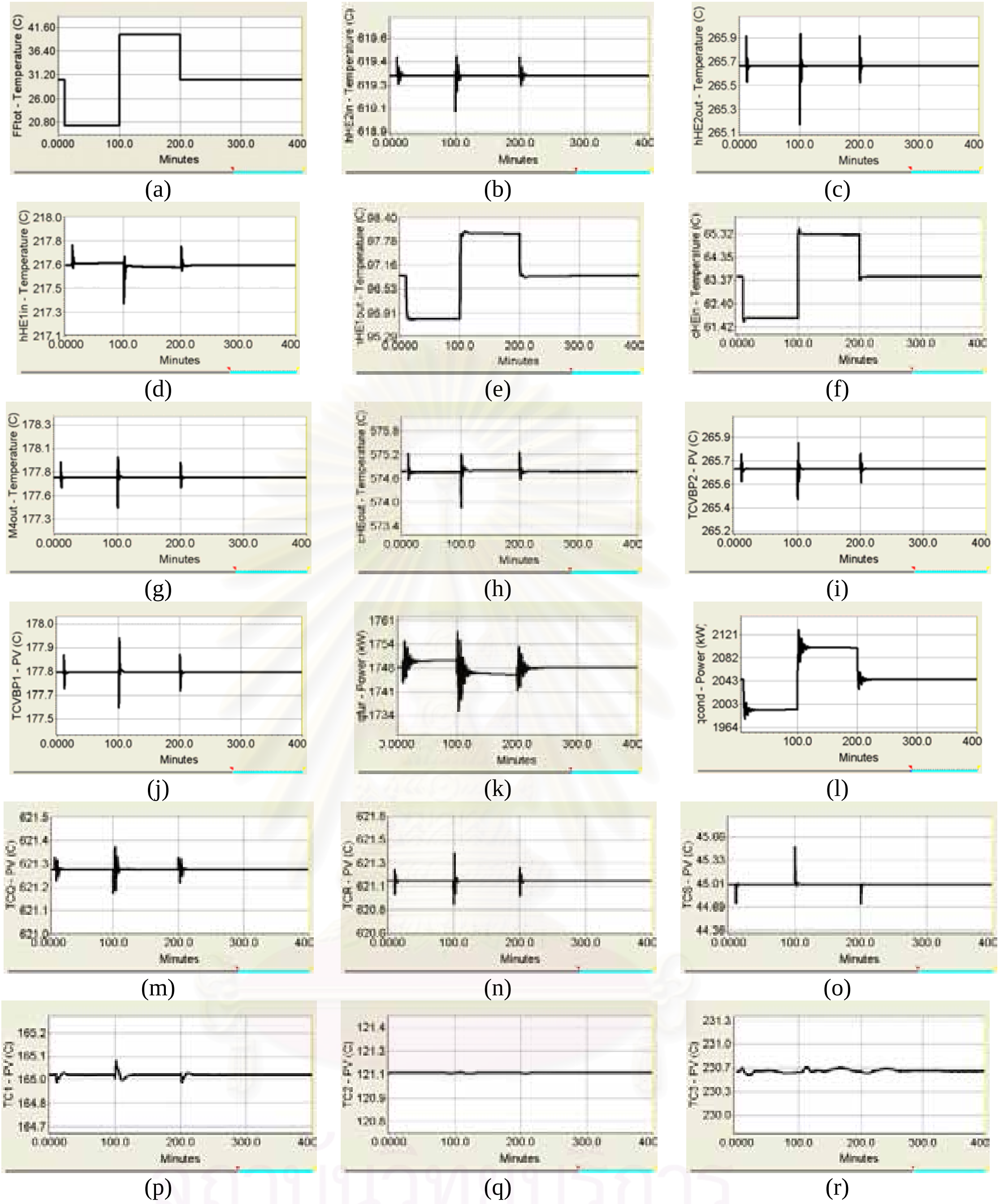


Figure 6.64 Dynamic Responses of the HDA Process Alternative 3 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

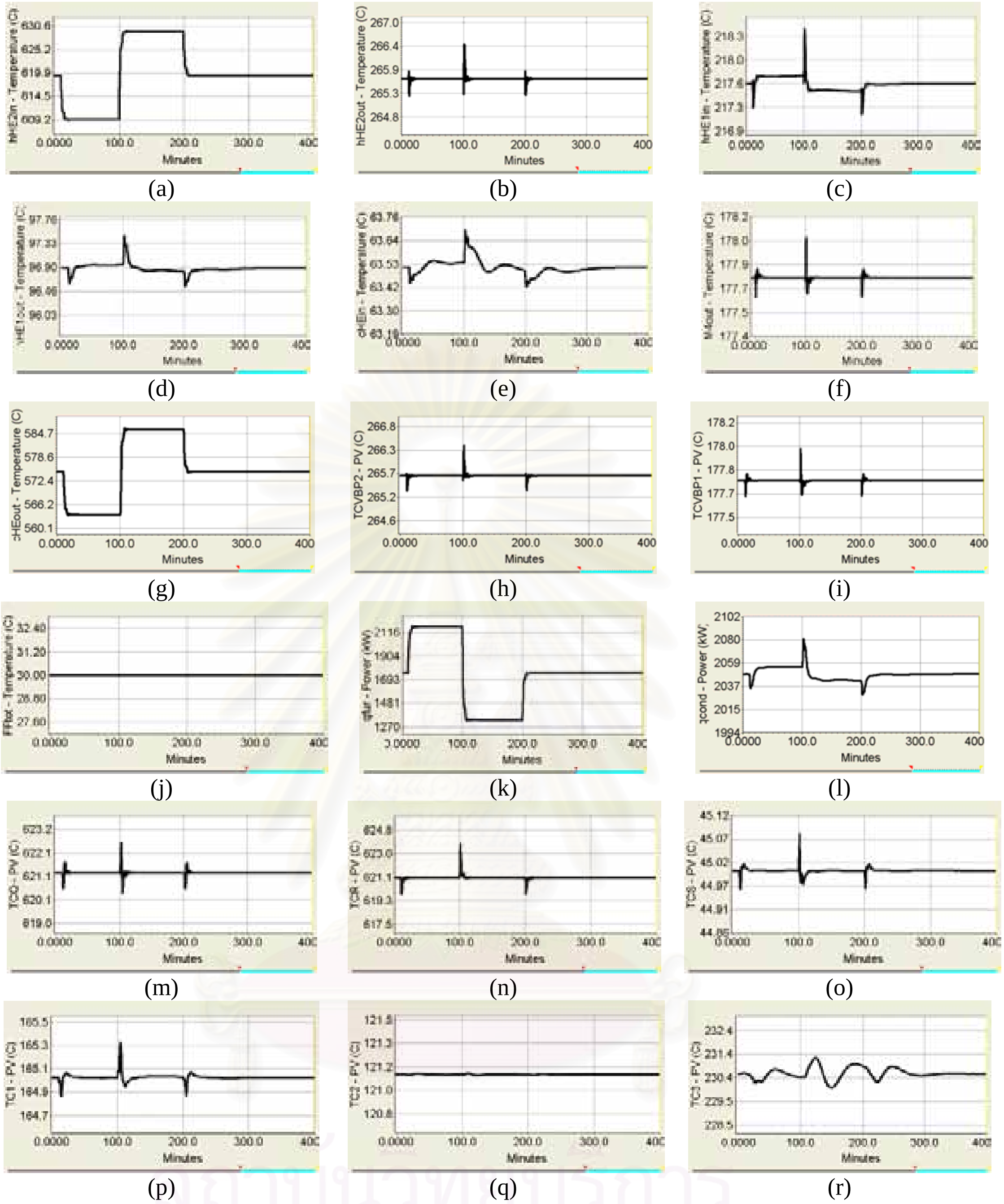
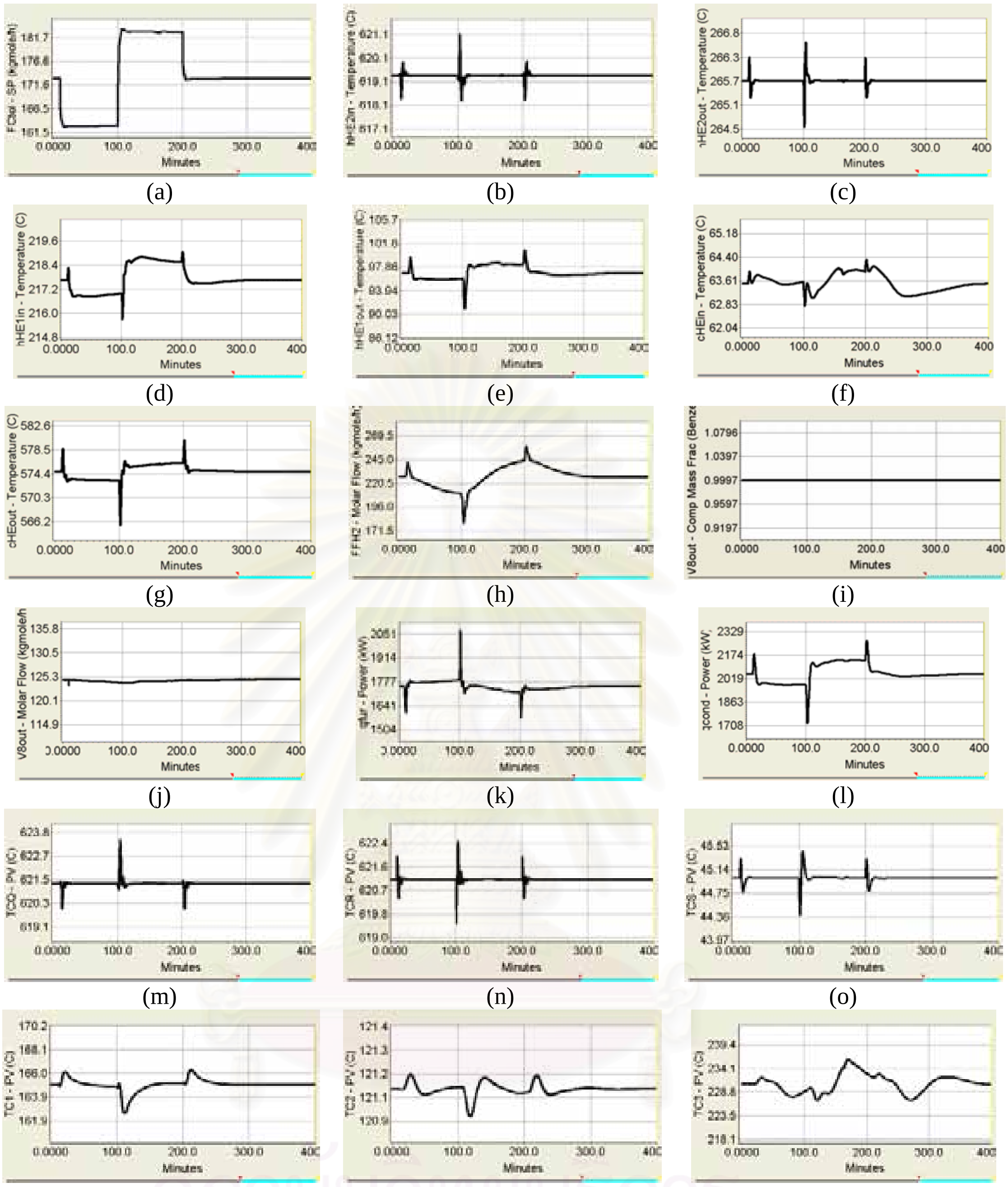


Figure 6.65 Dynamic Responses of the HDA Process Alternative 3 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.



(a) (b) (c) (d) (e) (f) (g) (h) (i) (j) (k) (l) (m) (n) (o) (p) (q) (r)

Figure 6.66 Dynamic Responses of the HDA Process Alternative 3 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

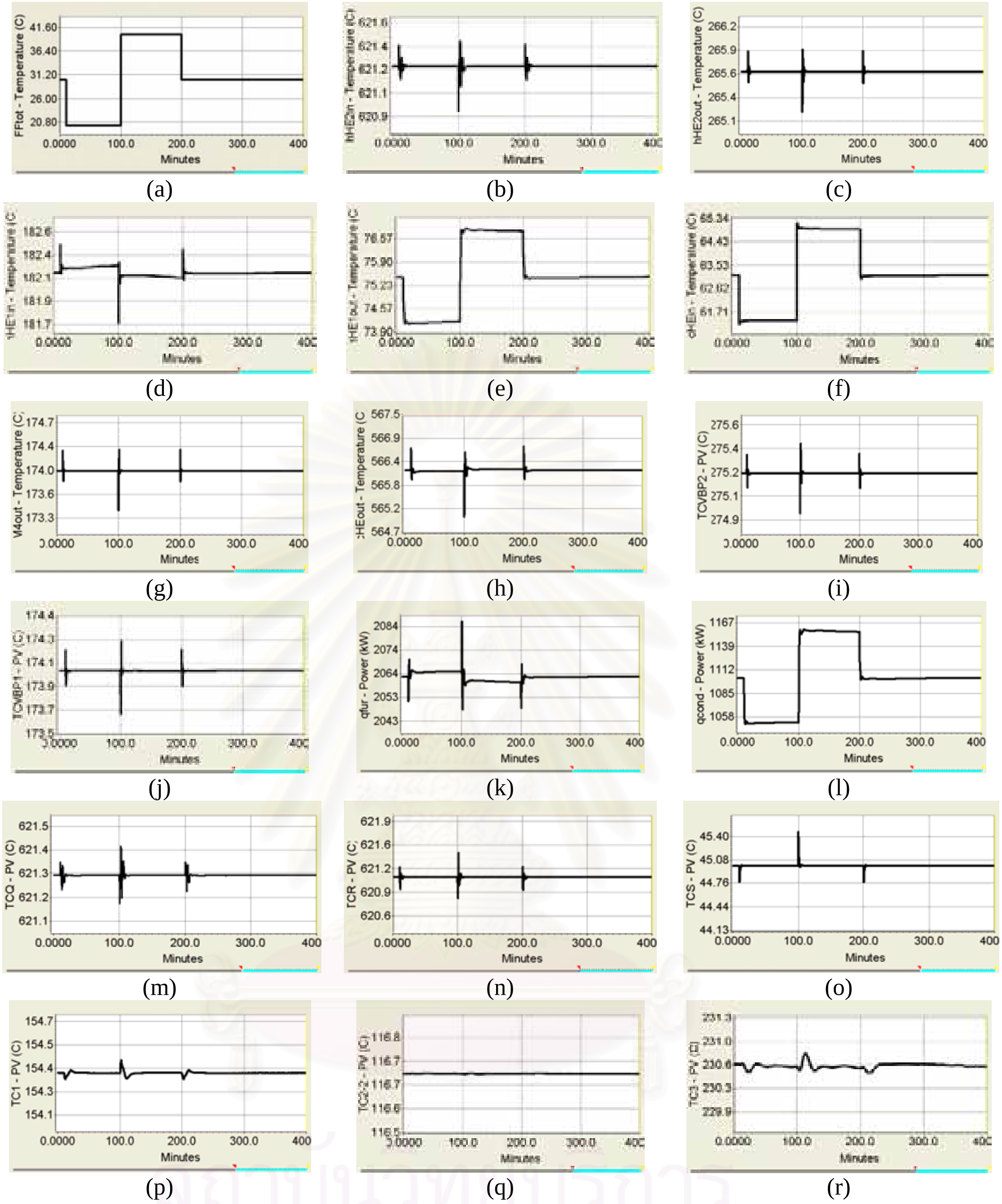


Figure 6.67 Dynamic Responses of the HDA Process Alternative 4 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

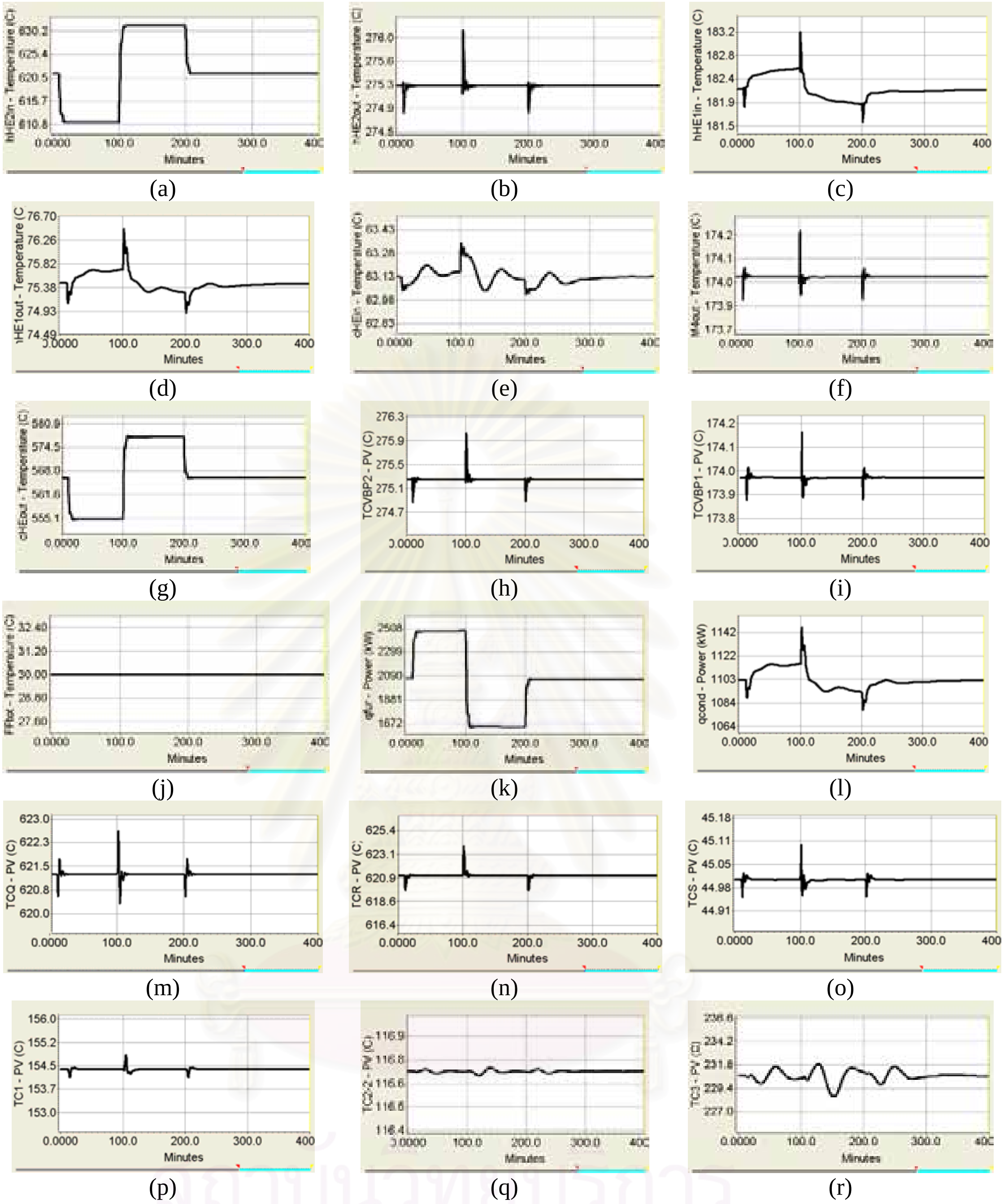


Figure 6.68 Dynamic Responses of the HDA Process Alternative 4 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

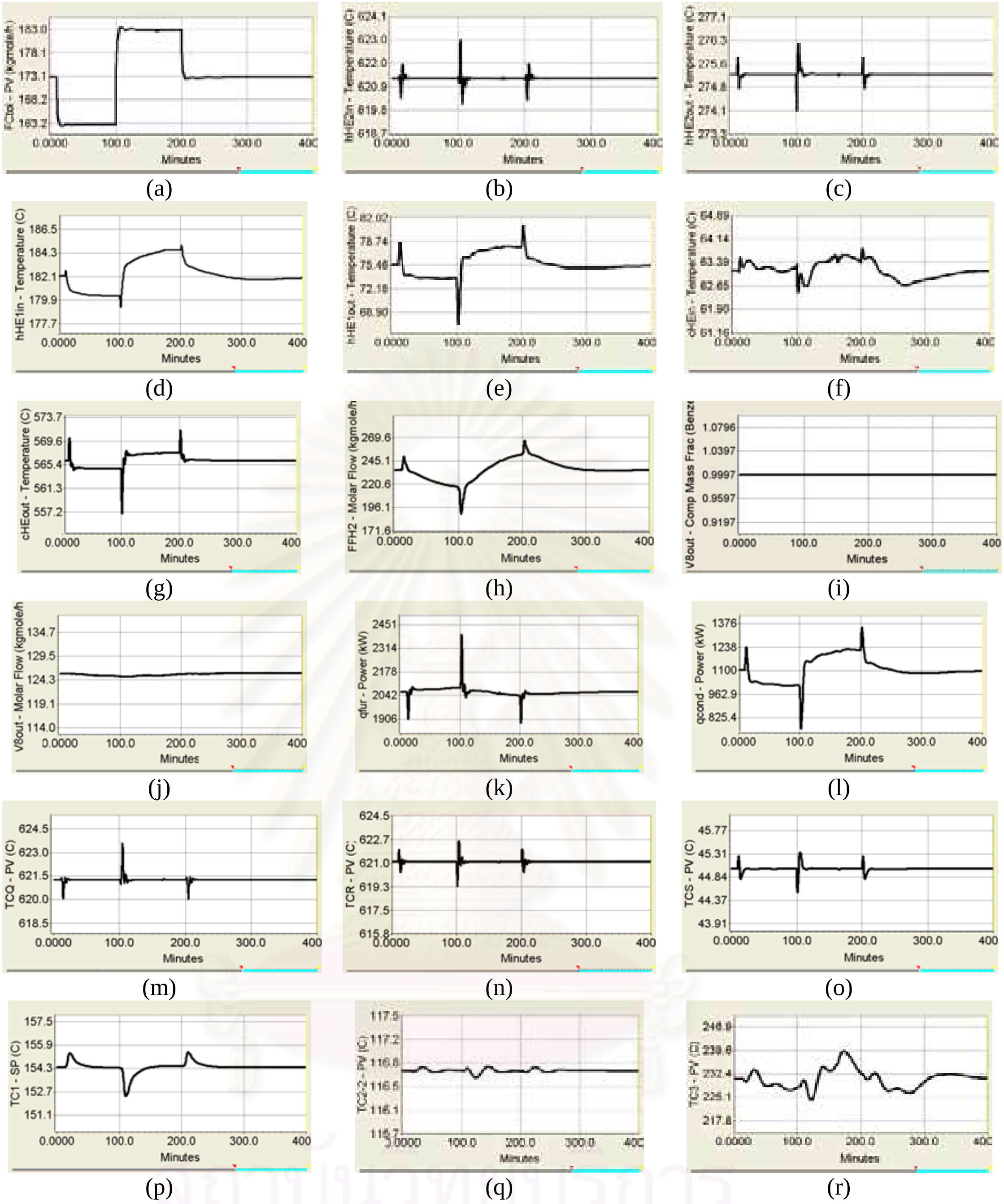
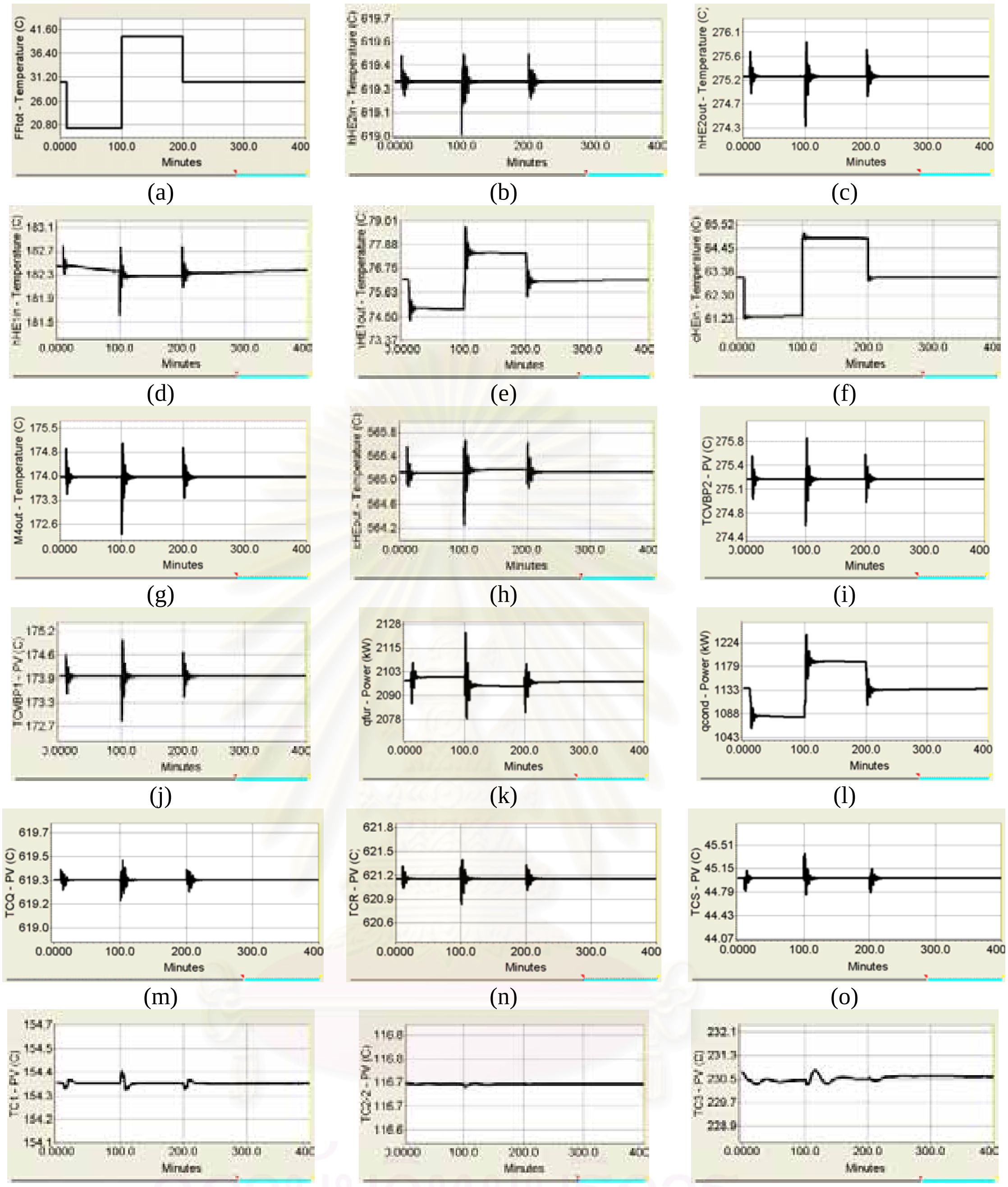


Figure 6.69 Dynamic Responses of the HDA Process Alternative 4 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.



(a) (b) (c)
(d) (e) (f)
(g) (h) (i)
(j) (k) (l)
(m) (n) (o)
(p) (q) (r)

Figure 6.70 Dynamic Responses of the HDA Process Alternative 4 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

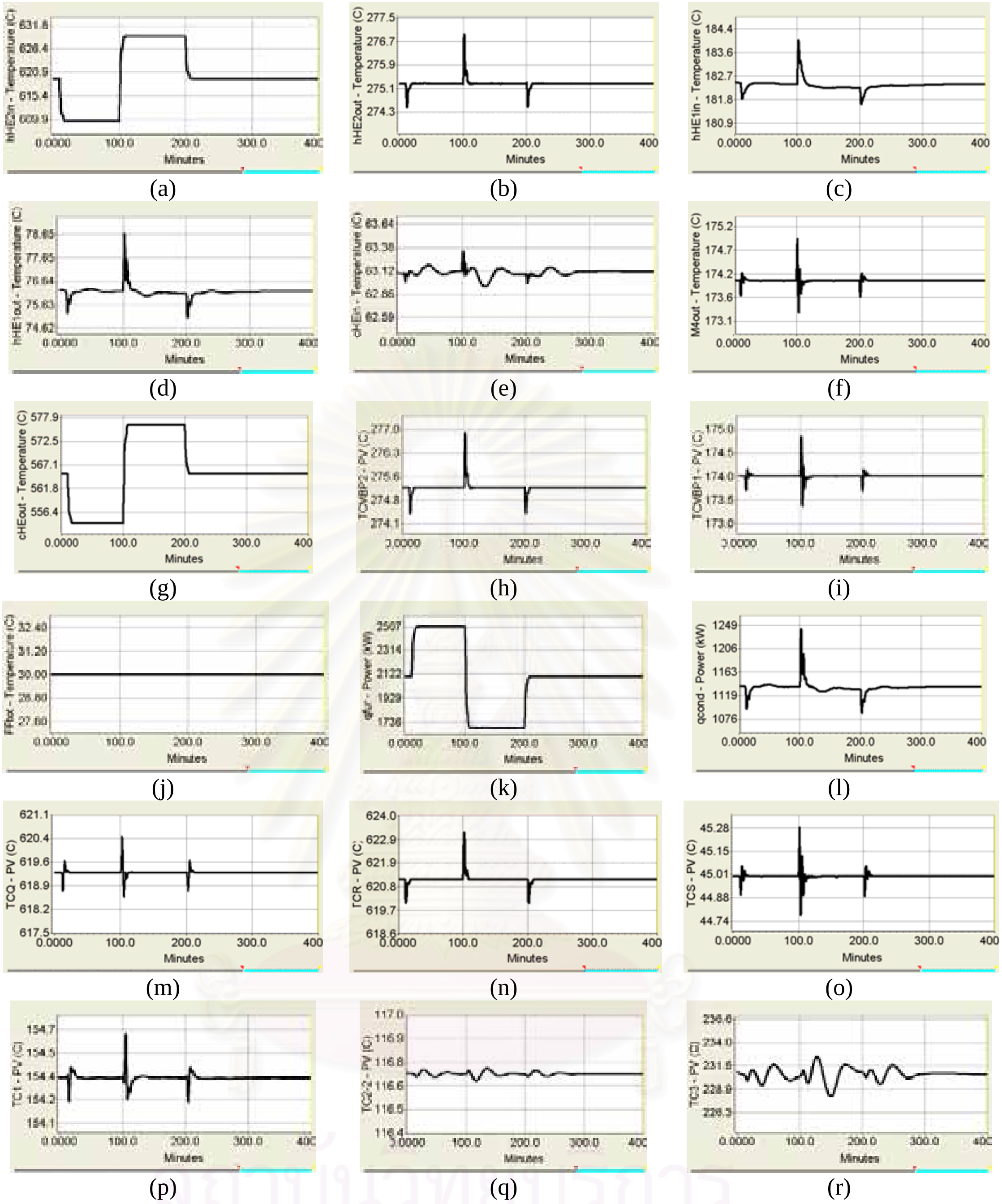


Figure 6.71 Dynamic Responses of the HDA Process Alternative 4 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

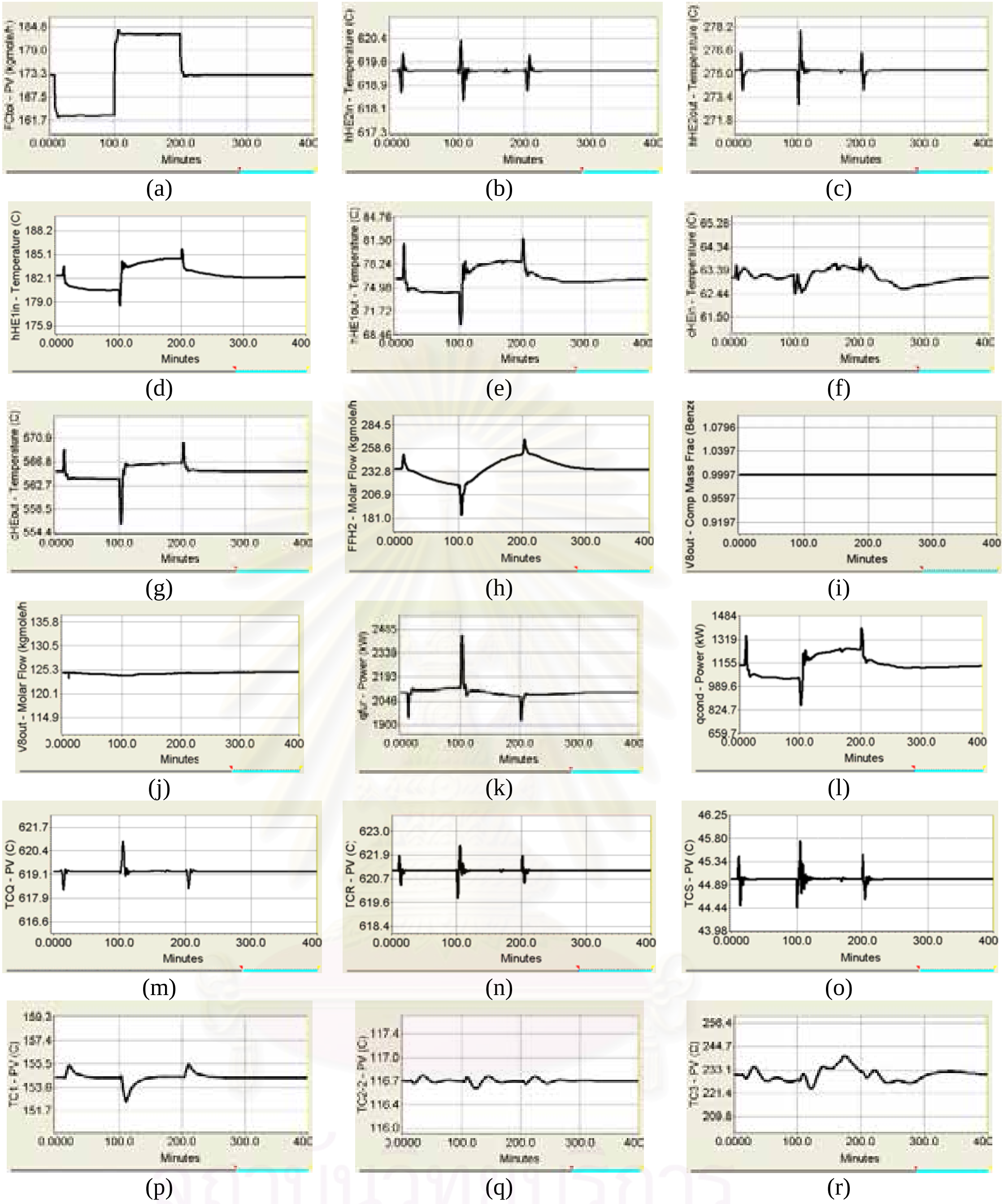


Figure 6.72 Dynamic Responses of the HDA Process Alternative 4 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

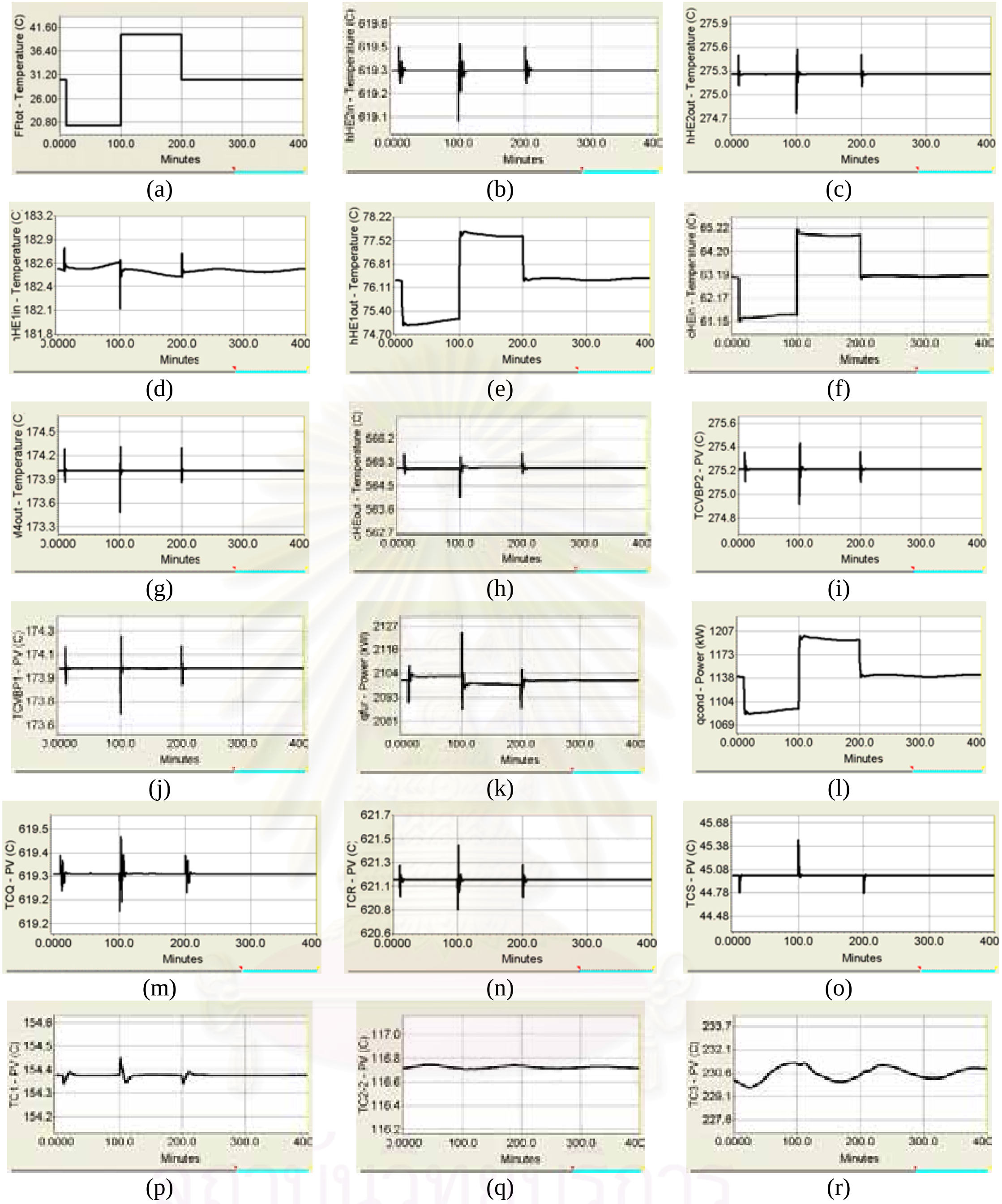


Figure 6.73 Dynamic Responses of the HDA Process Alternative 4 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

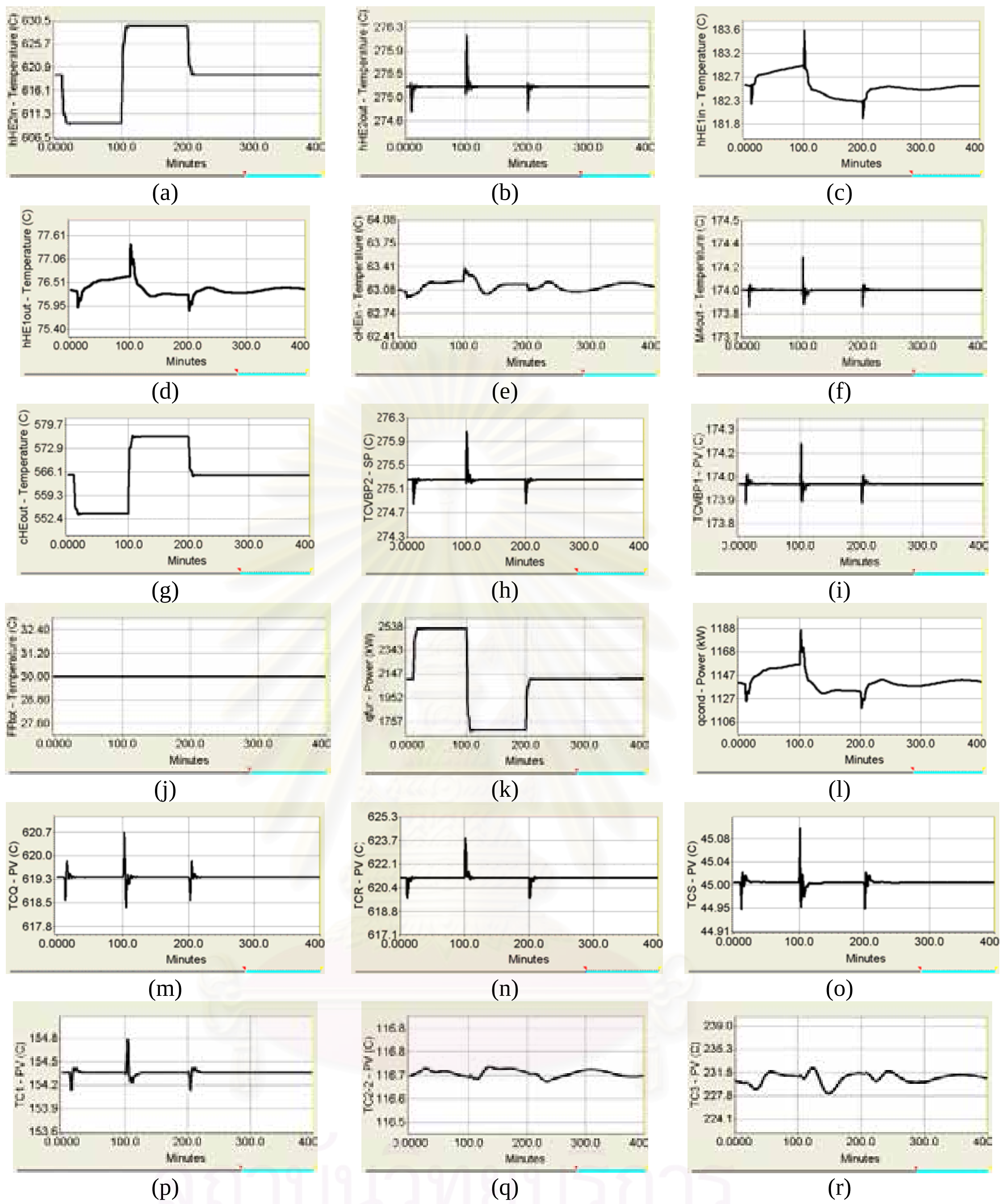


Figure 6.74 Dynamic Responses of the HDA Process Alternative 4 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

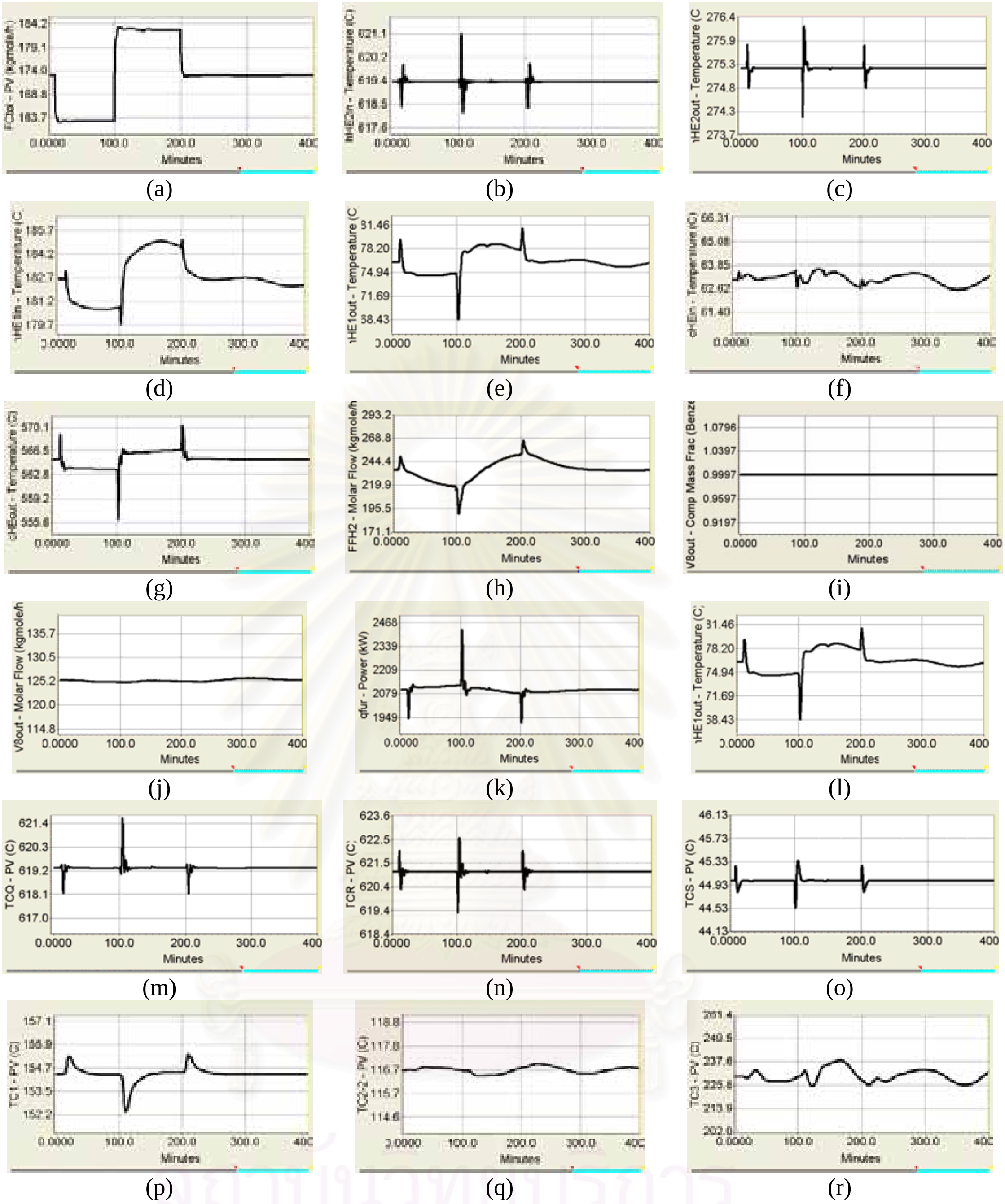


Figure 6.75 Dynamic Responses of the HDA Process Alternative 4 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

6.6 Dynamic Simulation Results for Heat-Integrated Plant of HDA Process

In order to illustrate the dynamic behavior of the control structure in the heat integrated plant of HDA process HIPs (HIP1, HIP2, HIP3, and HIP4) several disturbance loads are made. The dynamic results are explained in this part.

6.6.1 Dynamic Simulation Results for heat-integrated plant of HDA Process HIP1

Three disturbance loads are used to evaluate the dynamic performance of the new control structure (CS1, CS2 and CS3) for HDA process HIP1.

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

Figure 6.76, 6.79 and 6.82 show the dynamic responses of the control systems of HDA process HIP1 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 (Figure 6.76.a, 6.79.a and 6.82.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 heater or cooler as follows. In the first the cold inlet temperature of FEHE is decreased and then both the cold and hot outlet temperatures of FEHE decrease suddenly. The hot outlet temperature decreasing is a desired condition, hence the control action to control the cold outlet temperature of FEHE As a result, the cold and hot outlet temperature of FEHE rapidly drops to a new steady state value (Figure 6.76.c, 6.79.c and 6.82.c) and (Figure 6.76.e, 6.79.e and 6.82.e) and the cooler duty decreases (Figure 6.76.l, 6.79.l and 6.82.l) When the cold inlet temperature of FEHE increases, both the cold and hot outlet temperatures of FEHE increase. In order to the increasing cold outlet temperature is a desired condition, the control action to control the hot outlet temperature of FEHE. As a result, the cold outlet temperature of FEHE temperature quickly increases a new steady state value and the furnace duty decreases (Figure 6.76.k, 6.79.k and 6.82.k).

The hot outlet temperature of FEHE is slightly well controlled to prevent the thermal disturbance load propagation to furnace utility. This disturbance load is shifted to the cooler utility. The cooler duty will be increased in this case

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

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

Figure 6.77, 6.80 and 6.83 shows the dynamic responses of the control systems of HDA process HIP1 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 FEHE-hot-inlet temperature controller (i.e. TCX1) is decreased from 621.1 to 611.1^{o}C at time equals 10 minutes and the temperature is increased from 611.1 to 631.1^{o}C at time equals 100 minutes, then its temperature is returned to its nominal value of 621.1^{o}C at time equals 200 minutes (Figure 6.77.a, 6.80.a and 6.83.a). As can be seen, this temperature response is very fast, the new steady state is reached quickly

Since the hot outlet temperature of FEHE is controlled (Figure 6.77.b, 6.80.b and 6.83.b) 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, this disturbance load is shifted to the cooler utility. The cooler duty will be decreased in this case (Figure 6.77.l, 6.80.l and 6.83.l). Consider the case when the hot inlet temperature of FEHE increases, this is a desired condition to shift the disturbance load to the cold stream. Therefore, the cooler duty increases to a new steady state value.

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 CS1 is smoother than CS2 and CS3.

6.6.1.3 Change in the Total Toluene Feed Flowrates

Figure 6.78, 6.81 and 6.84 shows the dynamic responses of the control systems of HDA process HIP1 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 minutes (Figure 6.78.a, 6.81.a and 6.84.a).

The dynamic result can be seen that the drop in total toluene feed flowrates reduces the reaction rate, so fresh feed hydrogen flowrates (Figure 6.78.h, 6.81.h and 6.84.h). and the benzene product flowrates drops (Figure 6.78.j, 6.81.j and 6.84.j).and the benzene product quality increases (Figure 6.78.i, 6.81.i and 6.84.i). Consider the case when the total toluene feed flowrates increase and the benzene product flowrates increase because of the reaction rate enlargement. The benzene product quality will increase in this case. The deviation of benzene product quality from nominal value in CS1, CS2 and CS3 are slightly similar when total toluene feed flowrates change.

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

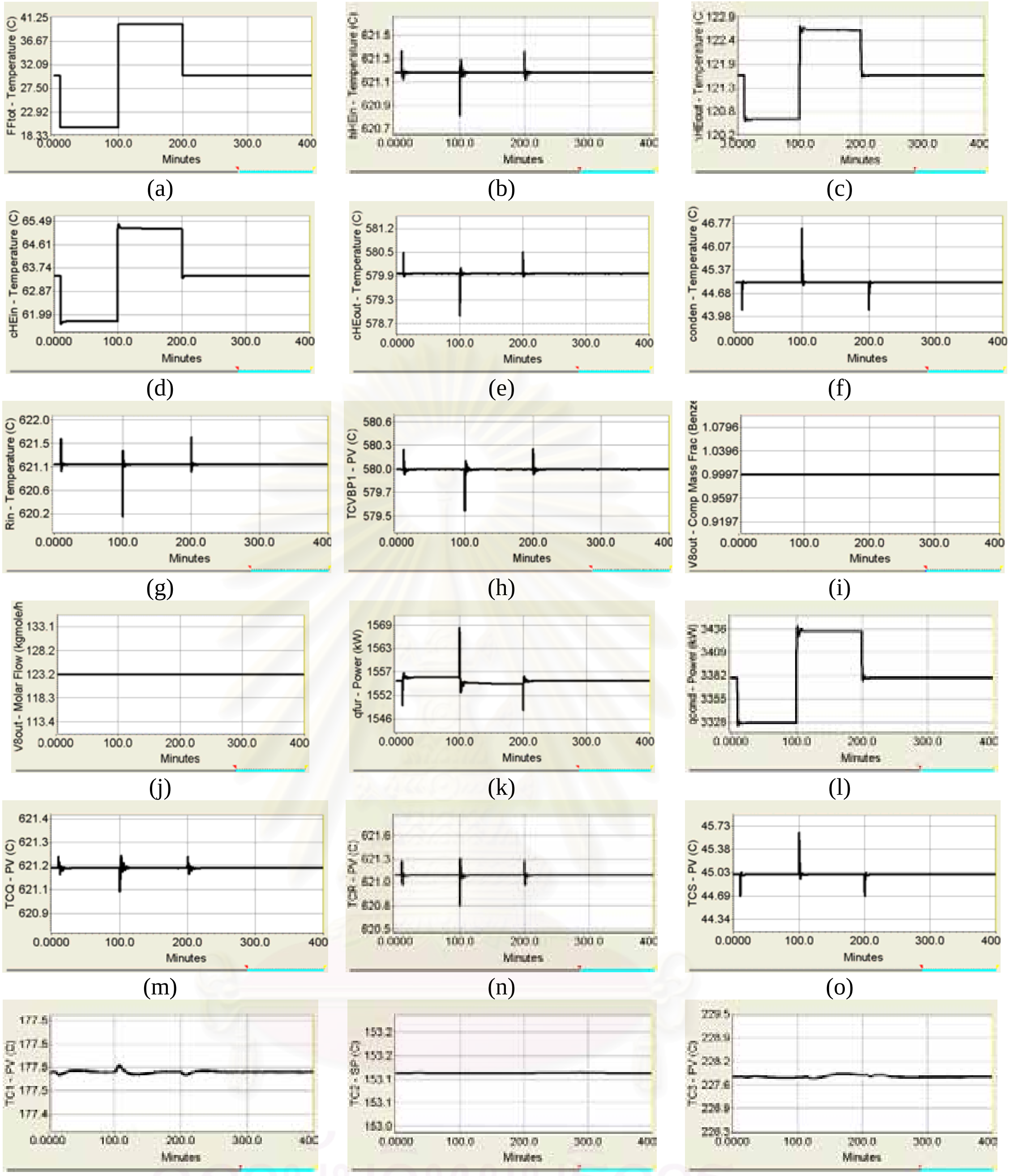


Figure 6.76 Dynamic Responses of the HDA Process of HIP 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

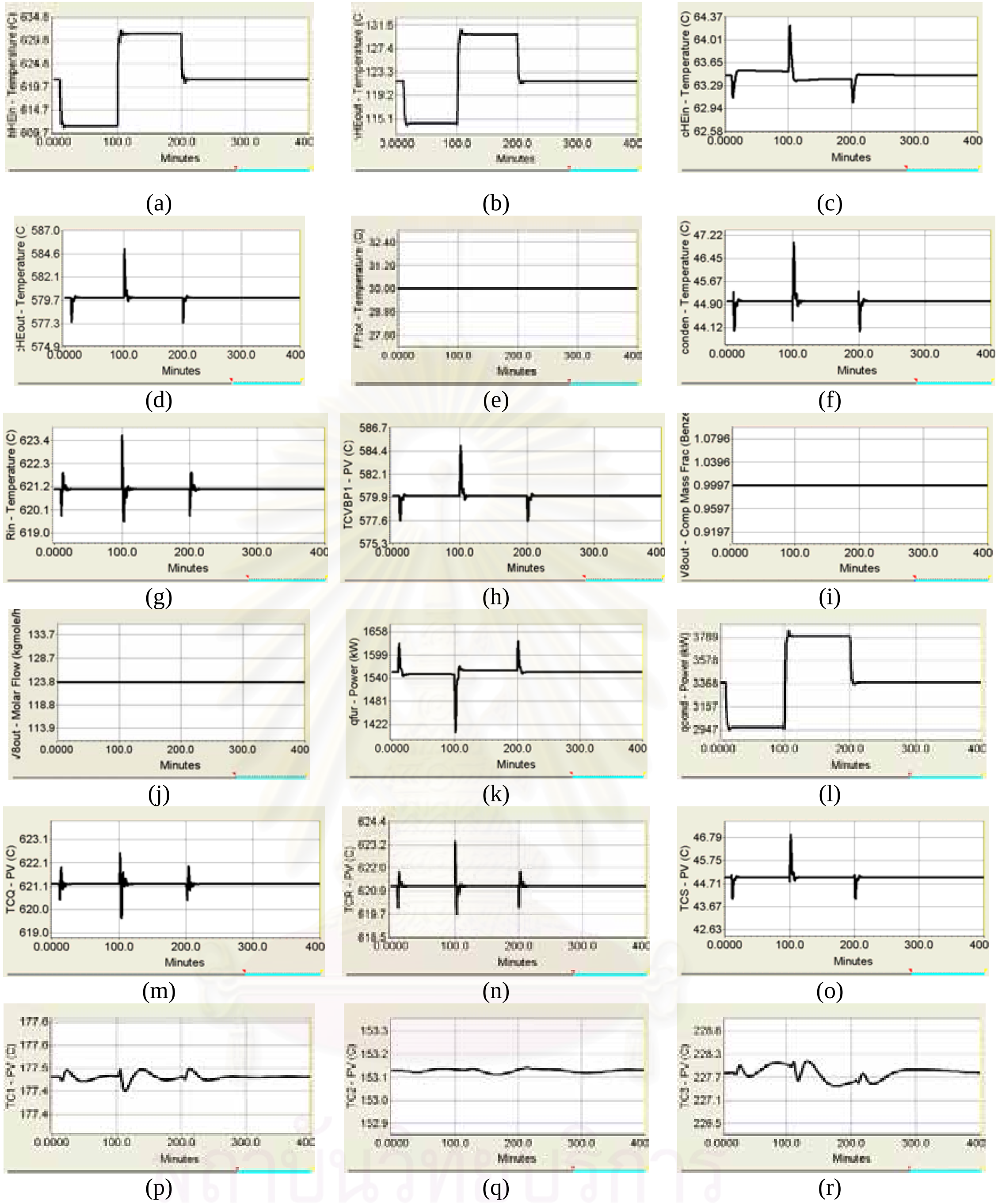


Figure 6.77 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

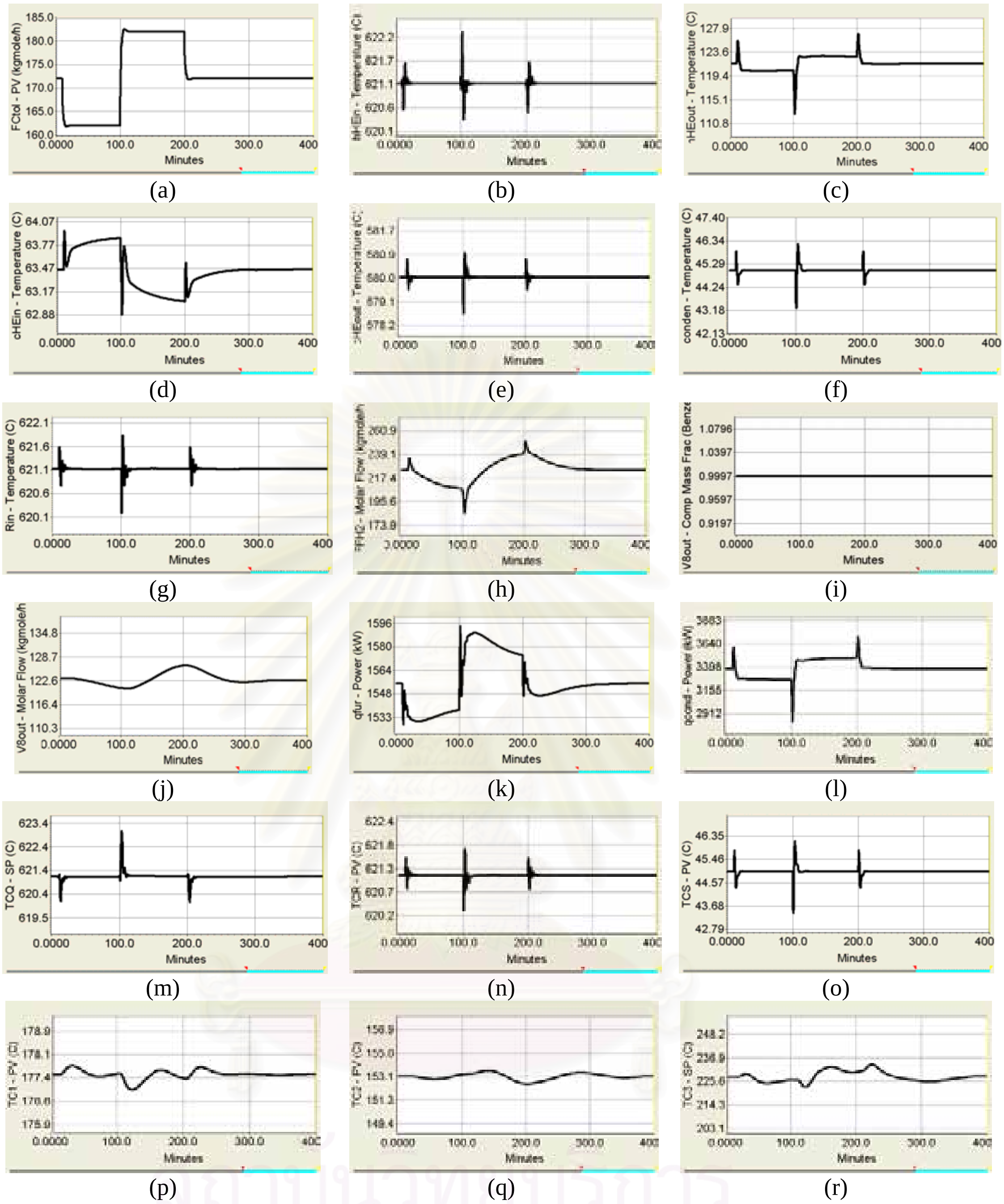


Figure 6.78 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

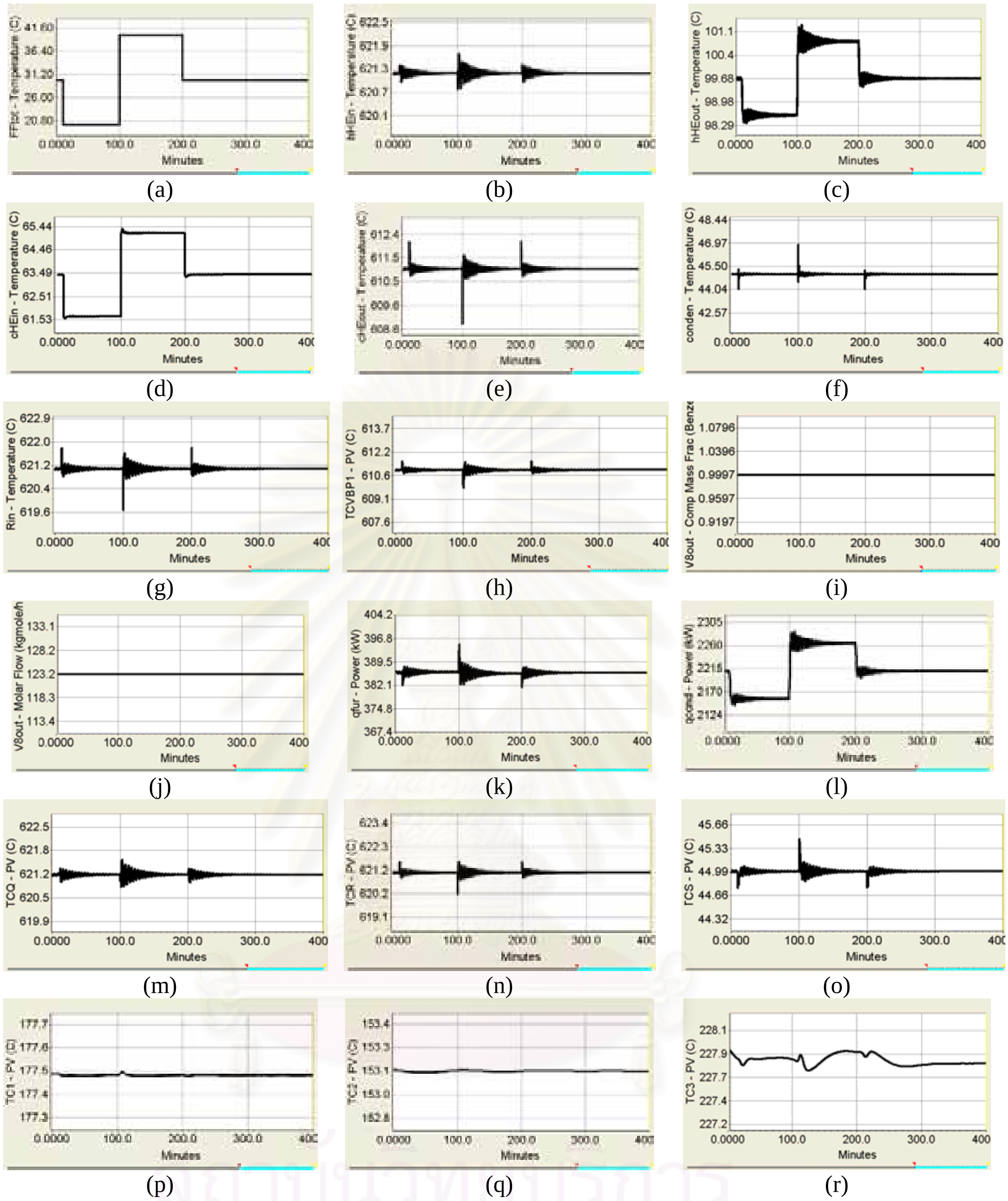


Figure 6.79 Dynamic Responses of the HDA Process of HIP 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

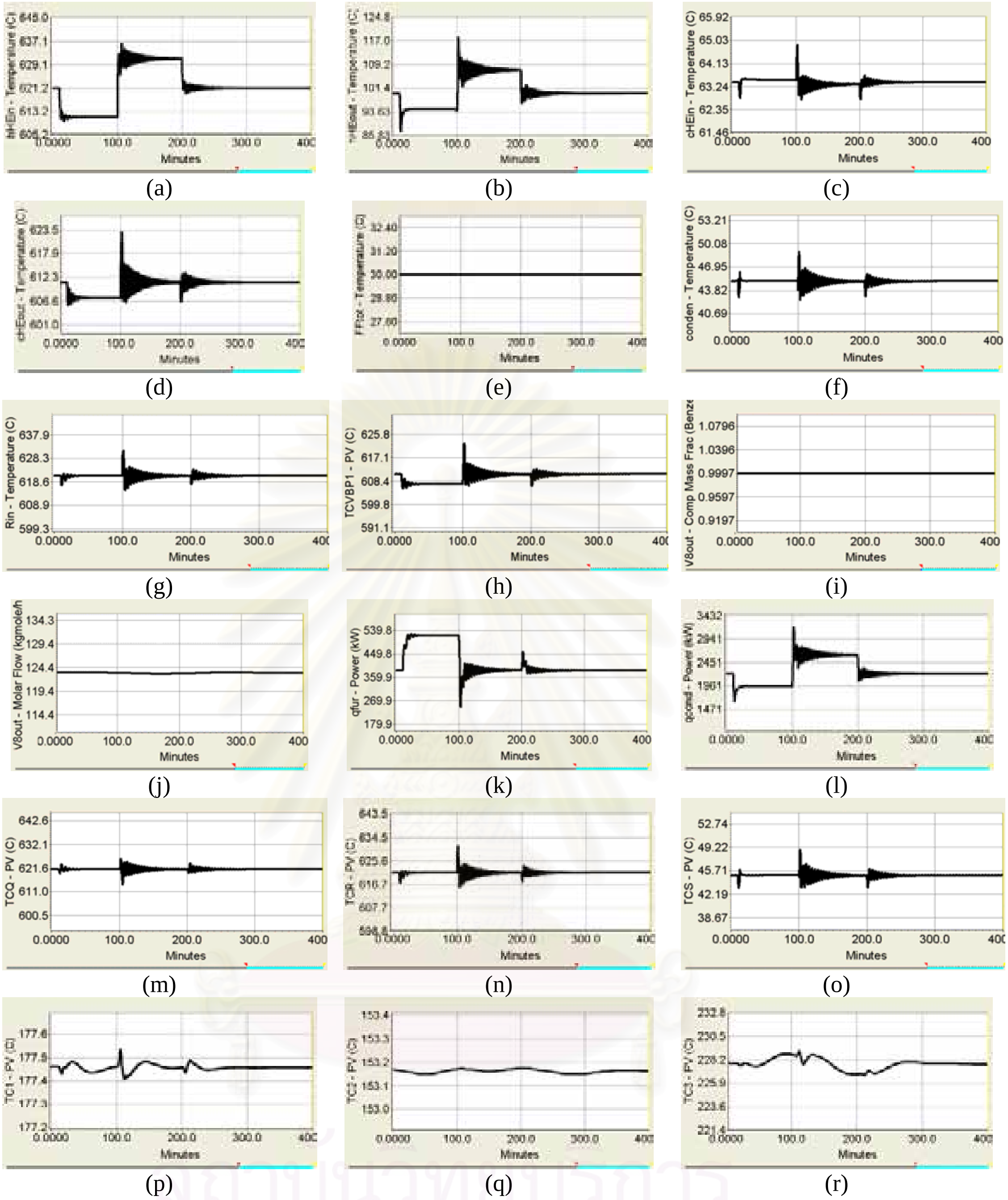


Figure 6.80 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

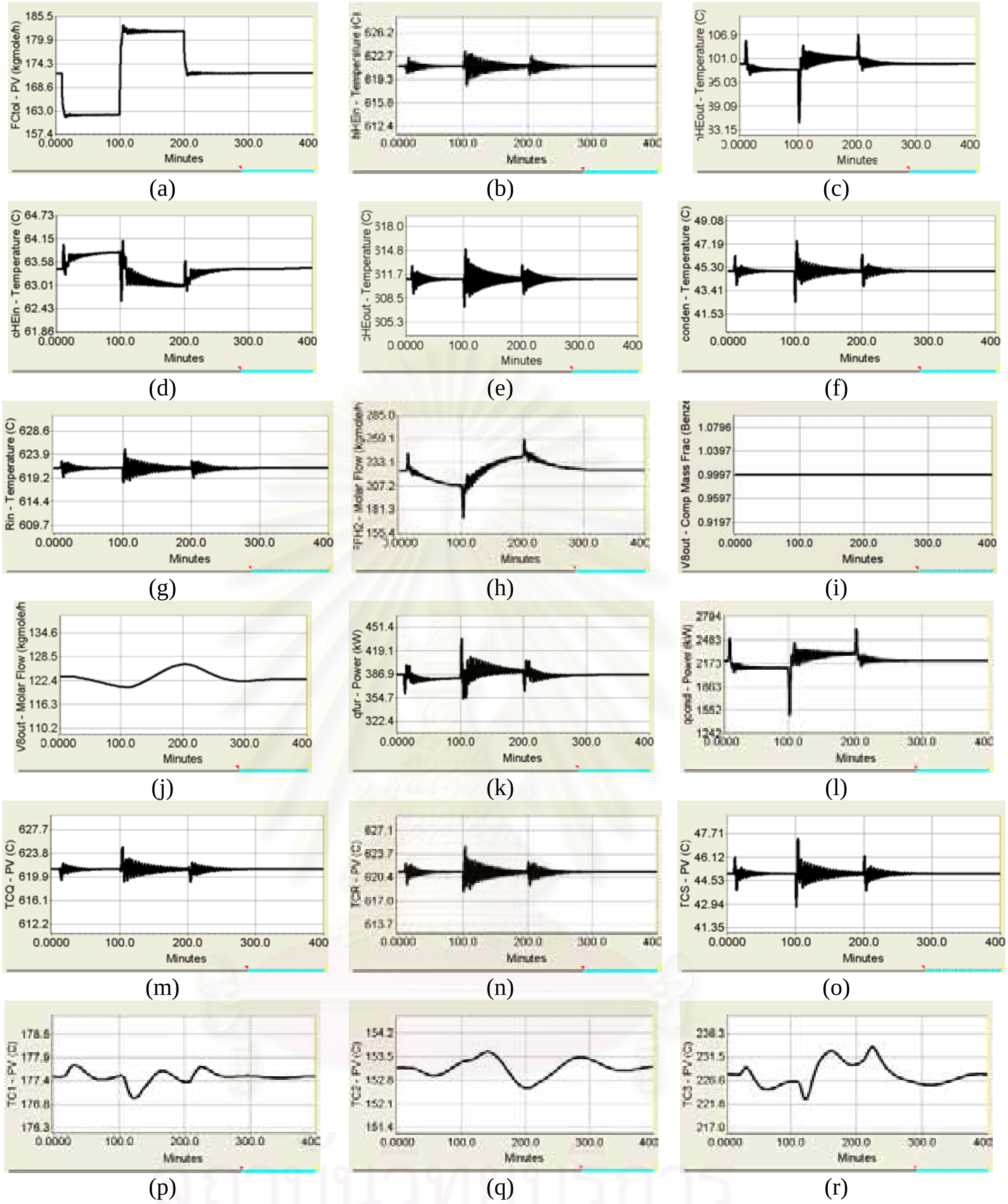


Figure 6.81 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

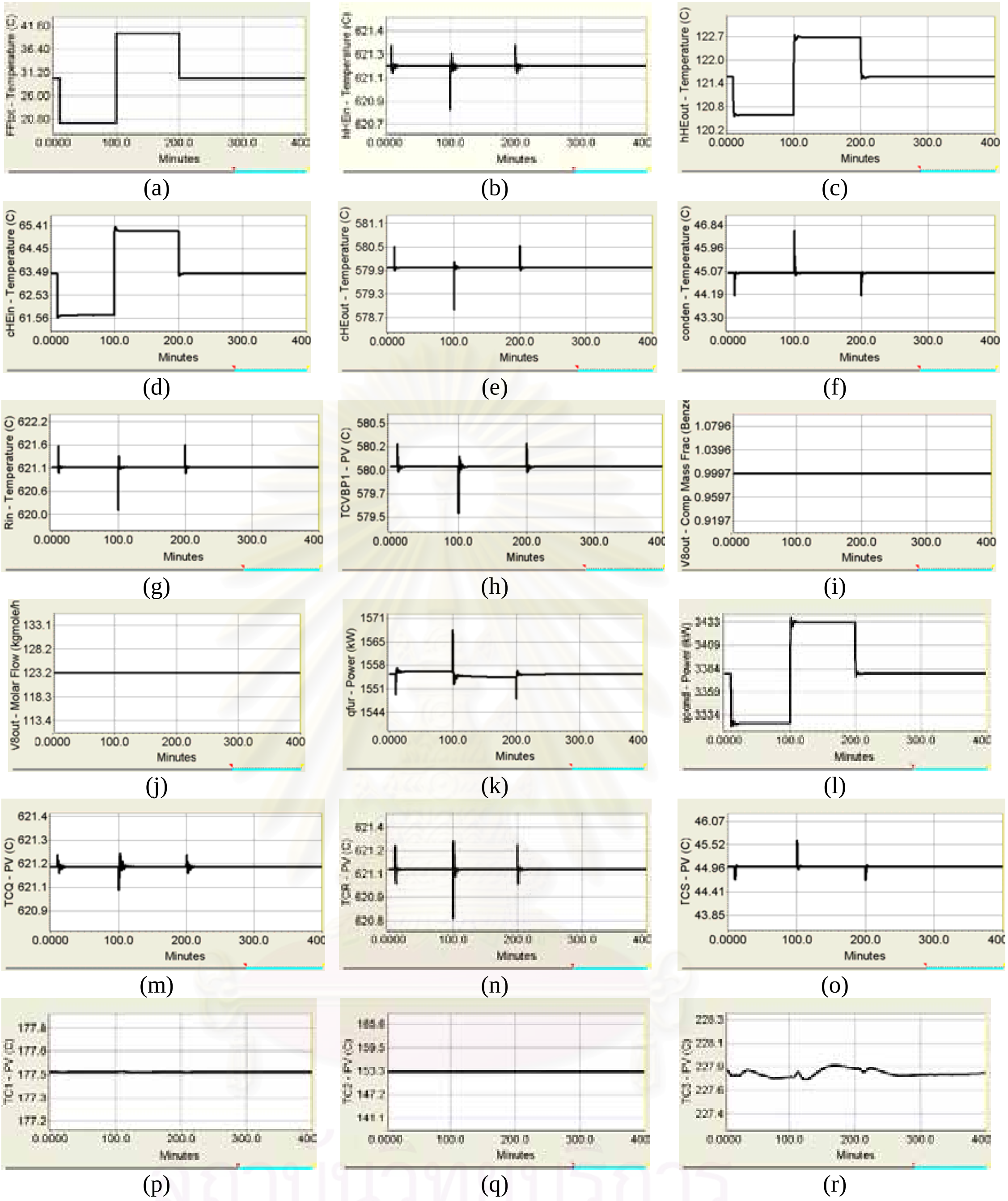


Figure 6.82 Dynamic Responses of the HDA Process of HIP 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

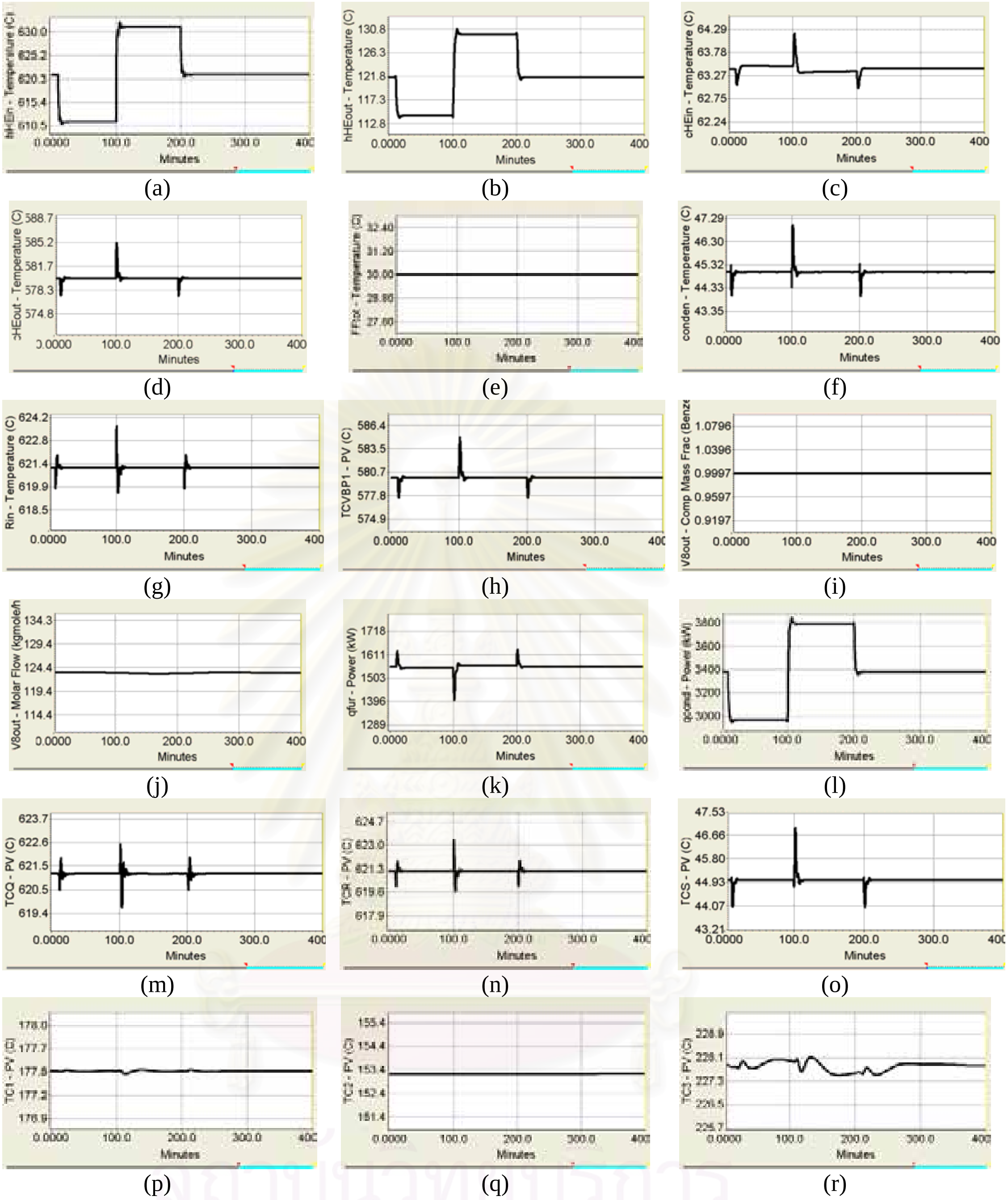


Figure 6.83 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

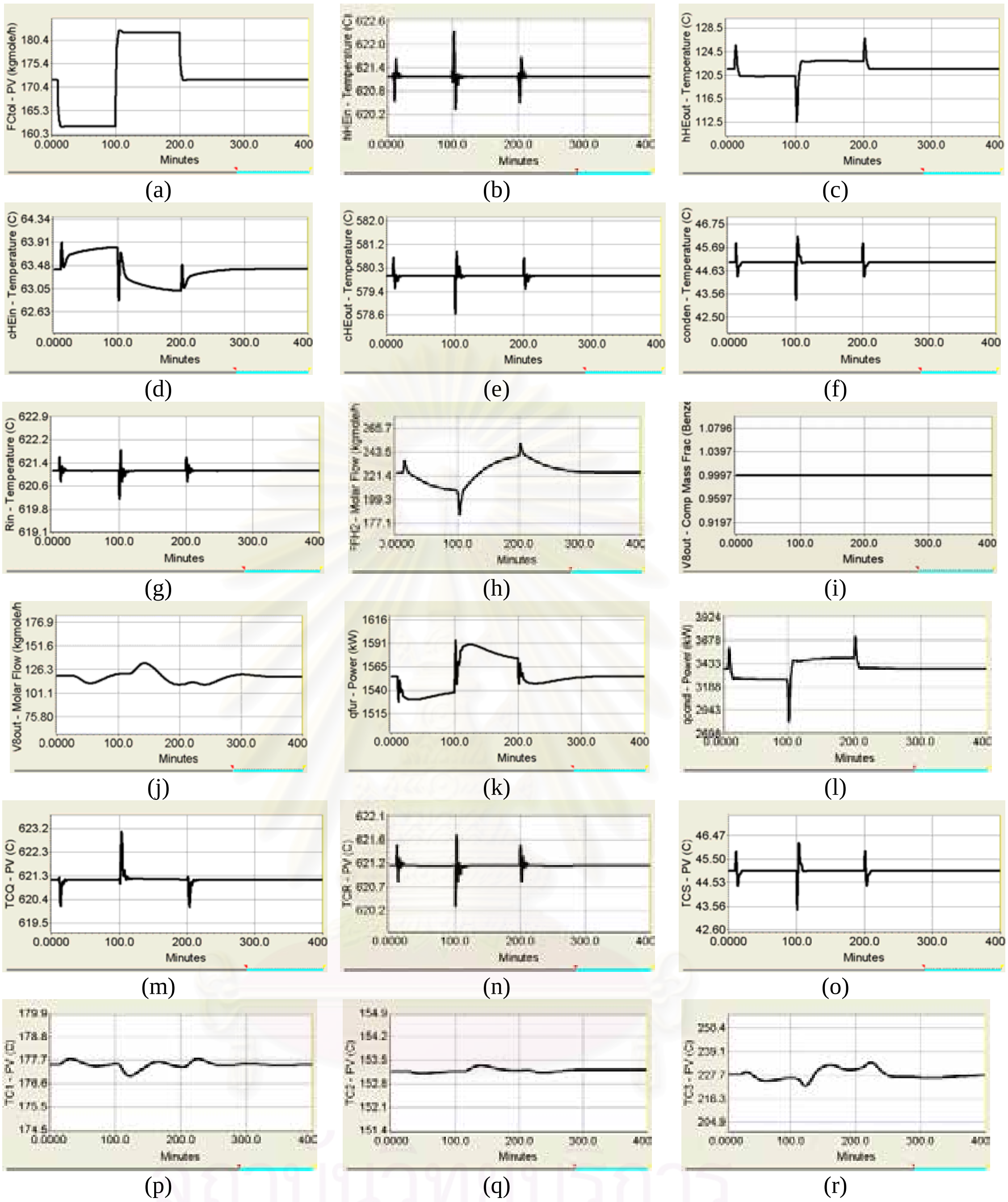


Figure 6.84 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

6.6.2 Dynamic Simulation Results for Heat-Integrated Plant of HDA Process HIP2

Three disturbance loads are also used to evaluate the dynamic performance of the new control structure (CS1, CS2 and CS3) for HDA process HIP2, (1) Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream) show the dynamic responses on Figure 6.85, 6.88 and 6.91, (2) Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream) show the dynamic responses on Figure 6.86, 6.89 and 6.92, (3) Change in the Total Toluene Feed Flowrates show the dynamic responses on Figure 6.87, 6.90 and 6.93, As can be see, the dynamic response for HDA process HIP2 are same in CS1, CS2 and CS3.

6.6.3 Dynamic Simulation Results for Heat-Integrated Plant of HDA Process HIP3

Three disturbance loads are also used to evaluate the dynamic performance of the new control structure (CS1, CS2 and CS3) for HDA process HIP3, (1) Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream) show the dynamic responses on Figure 6.94, 6.97 and 6.100, (2) Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream) show the dynamic responses on Figure 6.95, 6.98 and 6.101, (3) Change in the Total Toluene Feed Flowrates show the dynamic responses on Figure 6.96, 6.99 and 6.102, As can be see, the dynamic response for HDA process HIP3 are same in CS1, CS2 and CS3.

6.6.4 Dynamic Simulation Results for Heat-Integrated Plant of HDA Process HIP4

Three disturbance loads are also used to evaluate the dynamic performance of the new control structure (CS1, CS2 and CS3) for HDA process HIP4, (1) Change in the Heat Load Disturbance of cold stream (Reactor Feed Stream) show the dynamic responses on Figure 6.103, 6.106 and 6.109, (2) Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream) show the dynamic responses on Figure 6.104, 6.107 and 6.110, (3) Change in the Total Toluene Feed Flowrates show the dynamic responses on Figure 6.105, 6.108 and 6.111, As can be see, the dynamic response for HDA process HIP4 are same in CS1, CS2 and CS3.

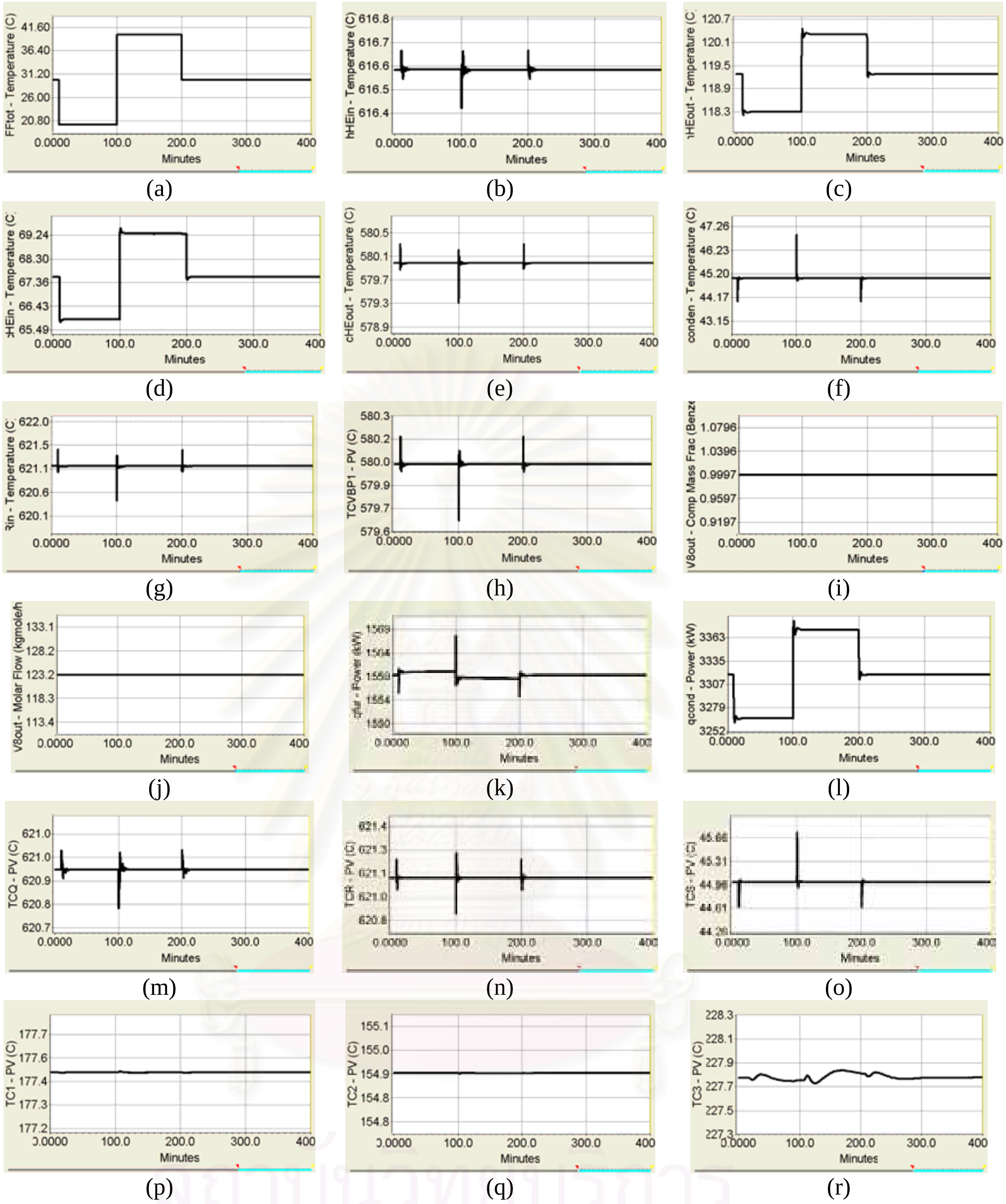


Figure 6.85 Dynamic Responses of the HDA Process of HIP 2 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

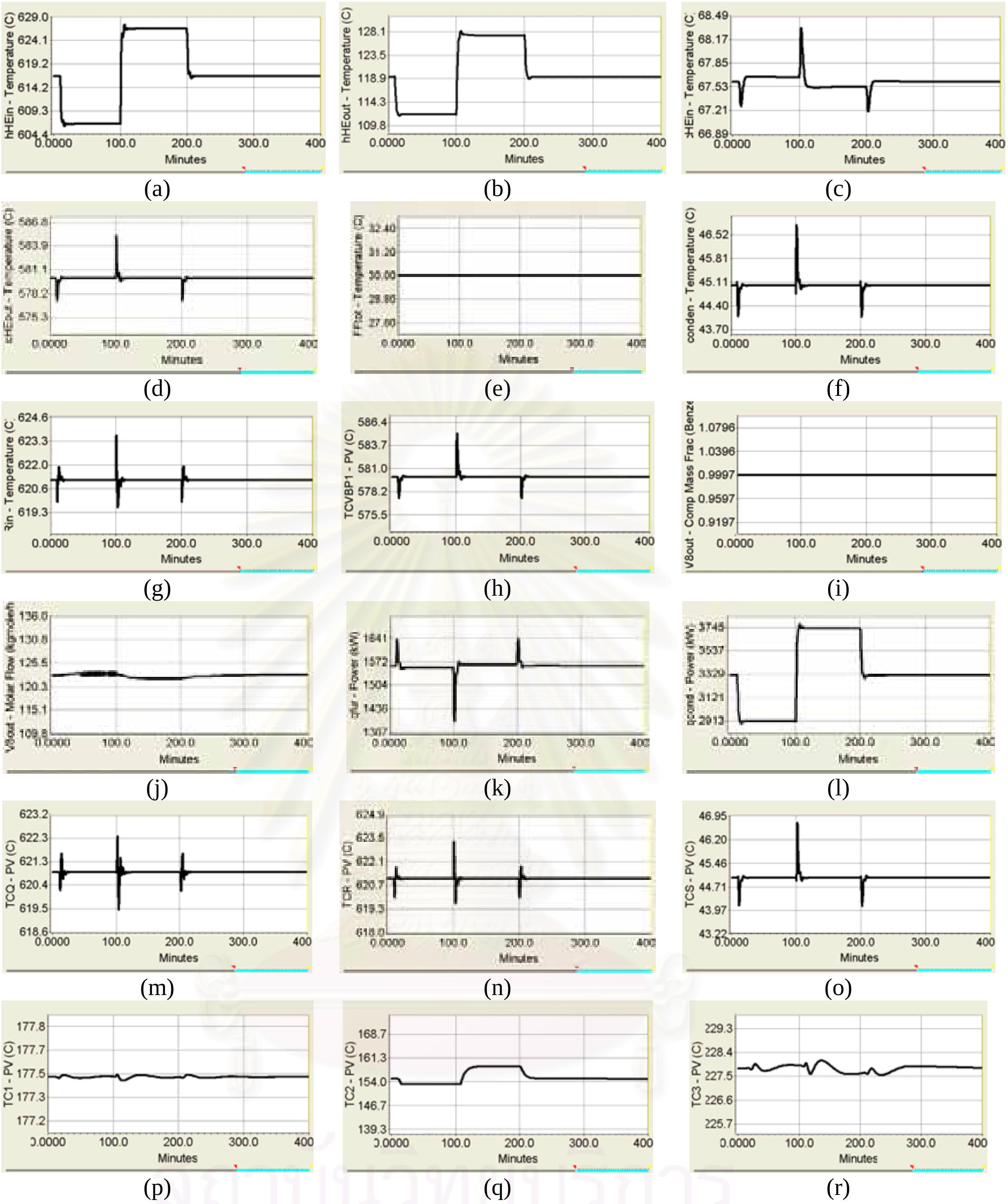


Figure 6.86 Dynamic Responses of the HDA Process of HIP 2 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

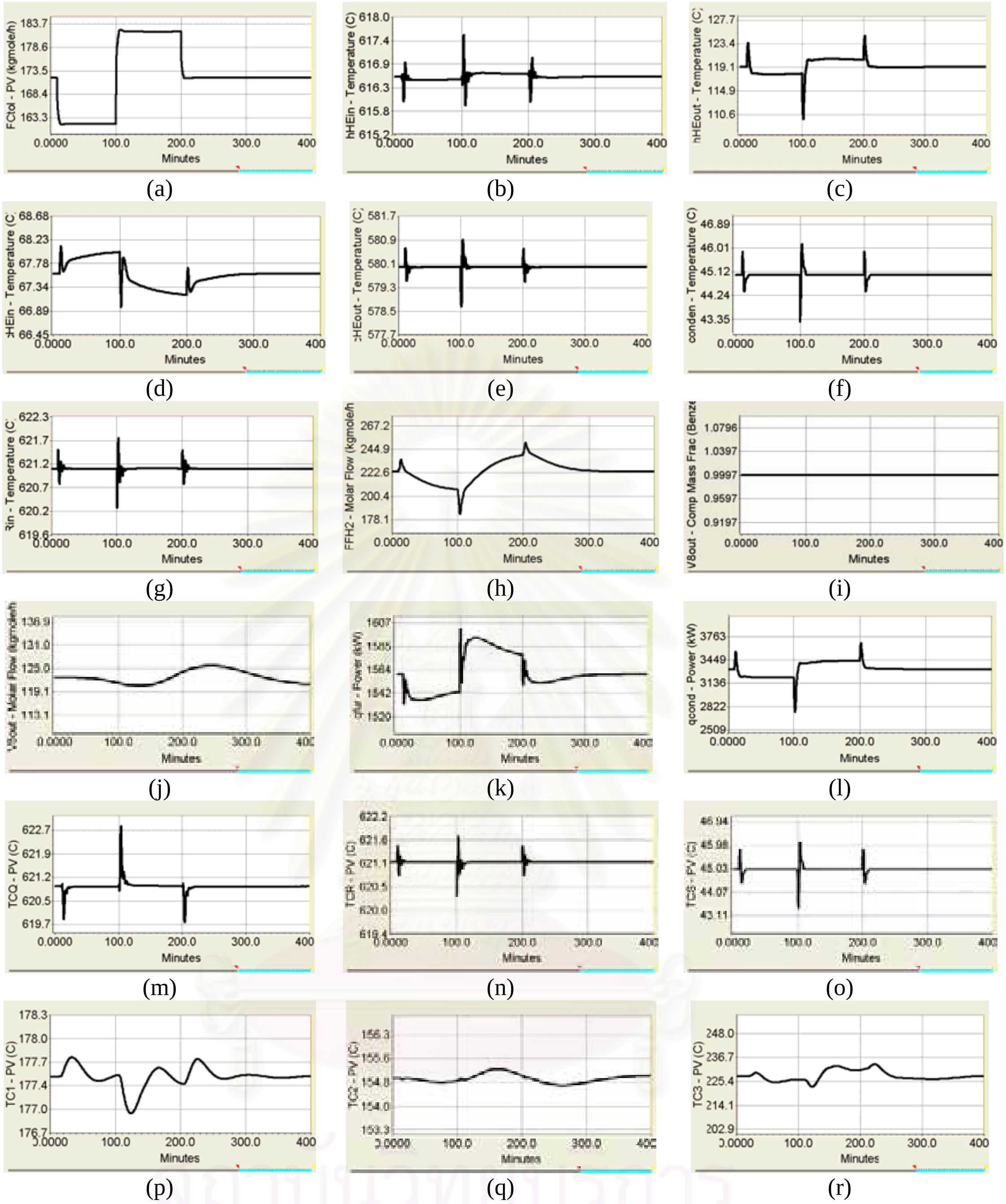


Figure 6.87 Dynamic Responses of the HDA Process of HIP 2 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

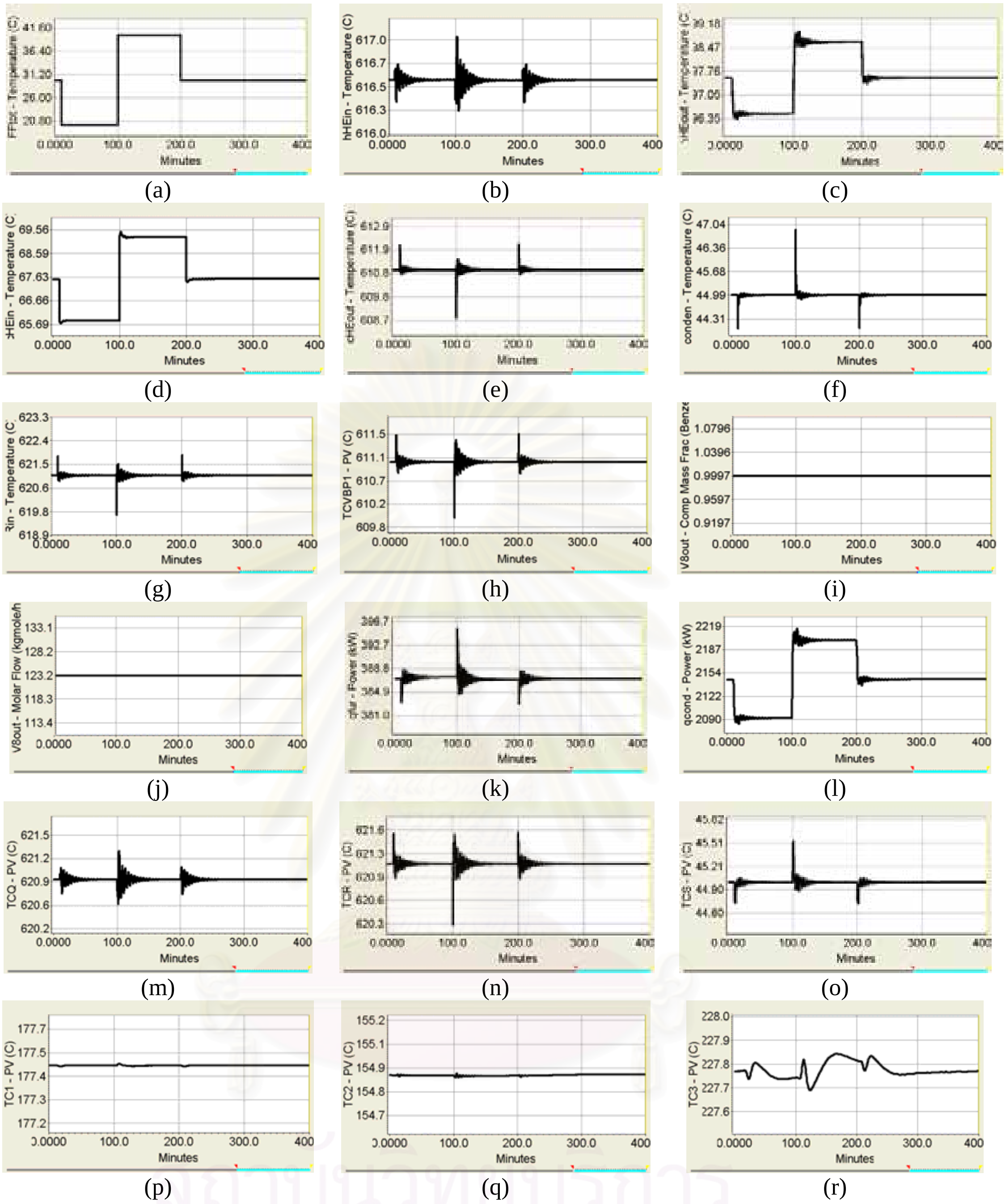


Figure 6.88 Dynamic Responses of the HDA Process of HIP 2 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

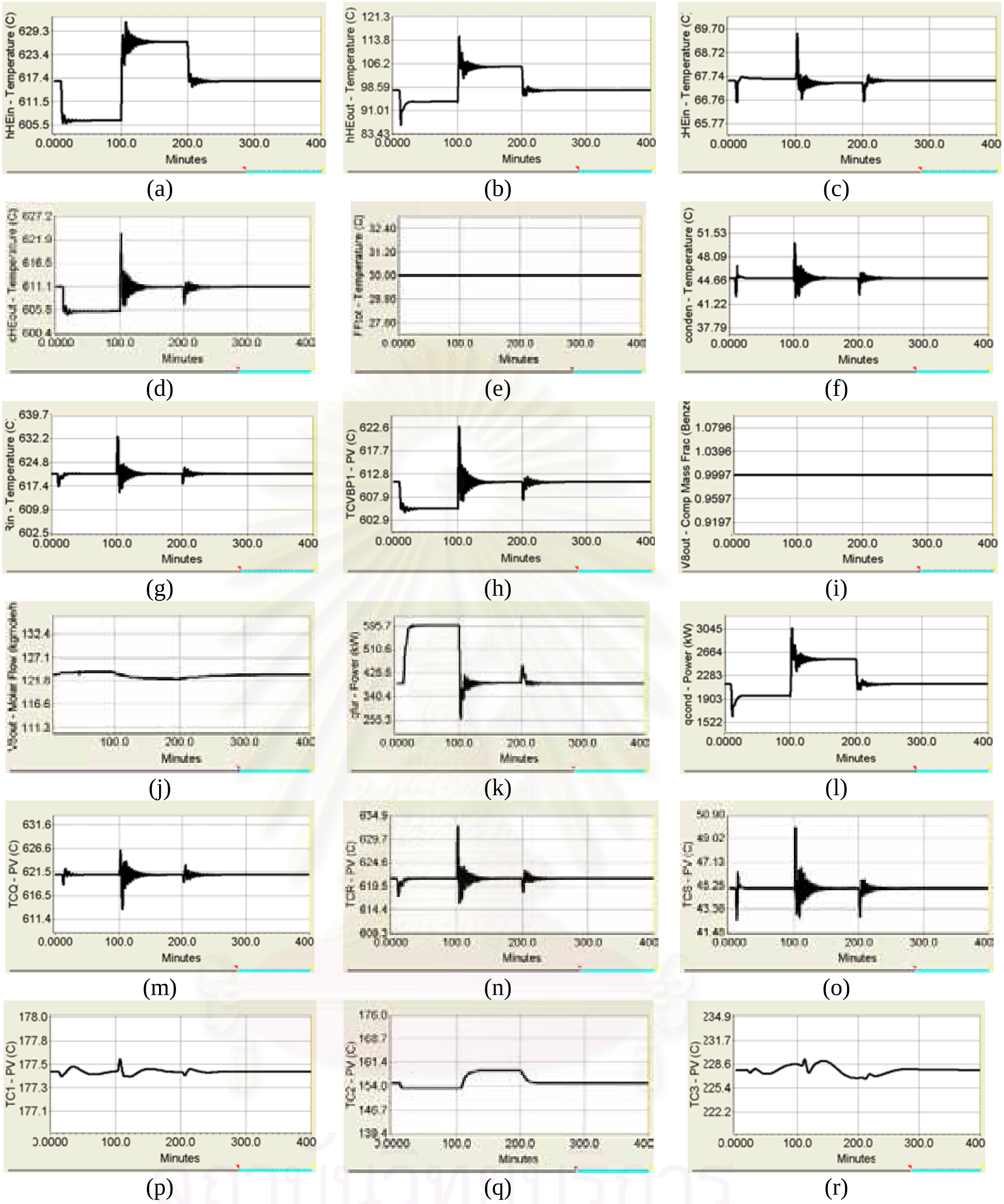


Figure 6.89 Dynamic Responses of the HDA Process of HIP 2 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

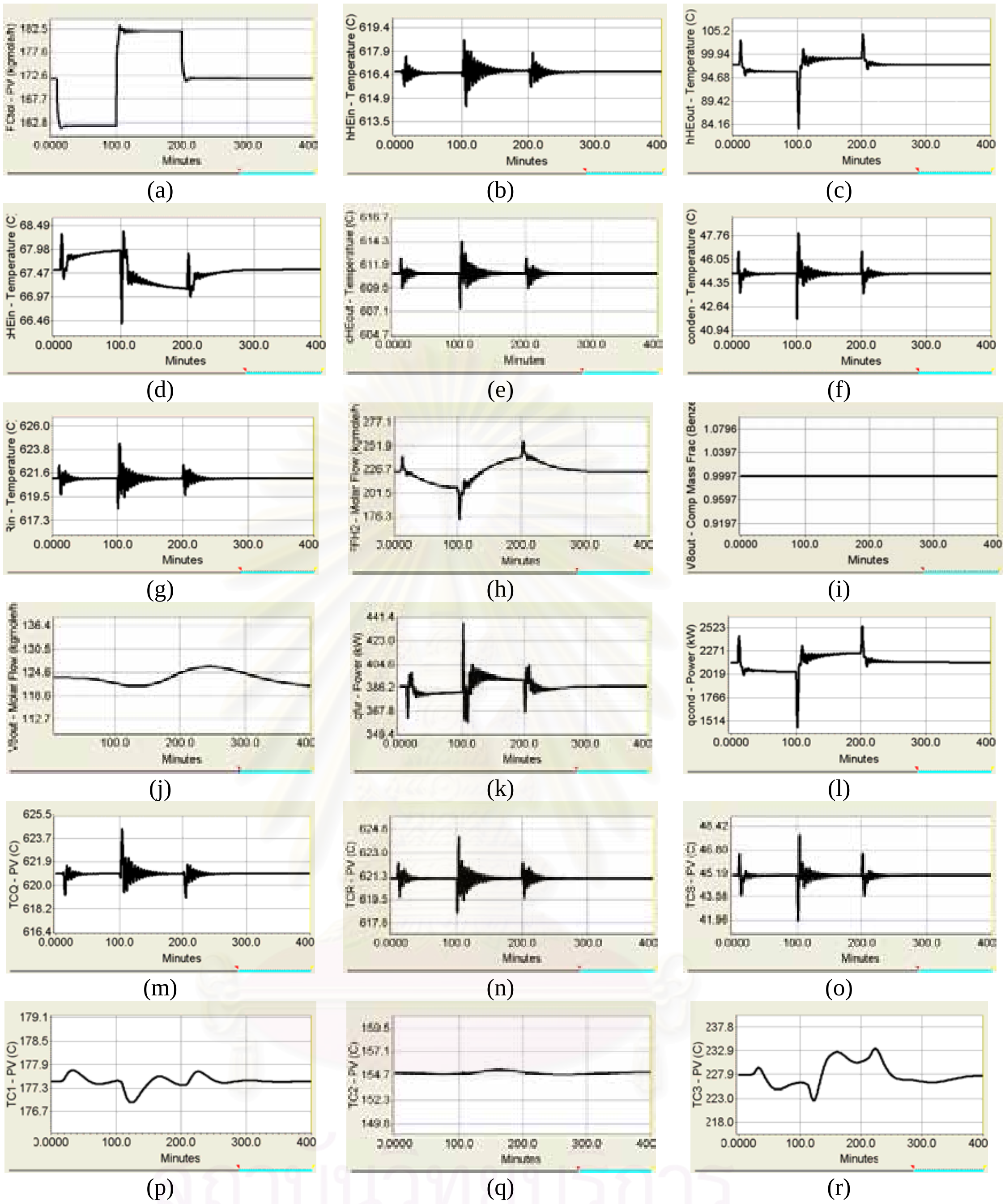
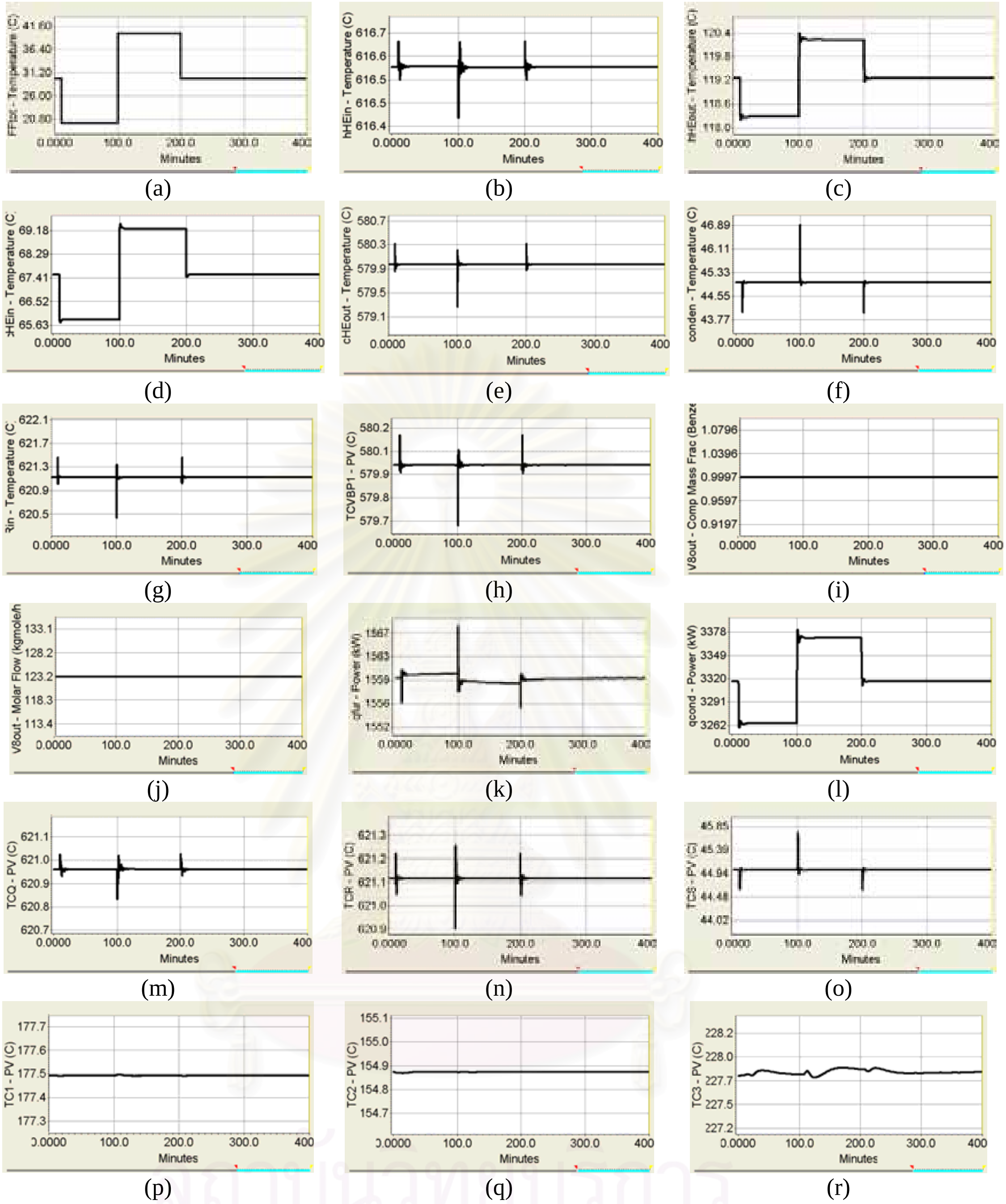


Figure 6.90 Dynamic Responses of the HDA Process of HIP 2 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.



(a) (b) (c) (d) (e) (f) (g) (h) (i) (j) (k) (l) (m) (n) (o) (p) (q) (r)

Figure 6.91 Dynamic Responses of the HDA Process of HIP 2 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

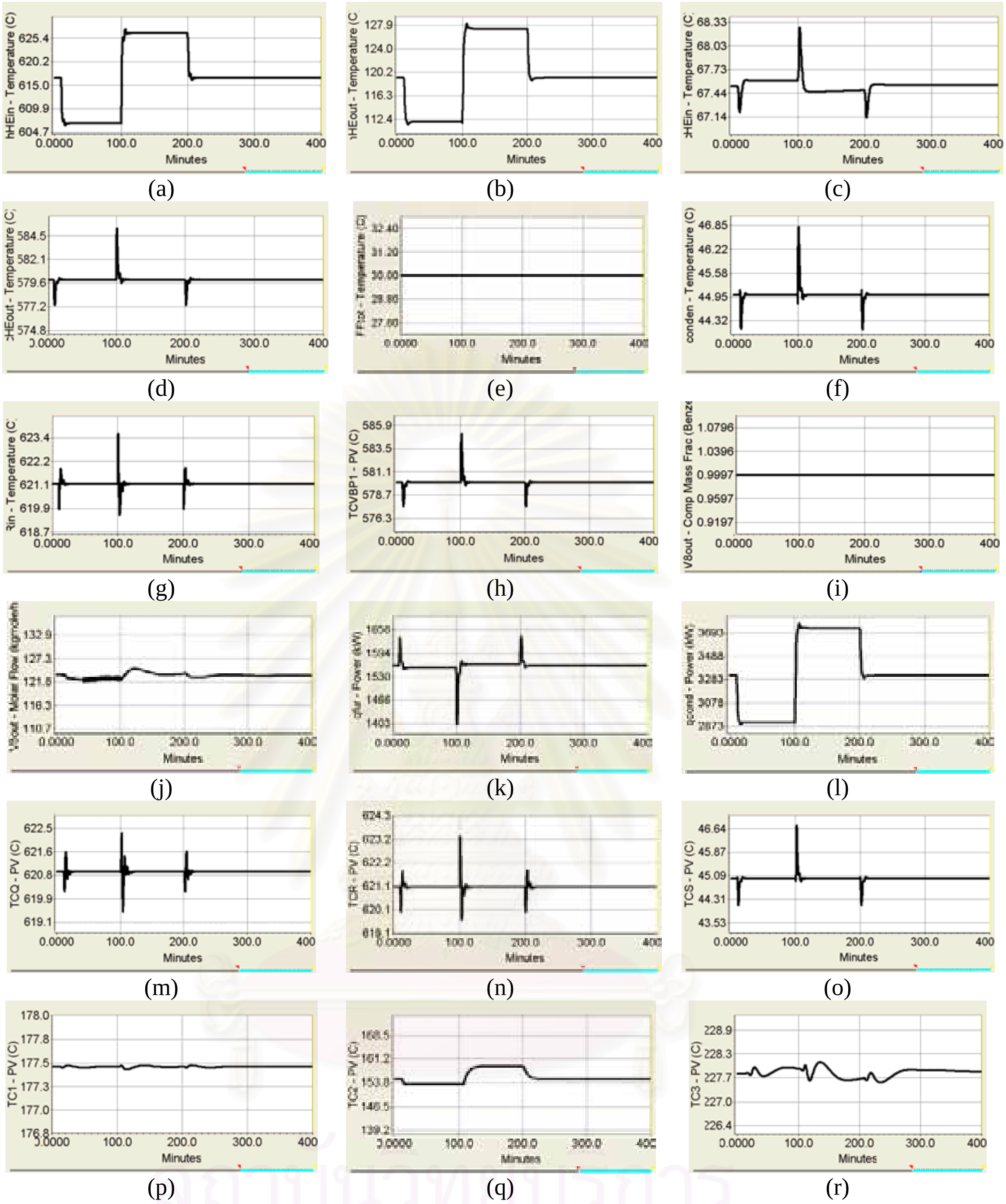


Figure 6.92 Dynamic Responses of the HDA Process of HIP 2 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) FEHE bypass stream, (i) benzene purity, (j) benzene flowrate, (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.

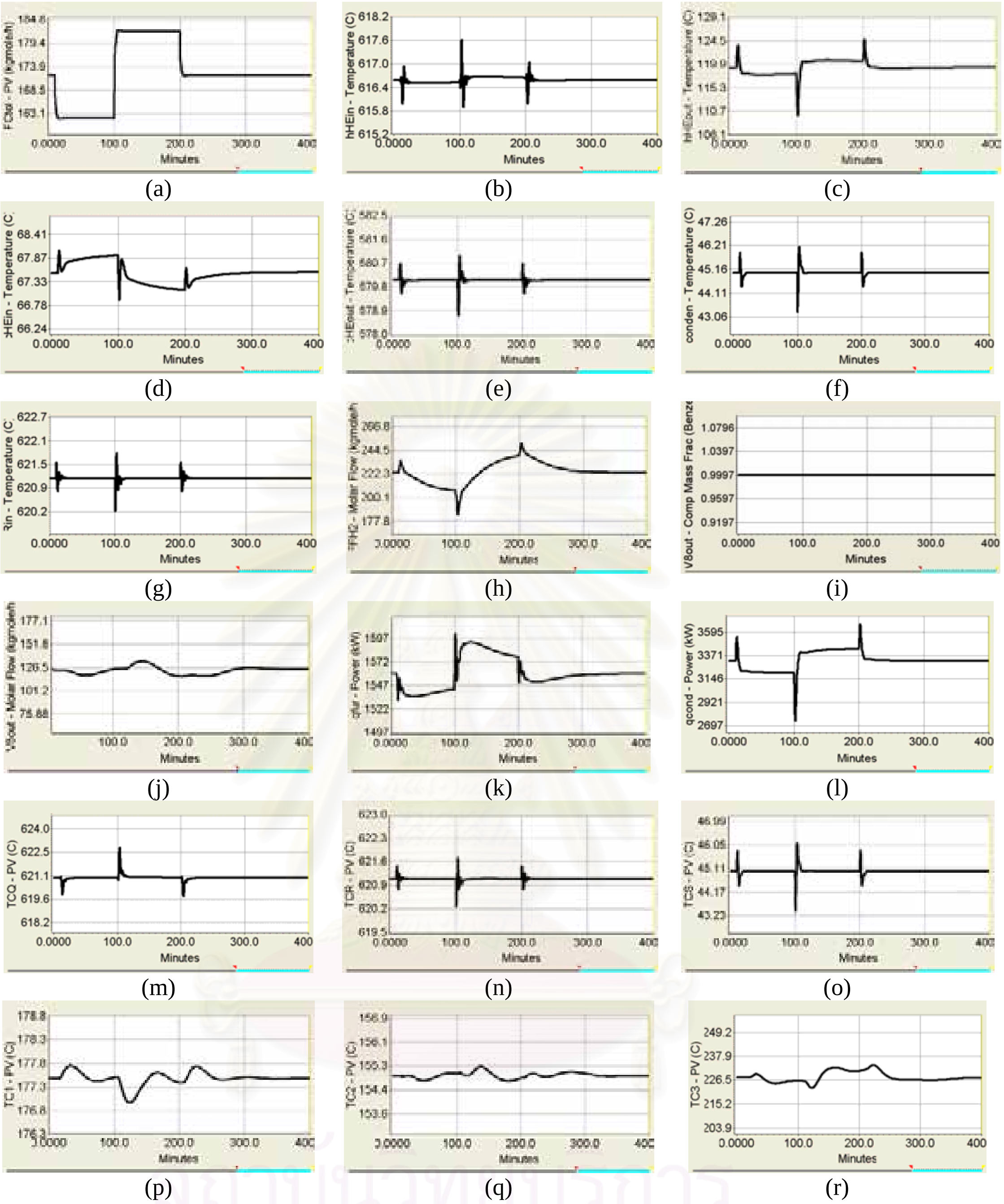


Figure 6.93 Dynamic Responses of the HDA Process of HIP 2 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

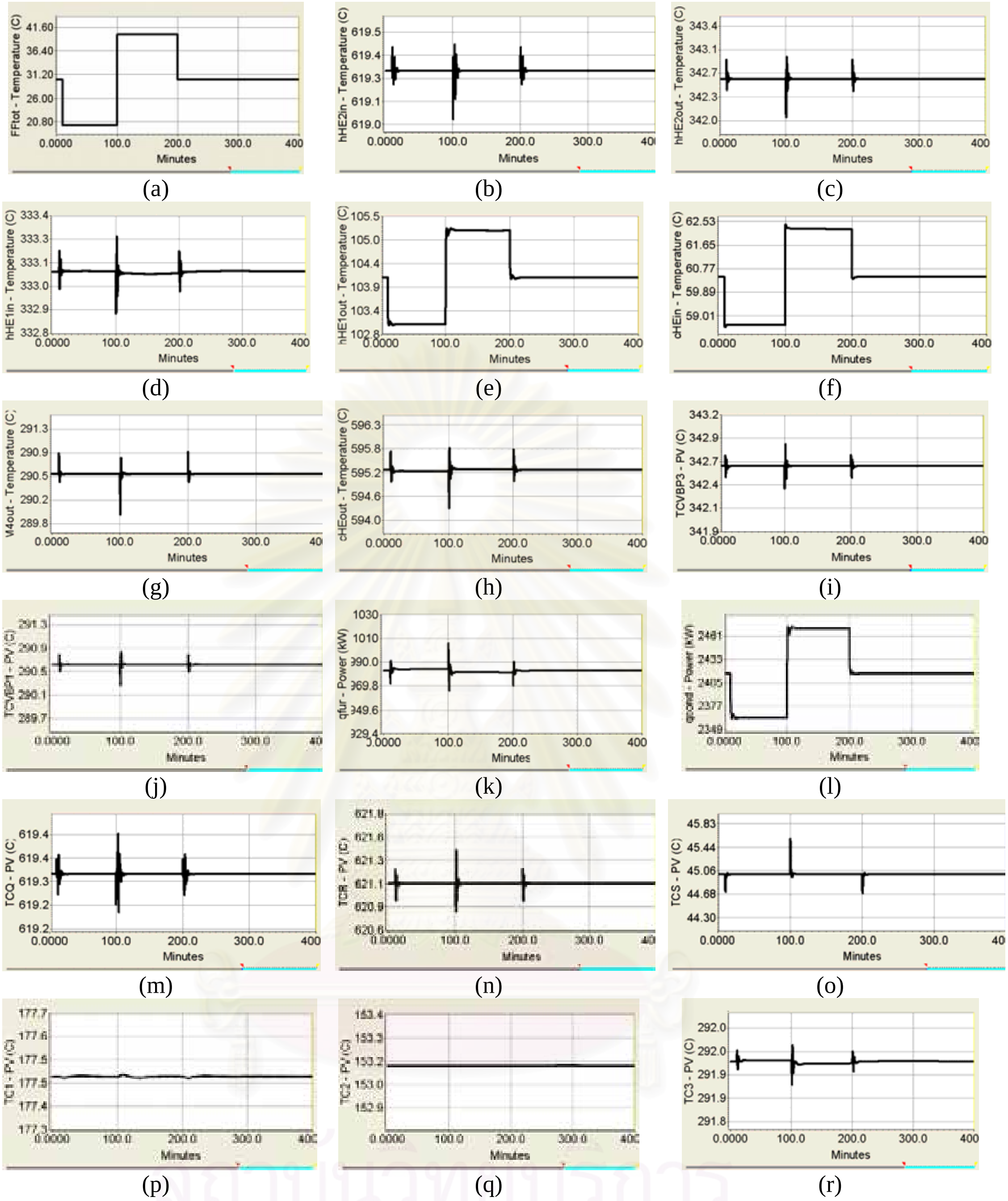


Figure 6.94 Dynamic Responses of the HDA Process of HIP 3 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

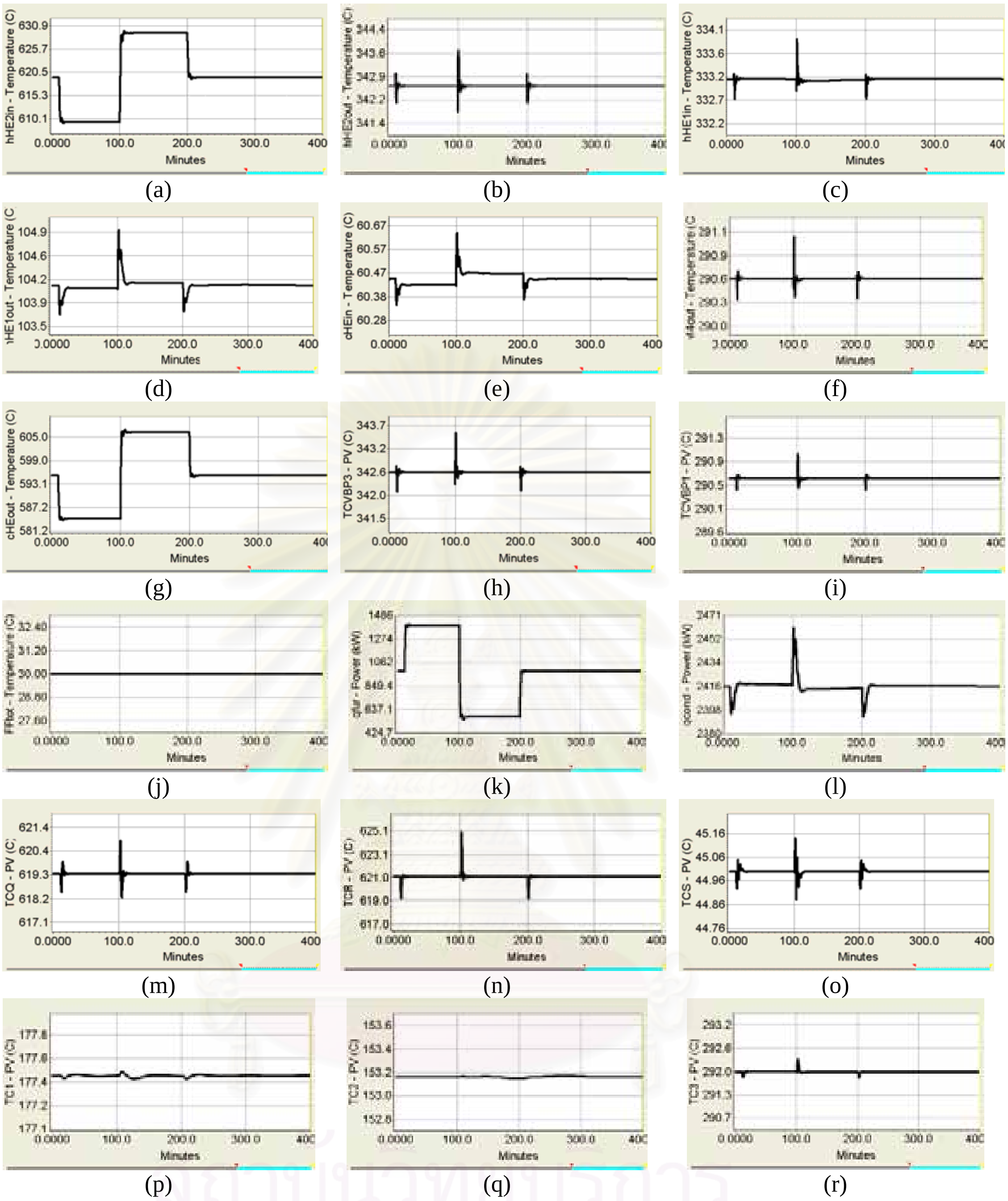


Figure 6.95 Dynamic Responses of the HDA Process of HIP 3 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

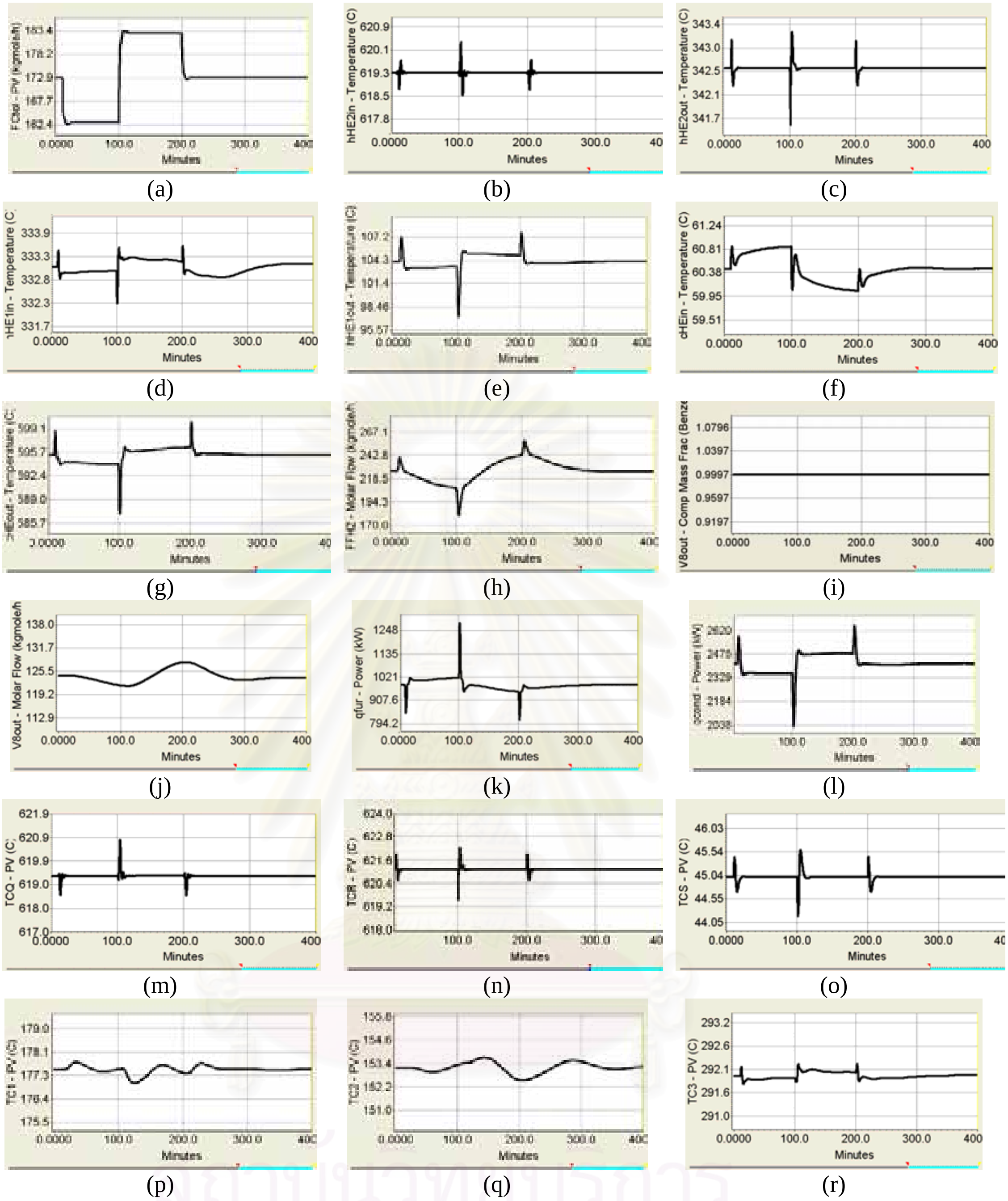


Figure 6.96 Dynamic Responses of the HDA Process of HIP 3 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

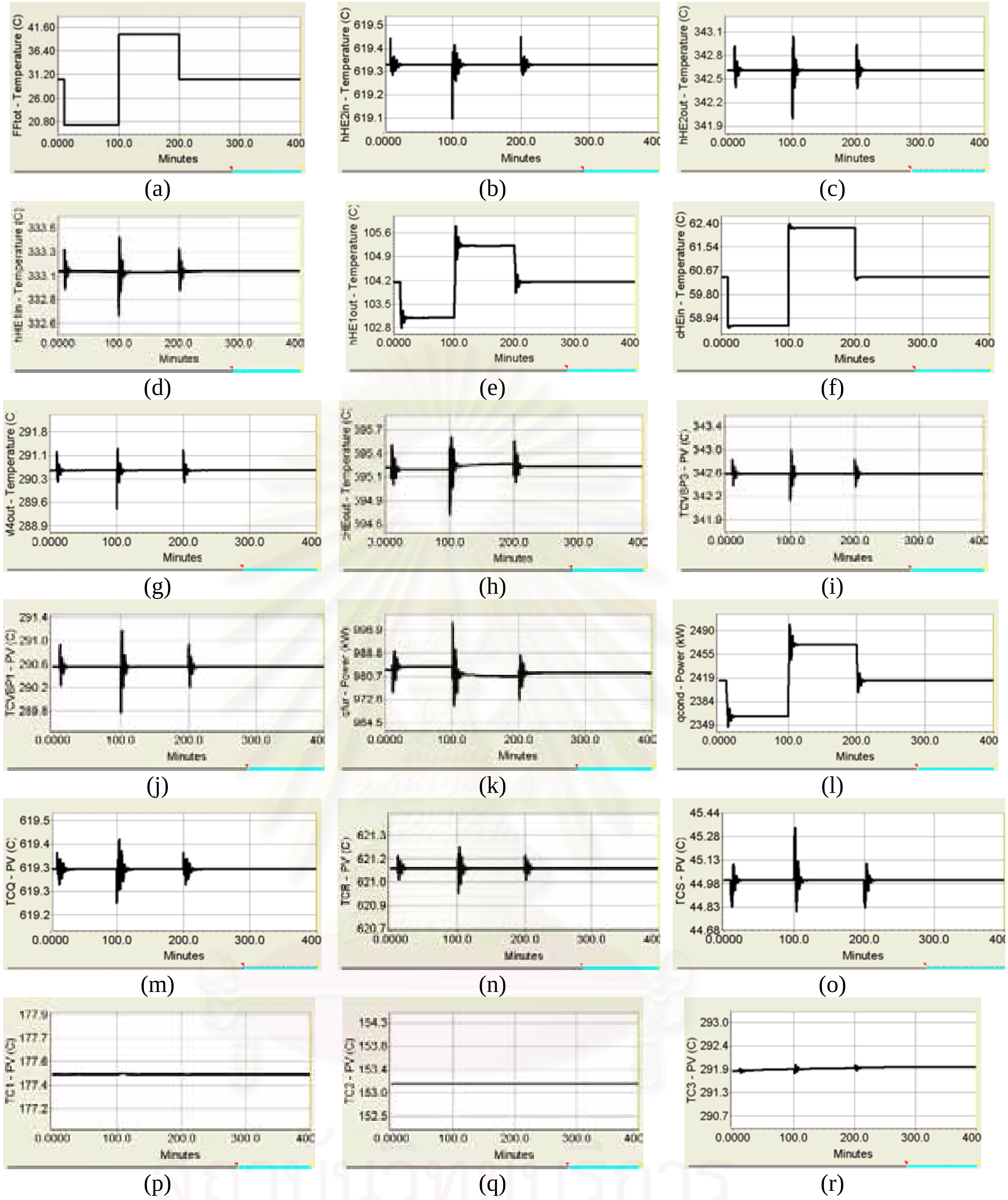


Figure 6.97 Dynamic Responses of the HDA Process of HIP 3 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

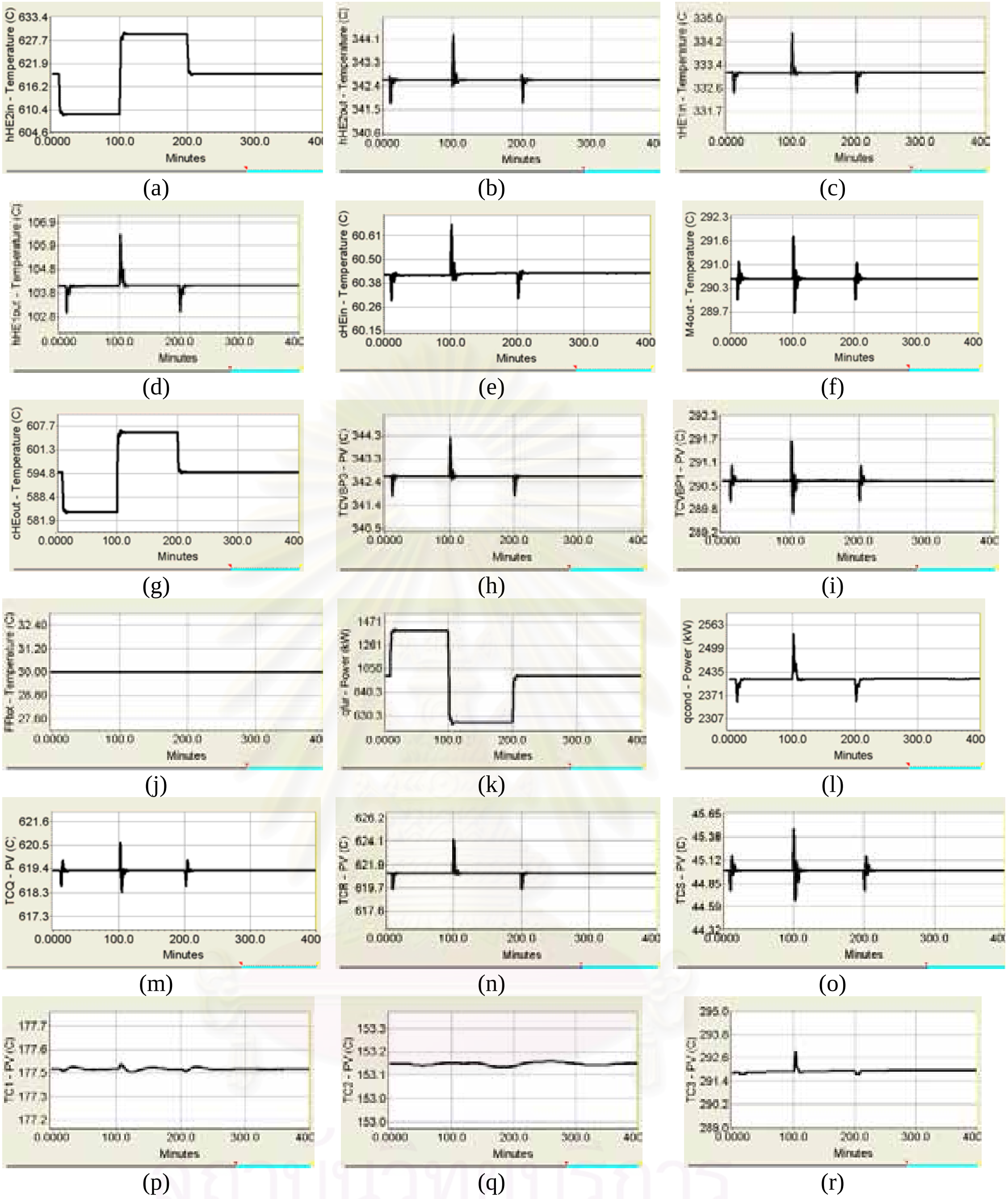


Figure 6.98 Dynamic Responses of the HDA Process of HIP 3 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

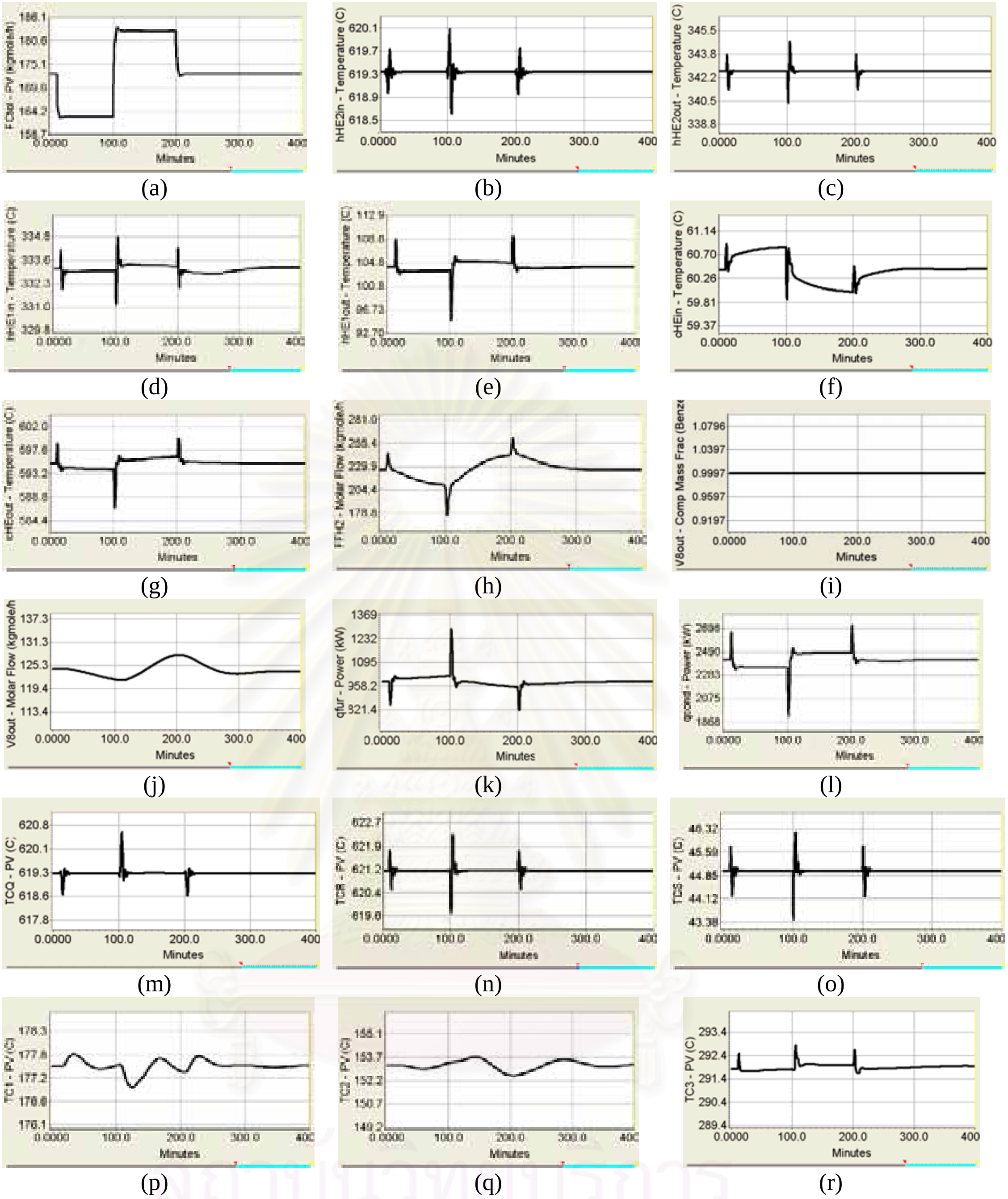


Figure 6.99 Dynamic Responses of the HDA Process of HIP 3 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

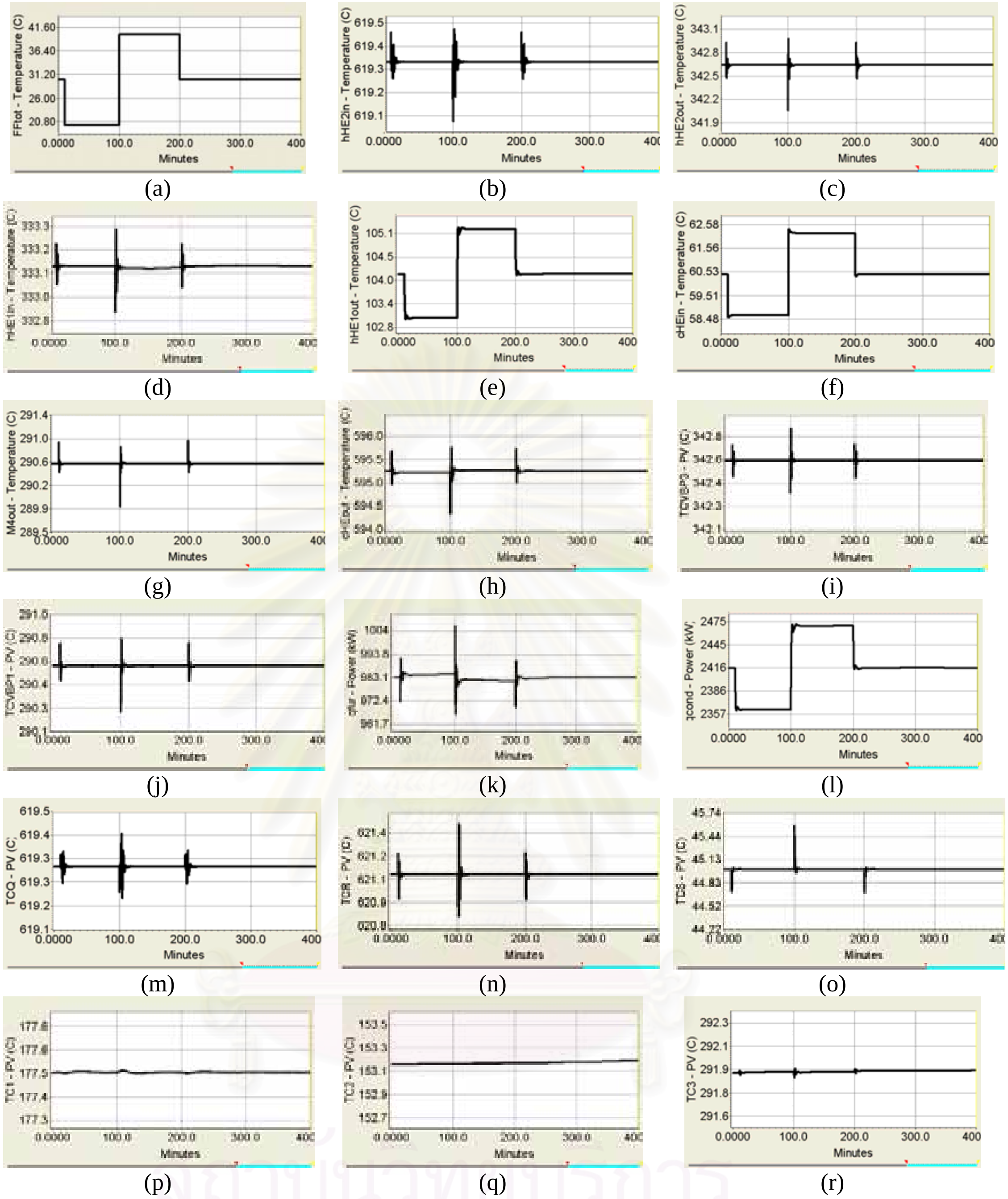


Figure 6.100 Dynamic Responses of the HDA Process of HIP 3 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

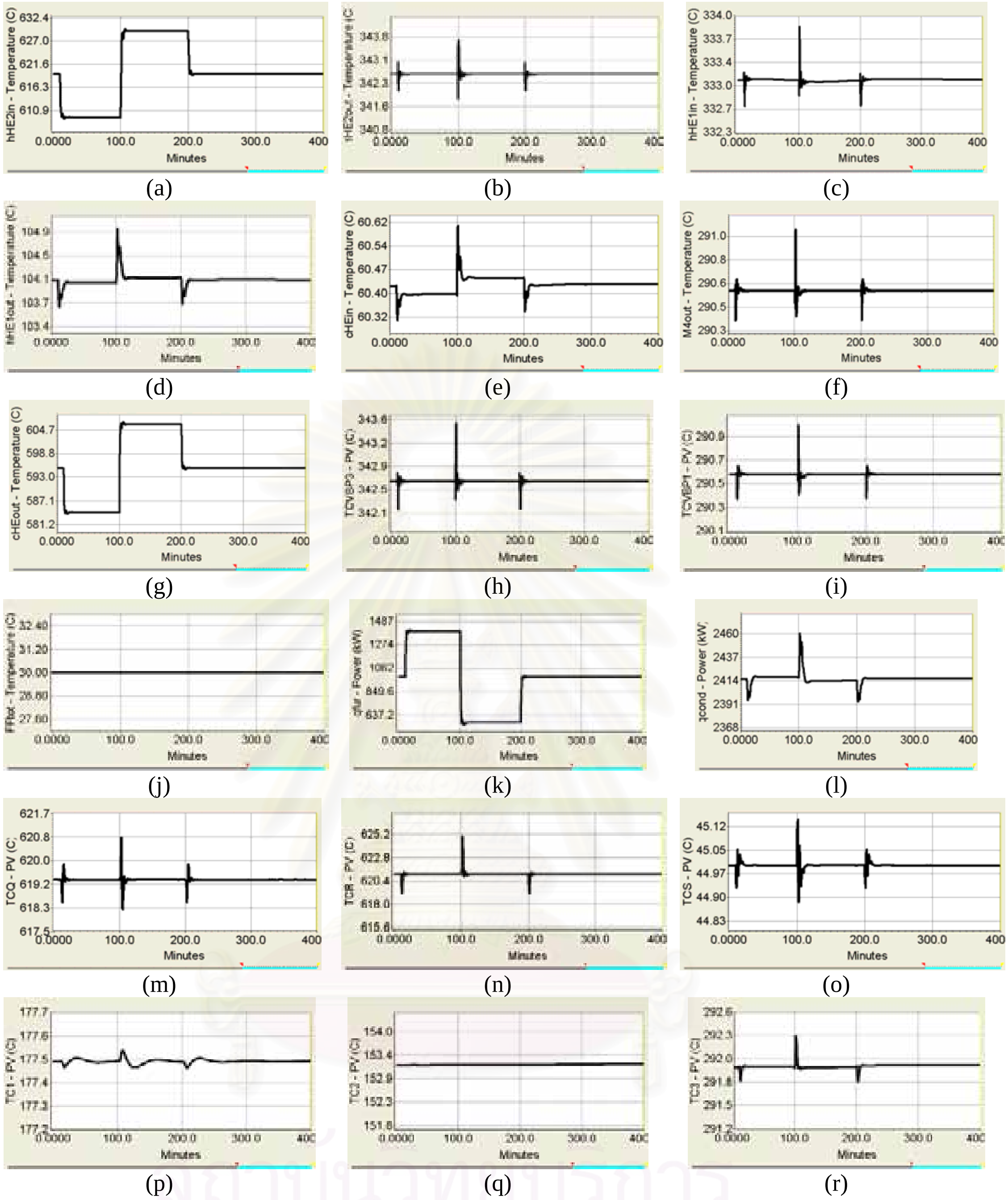


Figure 6.101 Dynamic Responses of the HDA Process of HIP 3 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

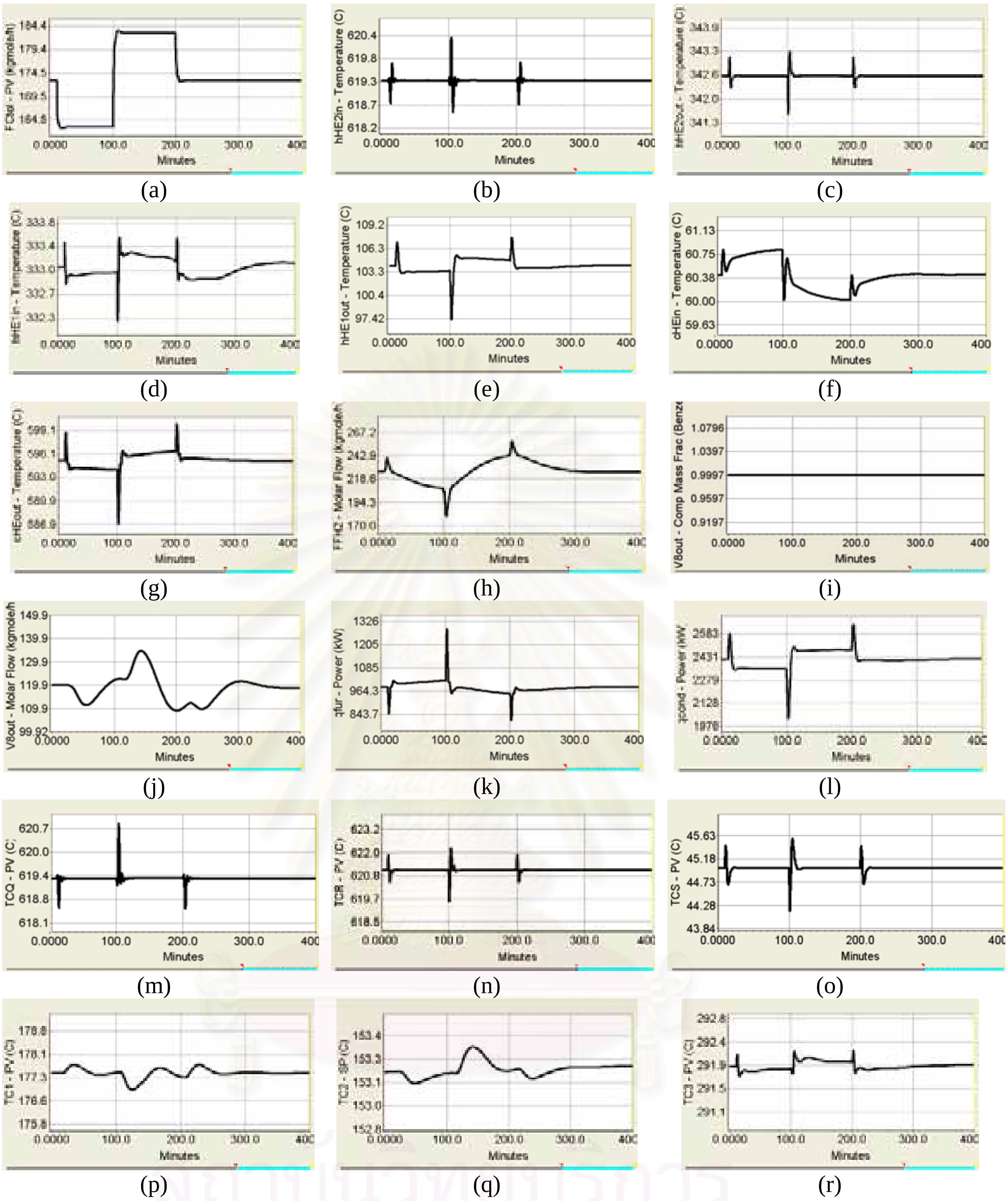
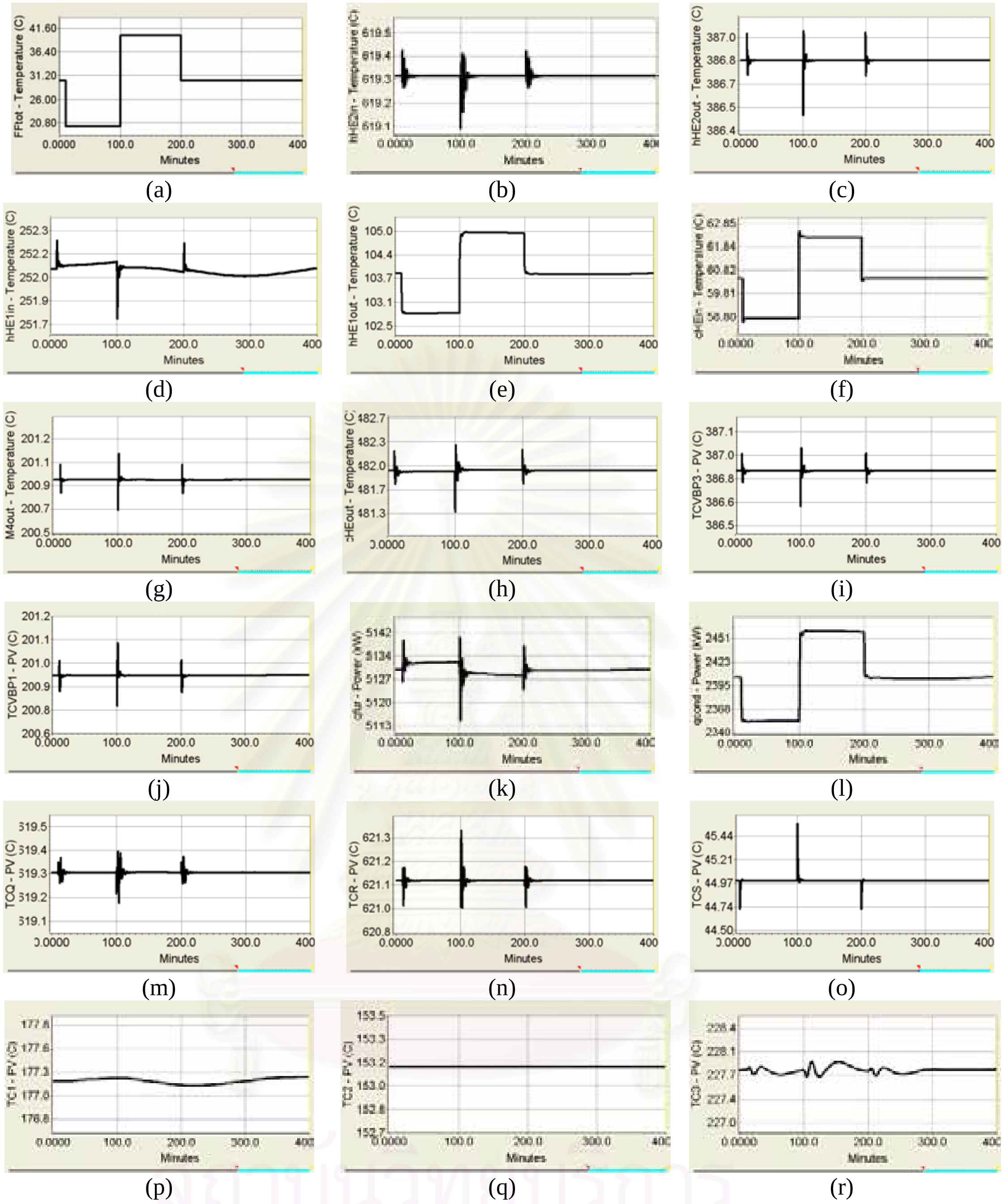


Figure 6.102 Dynamic Responses of the HDA Process of HIP 3 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.



(a) (b) (c) (d) (e) (f) (g) (h) (i) (j) (k) (l) (m) (n) (o) (p) (q) (r)

Figure 6.103 Dynamic Responses of the HDA Process of HIP 4 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

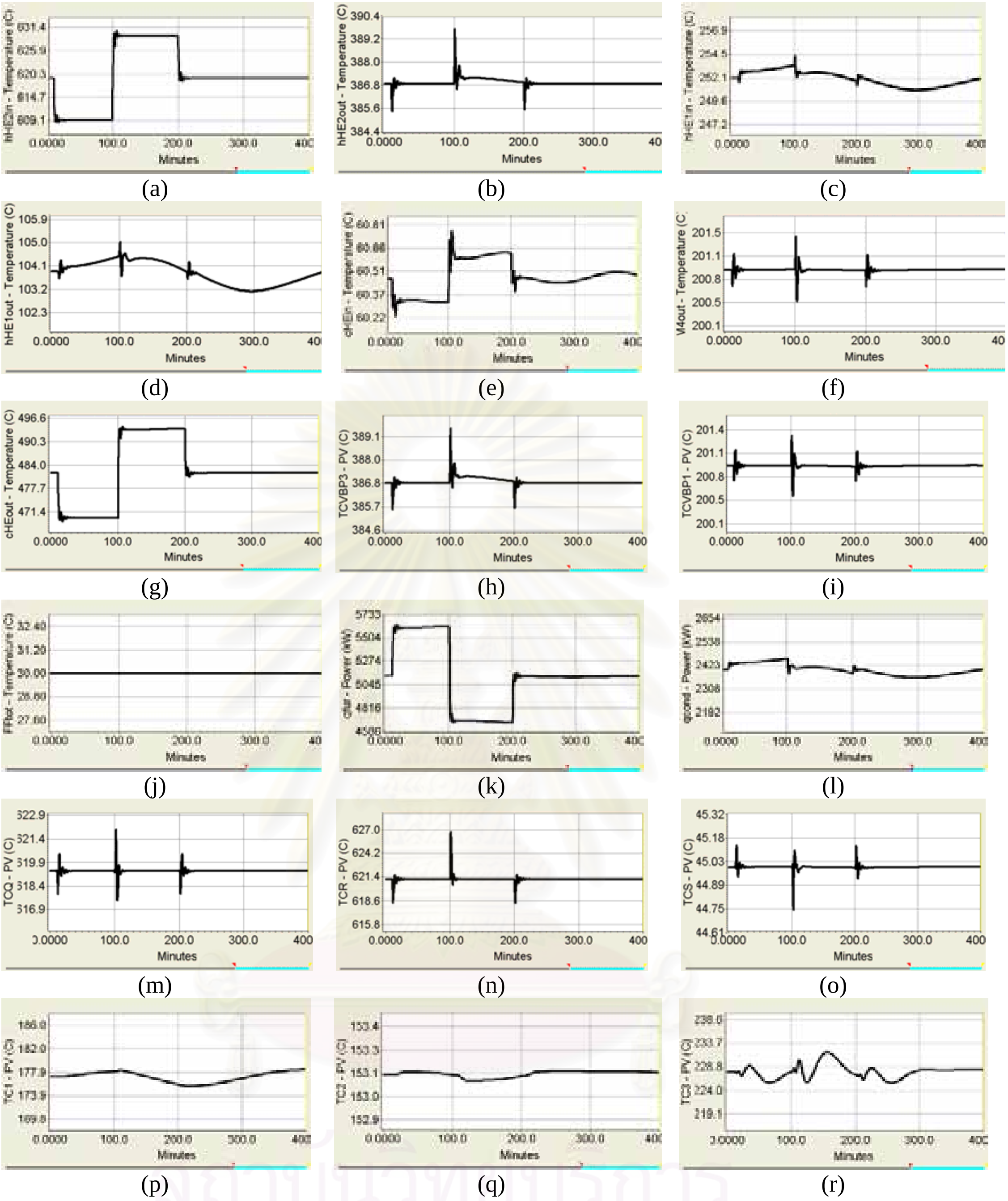


Figure 6.104 Dynamic Responses of the HDA Process of HIP 4 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

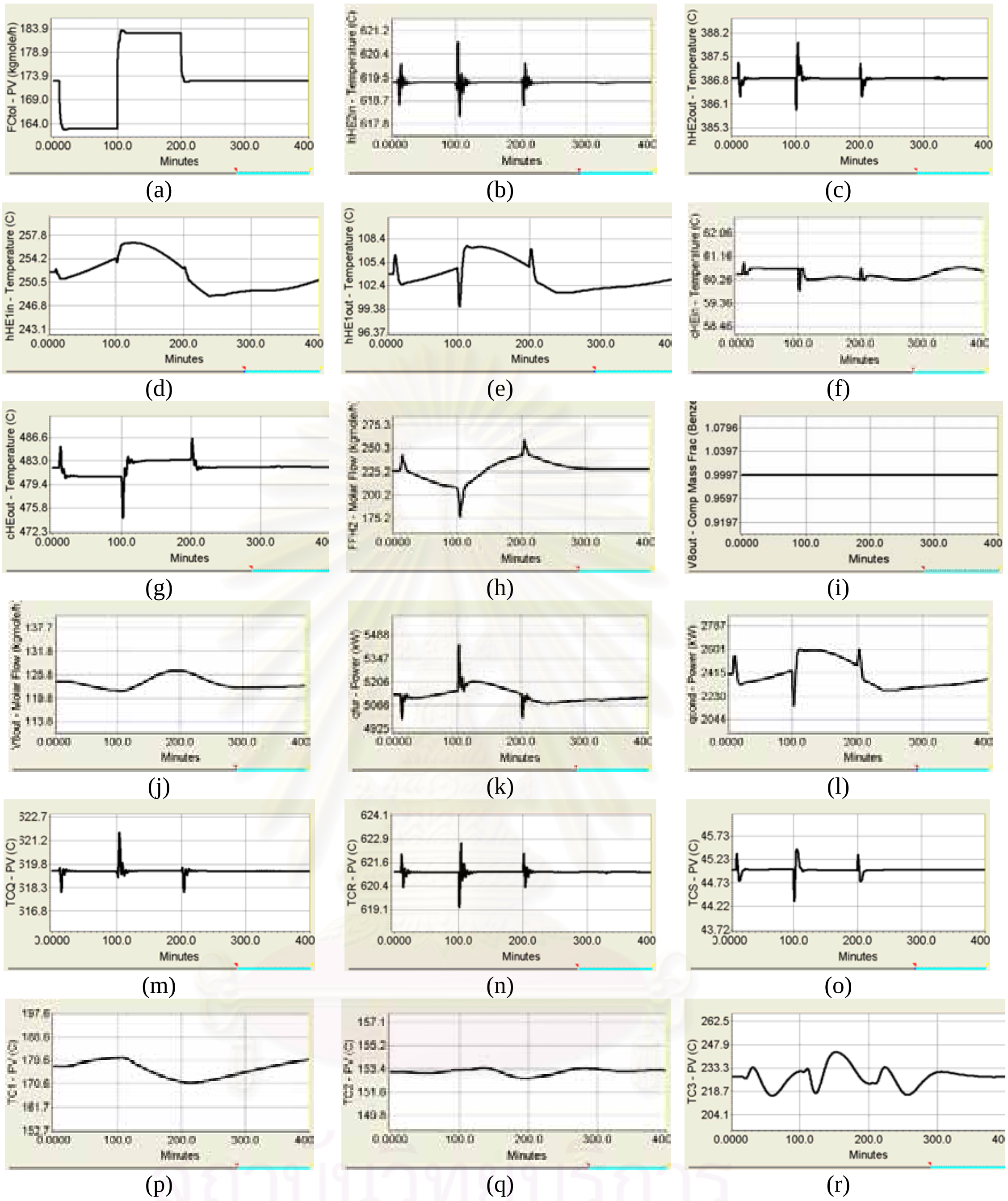


Figure 6.105 Dynamic Responses of the HDA Process of HIP 4 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

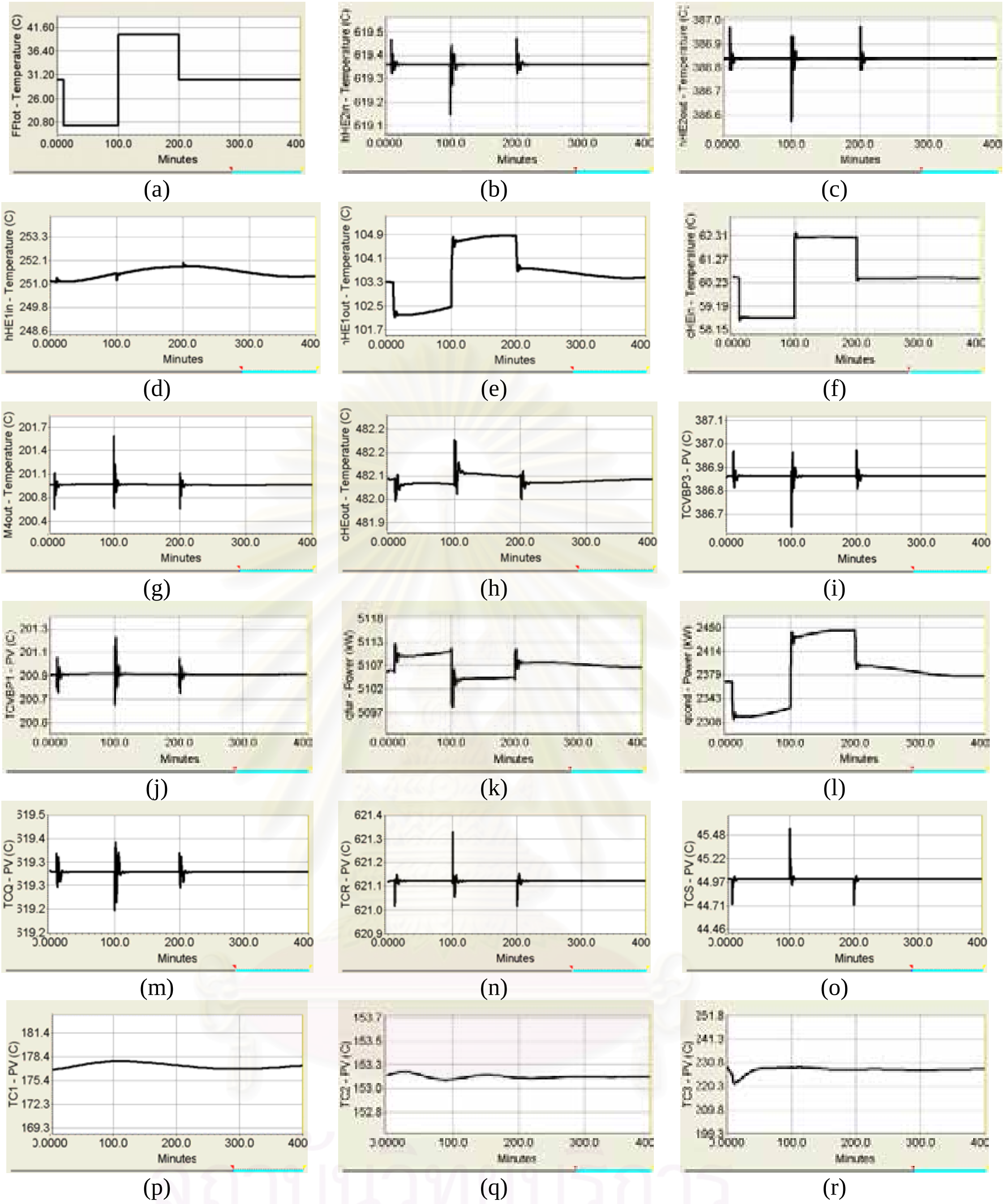


Figure 6.106 Dynamic Responses of the HDA Process of HIP 4 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

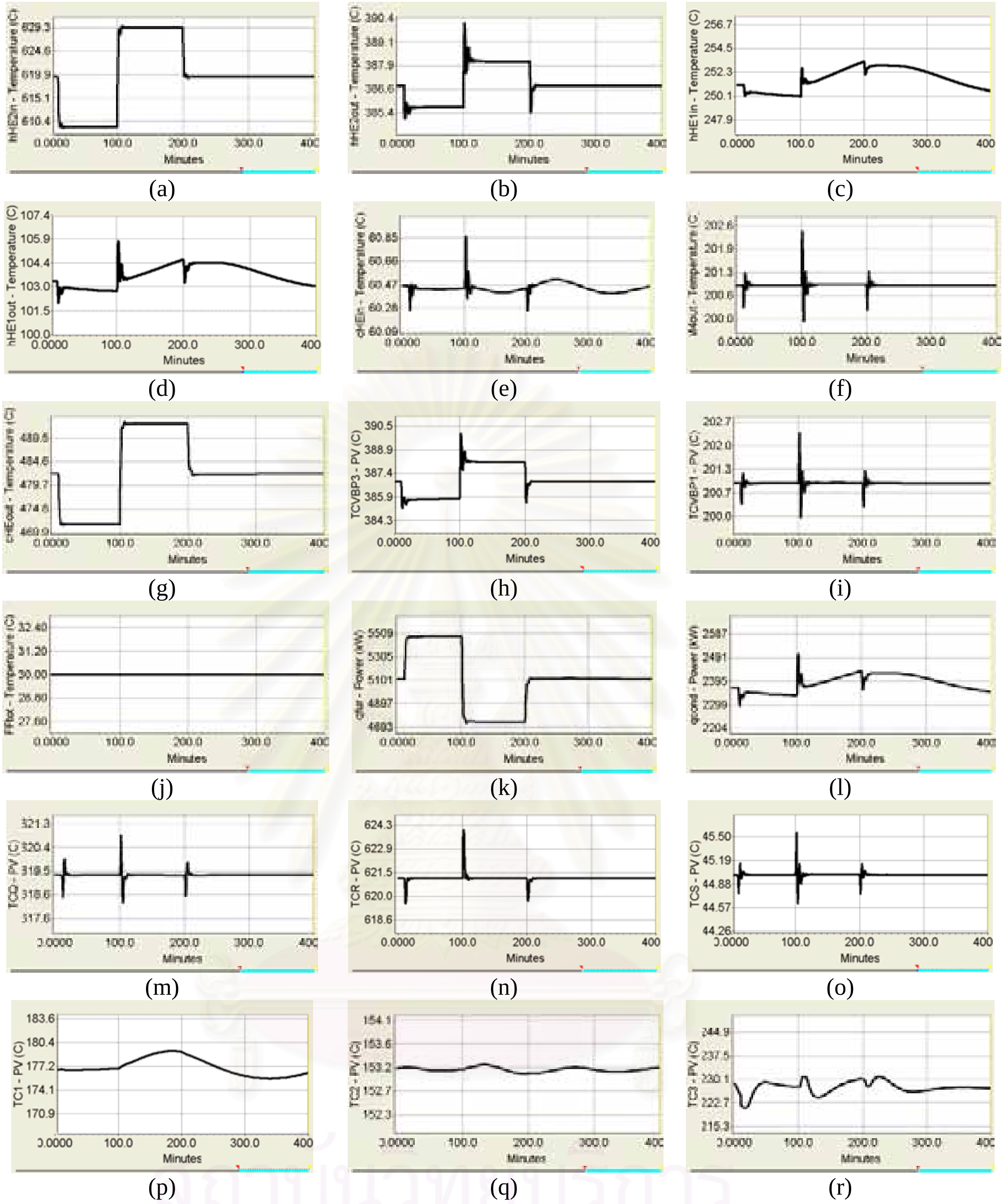


Figure 6.107 Dynamic Responses of the HDA Process of HIP 4 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

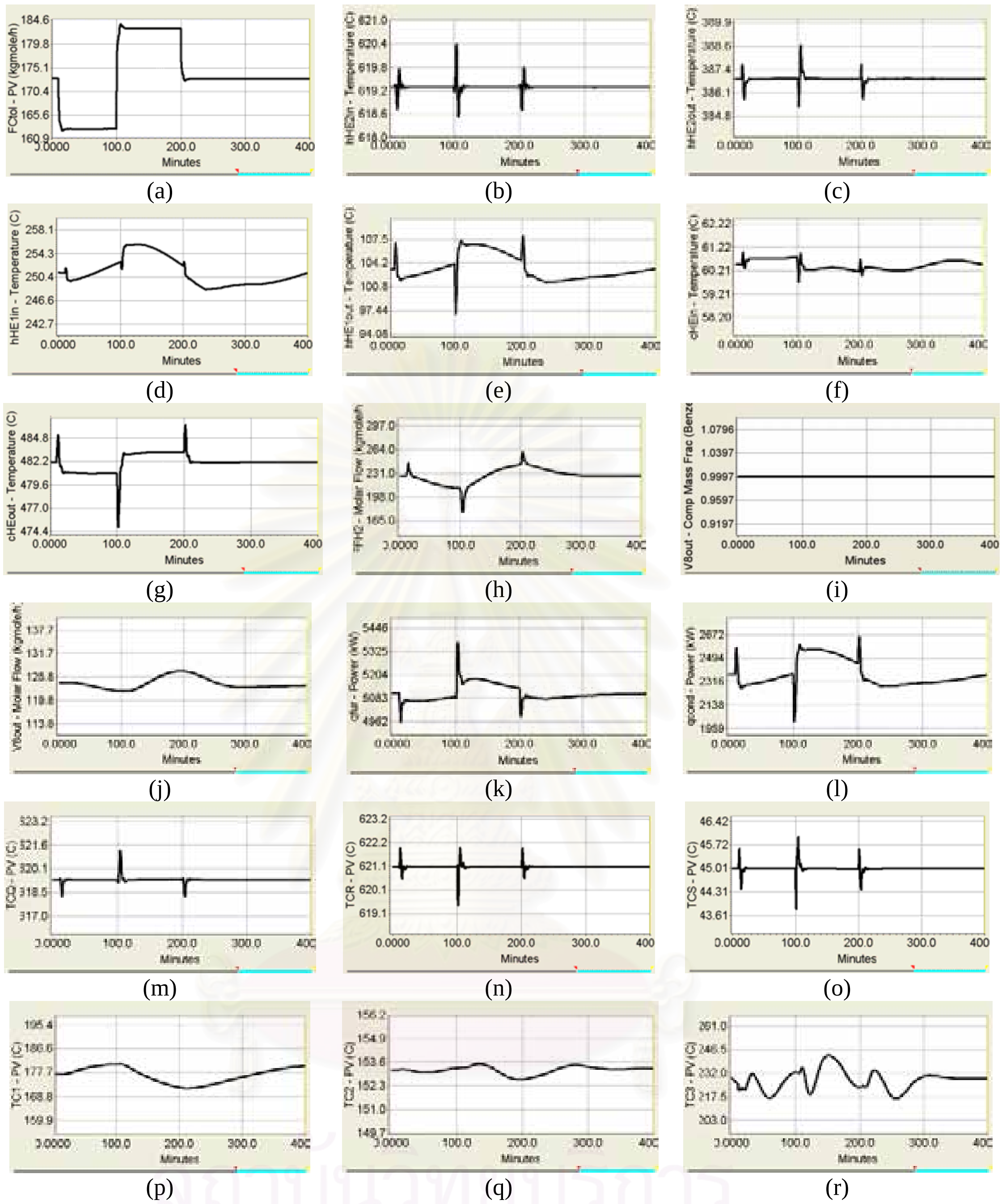


Figure 6.108 Dynamic Responses of the HDA Process of HIP 4 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

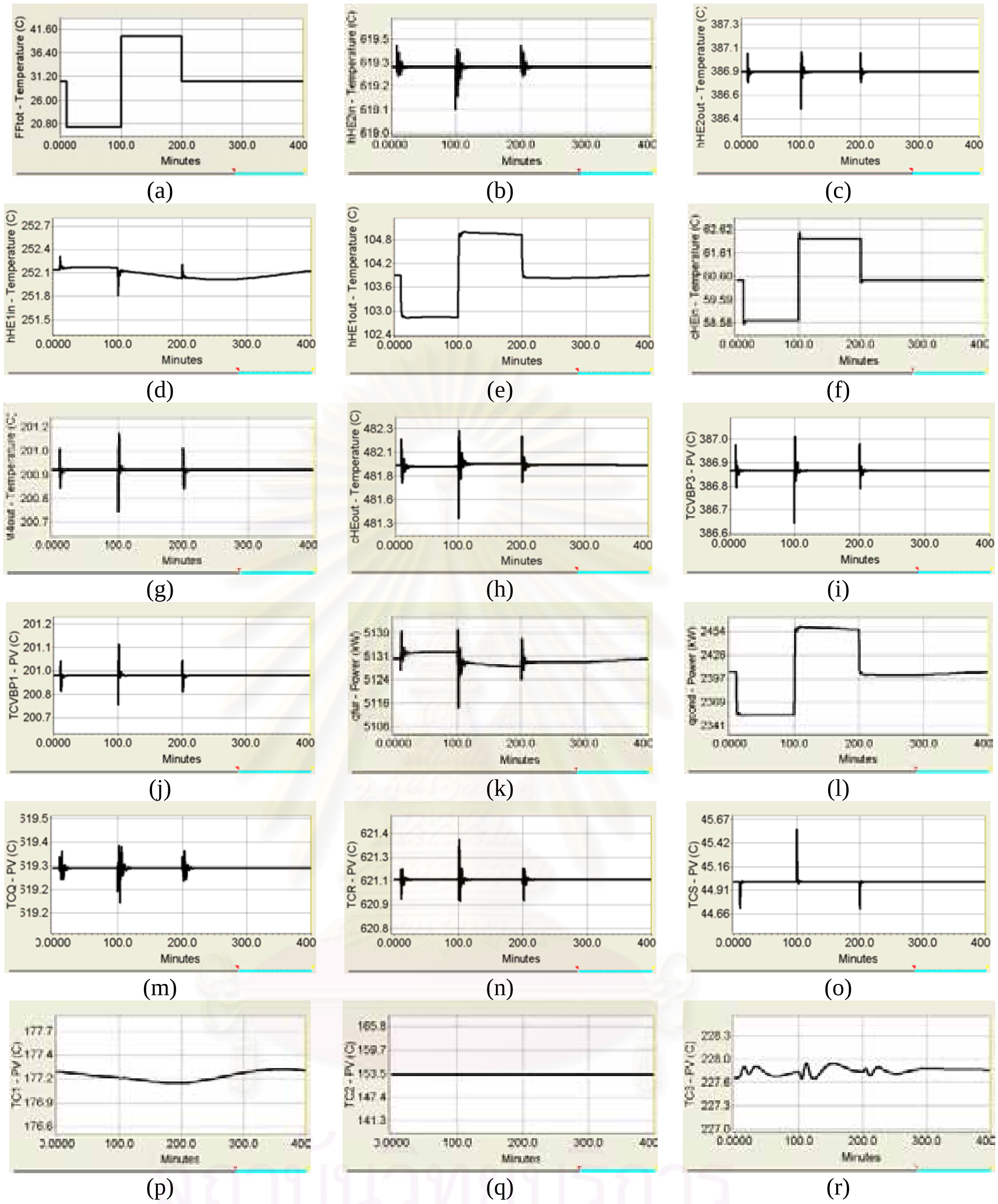


Figure 6.109 Dynamic Responses of the HDA Process of HIP 4 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold inlet temperature, (h) FEHE2 cold outlet temperature, (i) FEHE2 bypass stream, (j) FEHE1 bypass stream, (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.

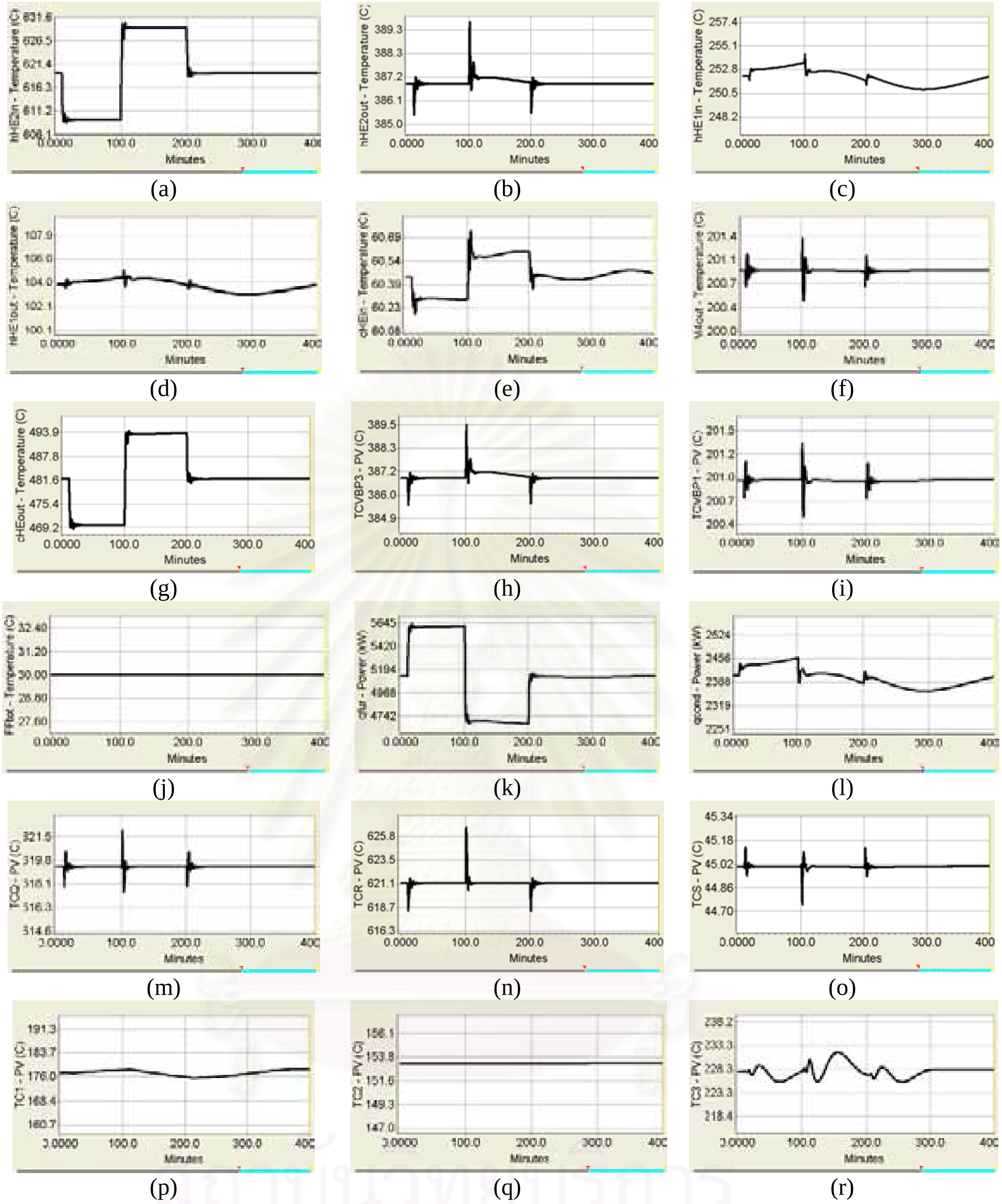


Figure 6.110 Dynamic Responses of the HDA Process of HIP 4 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE2 hot inlet temperature, (b) FEHE2 hot outlet temperature, (c) FEHE1 hot inlet temperature, (d) FEHE1 hot outlet temperature, (e) FEHE1 cold inlet temperature, (f) FEHE2 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) FEHE2 bypass stream, (i) FEHE1 bypass stream, (j) Fresh feed toluene 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.

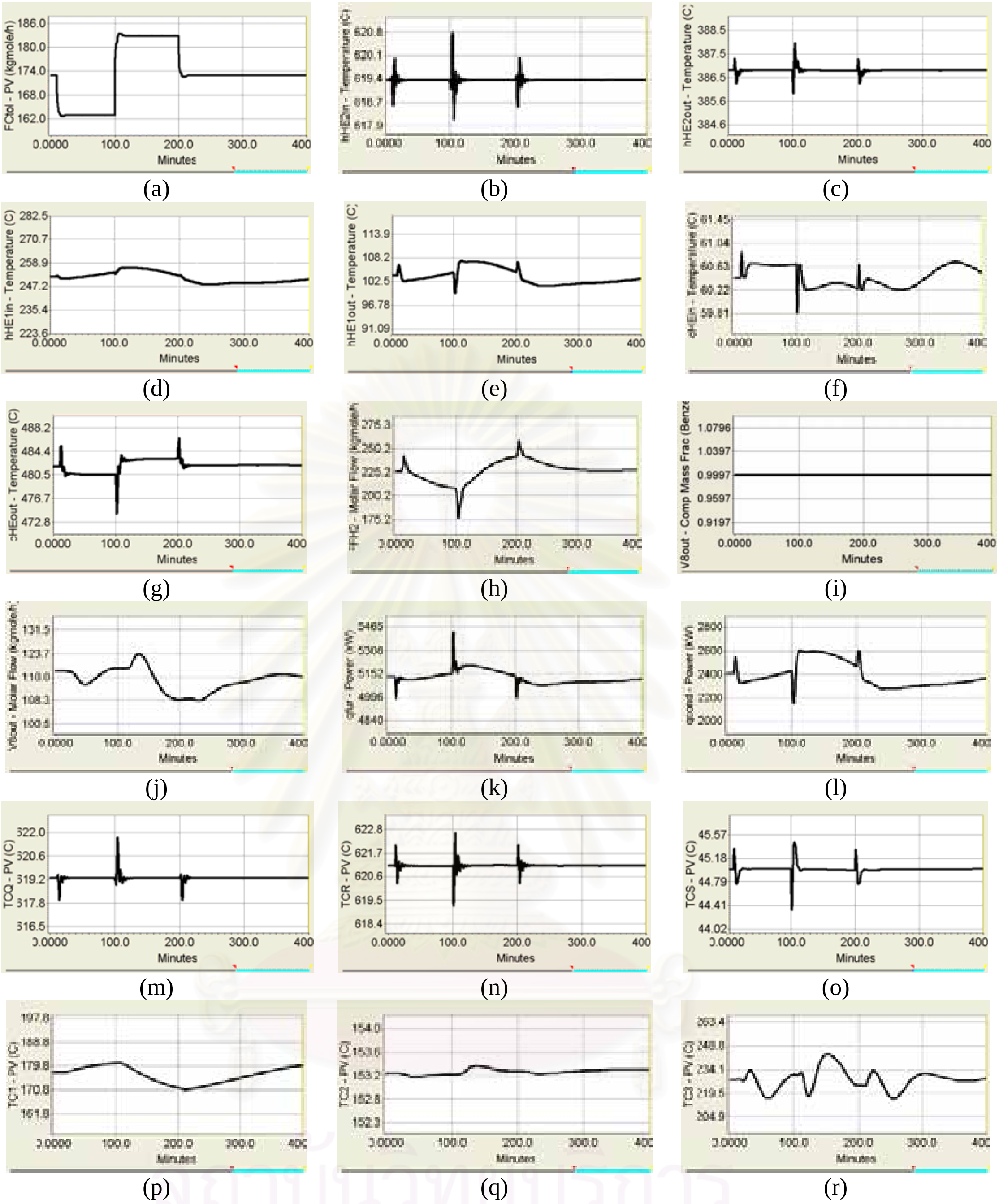


Figure 6.111 Dynamic Responses of the HDA Process of HIP 4 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE2 hot inlet temperature, (c) FEHE2 hot outlet temperature, (d) FEHE1 hot inlet temperature, (e) FEHE1 hot outlet temperature, (f) FEHE1 cold inlet temperature, (g) FEHE2 cold outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

6.7 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:

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

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. Table 6.25(a, b and c) - Table 6.28(a, b and c) show the IAE results of typical and heat-integrated plant of HDA process in alternative structures

As can be seen the similarity result the change in the disturbance loads of the hot and cold steam and the change in the disturbance loads of the total toluene feed flowrates, the value of IAE in process CS1 is smaller than another alternatives. All the change in the disturbance loads on the all typical (ALs) and all heat-integrated plant of HDA process (HIPs) case compared with the control structures (CS1, CS2, and CS3), i.e. the value of IAE in process CS1 is smaller than those on CS3 and CS2 in alternative structures respectively.

As can be seen, The IAE of control structures CS1 close to the IAE of CS3 but IAE of CS2 to high because CS2 exhibited very slow dynamics and is more sensitive to the disturbances. However, CS1 is the best control structure for handle disturbances due to it gives better control performances than CS3 and CS2 respectively.

Table 6.9a the IAE Values of Typical (AL1) and Heat-Integrated Plant of HDA process (HIP1) to a Change in the Disturbance Load of Hot Stream (Reactor Product Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL1	HIP1	AL1	HIP1	AL1	HIP1
TC1	0.0430	0.0158	0.0196	0.0031	0.0432	0.0157
TC2	0.0017	0.0075	0.0010	0.0307	0.0109	0.0023
TC3	0.3535	0.4046	0.1418	0.1513	0.3486	0.4110
TCVBP1	0.0847	0.0707	0.6434	0.4726	0.0847	0.0707
TCR	0.1393	0.0878	0.1156	0.1603	0.1393	0.0879
TCQ	0.1180	0.0779	0.0817	0.1296	0.1181	0.0779
TCS	0.1068	0.0718	0.0565	0.0653	0.1068	0.0718
Total	0.8471	0.7362	1.0596	1.0129	0.8516	0.7373

Table 6.9b the IAE Values of Typical (AL1) and Heat-Integrated Plant of HDA process (HIP1) to a Change in the Disturbance Load of Cold Stream (Reactor Feed Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL1	HIP1	AL1	HIP1	AL1	HIP1
TC1	0.1358	0.0115	0.1139	0.0099	0.1492	0.0111
TC2	0.0001	0.0151	0.0003	0.0061	0.0109	0.0045
TC3	0.0652	0.6009	0.1027	0.4208	0.0733	0.6331
TCVBP1	0.0023	0.0027	0.0065	0.0194	0.0037	0.0027
TCR	0.1421	0.0382	0.5028	0.3093	0.2144	0.0378
TCQ	0.1058	0.0287	0.4349	0.2603	0.1797	0.0277
TCS	0.3871	0.0710	0.3283	0.1372	0.4062	0.0701
Total	0.8383	0.7681	1.4893	1.1629	1.0373	0.7871

Table 6.9c the IAE Values of Typical (AL1) and Heat-Integrated Plant of HDA process (HIP1) to a Change in the Total Toluene Feed Flowrates

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL1	HIP1	AL1	HIP1	AL1	HIP1
TC1	0.1167	0.0408	0.1004	0.0370	0.1167	0.0407
TC2	0.0028	0.0223	0.0025	0.0204	0.0064	0.0043
TC3	1.0673	0.9971	0.9666	0.8978	1.0581	1.0038
TCVBP1	0.0209	0.0032	0.0738	0.0402	0.0501	0.0024
TCR	0.0079	0.0078	0.0413	0.1239	0.0153	0.0152
TCQ	0.0211	0.0111	0.0349	0.0665	0.0220	0.0111
TCS	0.0203	0.0066	0.0392	0.0266	0.0185	0.0047
Total	1.2571	1.0889	1.2587	1.2124	1.2871	1.0822

Table 6.10a the IAE Values of Typical (AL2) and Heat-Integrated Plant of HDA process (HIP2) to a Change in the Disturbance Load of Hot Stream (Reactor Product Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL2	HIP2	AL2	HIP2	AL2	HIP2
TC1	0.0731	0.0381	0.0661	0.0380	0.0729	0.0383
TC2	0.0037	0.0152	0.0033	0.0154	0.0111	0.0117
TC3	1.1112	0.9709	1.0356	0.9627	1.1470	0.9824
TCVBP1	0.0031	0.0033	0.0263	0.0222	0.0031	0.0033
TCR	0.0013	0.0077	0.0184	0.0664	0.0013	0.0077
TCQ	0.0118	0.0129	0.0173	0.0385	0.0118	0.0129
TCS	0.0130	0.0068	0.0452	0.0214	0.0130	0.0068
Total	1.2172	1.0549	1.2122	1.1647	1.2602	1.0630

Table 6.10b the IAE Values of Typical (AL2) and Heat-Integrated Plant of HDA process (HIP2) to a Change in the Disturbance Load of Cold Stream (Reactor Feed Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL2	HIP2	AL2	HIP2	AL2	HIP2
TC1	0.1811	0.0144	0.0728	0.0104	0.2526	0.0151
TC2	0.0019	0.0174	0.0003	0.0090	0.0031	0.0187
TC3	0.1012	0.7368	0.0318	0.4884	0.2968	0.7442
TCVBP1	0.0016	0.0022	0.0045	0.0125	0.0016	0.0022
TCR	0.0917	0.0419	0.3294	0.2834	0.0917	0.0424
TCQ	0.0940	0.0319	0.2043	0.1700	0.0940	0.1064
TCS	0.3629	0.1061	0.6526	0.2521	0.3632	0.0322
Total	0.8343	0.9508	1.2957	1.2257	1.1031	0.9612

Table 6.10c the IAE Values of Typical (AL2) and Heat-Integrated Plant of HDA process (HIP2) to a Change in the Total Toluene Feed Flowrates

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL2	HIP2	AL2	HIP2	AL2	HIP2
TC1	0.0731	0.0381	0.0661	0.0380	0.0729	0.0383
TC2	0.0037	0.0152	0.0033	0.0154	0.0111	0.0117
TC3	1.1112	0.9709	1.0356	0.9627	1.1470	0.9824
TCVBP1	0.0031	0.0033	0.0263	0.0222	0.0031	0.0033
TCR	0.0013	0.0077	0.0184	0.0664	0.0013	0.0077
TCQ	0.0118	0.0129	0.0173	0.0385	0.0118	0.0129
TCS	0.0130	0.0068	0.0452	0.0214	0.0130	0.0068
Total	1.2172	1.0549	1.2122	1.1647	1.2602	1.0630

Table 6.11a the IAE Values of Typical (AL3) and Heat-Integrated Plant of HDA process (HIP3) to a Change in the Disturbance Load of Hot Stream (Reactor Product Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL3	HIP3	AL3	HIP3	AL3	HIP3
TC1	0.0474	0.0217	0.0446	0.0133	0.0470	0.0221
TC2	0.3429	0.0191	0.2705	0.2766	0.5181	0.0922
TC3	0.0920	0.1023	0.0738	0.4517	0.0800	0.1323
TCVBP1	0.0265	0.0020	0.2205	0.0064	0.0262	0.0020
TCVBP2	0.0286	0.0739	0.1099	0.1378	0.0286	0.0742
TCR	0.1260	0.1486	0.1270	0.0991	0.1259	0.1486
TCQ	0.0847	0.0890	0.0747	0.0563	0.0846	0.0889
TCS	0.0060	0.0110	0.0344	0.0210	0.0059	0.0109
Total	0.7540	0.4676	0.9553	1.0624	0.9164	0.5713

Table 6.11b the IAE Values of Typical (AL3) and Heat-Integrated Plant of HDA process (HIP3) to a Change in the Disturbance Load of Cold Stream (Reactor Feed Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL3	HIP3	AL3	HIP3	AL3	HIP3
TC1	0.2726	0.0404	0.2091	0.0297	0.3250	0.0403
TC2	0.0008	0.0285	0.0004	0.5139	0.0098	0.1835
TC3	0.0716	0.0101	0.0341	0.1659	0.0675	0.0452
TCVBP1	0.0019	0.0114	0.0085	0.0395	0.0019	0.0114
TCVBP2	0.1815	0.0029	0.4116	0.0056	0.1810	0.0029
TCR	0.2270	0.2807	0.1833	0.2254	0.2266	0.2797
TCQ	0.1754	0.1762	0.1242	0.1372	0.1745	0.1751
TCS	0.2647	0.2956	0.3356	0.3872	0.2644	0.2955
Total	1.1956	0.8459	1.3070	1.5045	1.2508	1.0337

Table 6.11c the IAE Values of Typical (AL3) and Heat-Integrated Plant of HDA process (HIP3) to a Change in the Total Toluene Feed Flowrates

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL3	HIP3	AL3	HIP3	AL3	HIP3
TC1	0.3233	0.0576	0.3249	0.0441	0.3234	0.0575
TC2	0.0142	0.0108	0.0134	0.0089	0.0231	0.0020
TC3	0.5991	0.5105	0.5944	0.6064	0.5834	0.5299
TCVBP1	0.0242	0.0101	0.1003	0.0226	0.0242	0.0101
TCVBP2	0.0226	0.0877	0.0796	0.2093	0.0225	0.0876
TCR	0.0206	0.2088	0.0223	0.1728	0.0205	0.2090
TCQ	0.0269	0.0321	0.0261	0.0257	0.0268	0.0321
TCS	0.0106	0.0279	0.0179	0.0309	0.0106	0.0279
Total	1.0416	0.9455	1.1789	1.1206	1.0346	0.9560

Table 6.12a the IAE Values of Typical (AL4) and Heat-Integrated Plant of HDA process (HIP4) to a Change in the Disturbance Load of Hot Stream (Reactor Product Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL4	HIP4	AL4	HIP4	AL4	HIP4
TC1	0.1919	0.3296	0.1893	0.3074	0.1910	0.3314
TC2	0.0205	0.0100	0.0227	0.0101	0.1462	0.0027
TC3	0.8276	0.7175	0.8628	0.7520	0.9540	0.7252
TCVBP1	0.0140	0.0023	0.0860	0.0047	0.0135	0.0023
TCVBP2	0.0237	0.0017	0.0702	0.0031	0.0235	0.0017
TCR	0.0238	0.0071	0.0221	0.0050	0.0241	0.0070
TCQ	0.0270	0.0074	0.0252	0.0062	0.0275	0.0074
TCS	0.0086	0.0036	0.0114	0.0046	0.0087	0.0036
Total	1.1371	1.0792	1.2896	1.0931	1.3885	1.0813

Table 6.12b the IAE Values of Typical (AL4) and Heat-Integrated Plant of HDA process (HIP4) to a Change in the Disturbance Load of Cold Stream (Reactor Feed Stream)

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL4	HIP4	AL4	HIP4	AL4	HIP4
TC1	0.1409	0.1704	0.0695	0.2210	0.1329	0.3113
TC2	0.0178	0.1421	0.0234	0.0154	0.8302	0.0425
TC3	0.0051	0.5494	0.0095	0.9335	0.0740	0.6380
TCVBP1	0.1790	0.0029	0.8263	0.0012	0.1765	0.0030
TCVBP2	0.1942	0.0034	0.4656	0.0004	0.1866	0.0035
TCR	0.2634	0.0028	0.2487	0.0001	0.2586	0.0028
TCQ	0.1890	0.0021	0.1659	0.0002	0.1910	0.0022
TCS	0.2191	0.0027	0.2983	0.0004	0.2358	0.0027
Total	1.2084	0.8759	2.1071	1.1723	2.0855	1.0060

Table 6.12c the IAE Values of Typical (AL4) and Heat-Integrated Plant of HDA process (HIP4) to a Change in the Total Toluene Feed Flowrates

Controller	Integral Absolute Error (IAE)					
	CS1		CS2		CS3	
	AL4	HIP4	AL4	HIP4	AL4	HIP4
TC1	0.1919	0.3296	0.1893	0.3074	0.1910	0.3314
TC2	0.0205	0.0100	0.0227	0.0101	0.1462	0.0027
TC3	0.8276	0.7175	0.8628	0.7520	0.9540	0.7252
TCVBP1	0.0140	0.0023	0.0860	0.0047	0.0135	0.0023
TCVBP2	0.0237	0.0017	0.0702	0.0031	0.0235	0.0017
TCR	0.0238	0.0071	0.0221	0.0050	0.0241	0.0070
TCQ	0.0270	0.0074	0.0252	0.0062	0.0275	0.0074
TCS	0.0086	0.0036	0.0114	0.0046	0.0087	0.0036
Total	1.1371	1.0792	1.2896	1.0931	1.3885	1.0813

CHAPTER VII

CONCLUSIONS AND RECOMMENDATIONS

7.1 Conclusion

In this thesis, we are considered two huge works, Primary work, we design a new sequence of separation part for HDA process, First studied by Luyben, et al. (1998) and further explored by Ploypaisansang (2003). For the new sequence design, three distillation columns are find the sequence for minimize the energy usages and economic evaluation is based on the total annual cost in the separation section. In the new sequence we can save the energy usage 7.2% and the total annual cost (TAC) 16.9% compared with the typical sequence. So we call this concept is heat-integrated process (HIP).

Secondary work, we are considered the heat-integrated processes (HIPs) design altogether with plantwide control structure selection for reduction of energy consumption and maintaining good control performance. We look at 8 alternatives of various heat-integrated processes (HIPs) structure (base on Luyben's and Douglas design, alternative 1, 2, 3 and 4) and our work we design 4 new heat-integrated process (HIP1, 2, 3 and 4) of HDA plant and three new design of plantwide control structures (CS1, CS2 and CS3) to implemented in all alternative structures. In our work we can save the energy usage 5.7-24.7 % from the alternative1.

The plantwide control structures are designed using the disturbance load propagation method and HEN design follows resilient HEN synthesis method (Wongsri, M., 1990) and heat pathway heuristics,(HPH) (Wongsri, M. and Hermawan Y.D., 2005), respectively. In general the HPH is very useful in terms of heat load or disturbance management to achieve the highest possible dynamic MER.

Two kinds of disturbances: thermal and material disturbances are used in evaluation of the plantwide control structures. The performances of the heat integrated plants (HIPs) and the control structures evaluated dynamically by commercial software HYSYS.

Since the major of control loop is similar, the dynamic response and dynamic performance of the three new control structures are slightly deference. The IAE method is used to evaluate the dynamic performance of the designed control systems.

As can be see, The IAE of control structures CS1 close to the IAE of CS3 but IAE of CS2 to high because CS2 exhibited very slow dynamics and is more sensitive to the disturbances. However, CS1 is the best control structure for handle disturbances due to it gives better control performances than CS3 and CS2 respectively.

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

7.2 Recommendations

1. Study and design the highly complex heat-integrated plants (HIPs) of HDA process point of view.

2. Study the controllability characteristics of highly complex heat-integrated plant of HDA process.

3. Study and design the control structure of complex heat-exchanger networks (HENs) and heat-integrated plants (HIPs) of the other process in plantwide control point of view.

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APPENDICES

APPENDIX A

PROCESS STREAM DATA OF HEAT-INTEGRATED PROCESS HDA PLANT

Table A.1 Process Stream Data for Heat-integrated process of HDA Process HIP1

Stream name	**Rtol**	**FFH2**	**FFtot**	**v1out**	**v2out**
Temperature [C]	140.2327	30.0000	30.0000	29.9944	30.1915
Pressure [bar]	39.6449	43.0922	43.0922	39.6449	39.6449
MolarFlow[kgmole/hr]	39.8304	224.2800	132.1595	224.2800	132.1595
H_2,mole fraction	0.0000	0.8025	0.0000	0.8025	0.0000
CH_4	0.0000	0.1975	0.0000	0.1975	0.0000
C_6H_6	0.0008	0.0000	0.0000	0.0000	0.0000
C_7H_8	0.9992	0.0000	1.0000	0.0000	1.0000
$C_{12}H_{10}$	0.0000	0.0000	0.0000	0.0000	0.0000

Stream name	**liq**	**grecycle**	**purge**	**v4out**	**p1out**
Temperature [C]	45.0000	45.0000	45.0000	43.6075	45.2575
Pressure [bar]	31.2918	31.2918	31.2918	24.4074	36.5413
MolarFlow[kgmole/hr]	224.4228	1604.5511	221.4034	221.4034	224.4228
H_2,mole fraction	0.0001	0.0713	0.0713	0.0713	0.0001
CH_4	0.0087	0.8466	0.8466	0.8466	0.0087
C_6H_6	0.6956	0.0720	0.0720	0.0720	0.6956
C_7H_8	0.2639	0.0102	0.0102	0.0102	0.2639
$C_{12}H_{10}$	0.0316	0.0000	0.0000	0.0000	0.0316

Stream name	**d3**	**b3**	**P3out**	**V3out**	**V10out**
Temperature [C]	137.6255	292.8159	140.1496	140.2327	271.5070
Pressure [bar]	2.0684	2.1442	42.8534	39.6449	1.4180
MolarFlow[kgmole/hr]	39.8304	2.8516	39.8304	39.8304	2.8516
H_2,mole fraction	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000
C_6H_6	0.0008	0.0000	0.0008	0.0008	0.0000
C_7H_8	0.9992	0.0002	0.9992	0.9992	0.0002
$C_{12}H_{10}$	0.0000	0.9998	0.0000	0.0000	0.9998

Stream name	**V7out**	**d1**	**V6out**	**quench**	**cHEout**
Temperature [C]	54.5081	154.7871	153.7191	45.3982	580.0003
Pressure [bar]	2.6936	6.8400	6.1412	33.4989	37.5695
MolarFlow[kgmole/hr]	9.0659	132.2172	132.2172	49.5230	2000.8213
H_2,mole fraction	0.0086	0.0002	0.0002	0.0001	0.0501
CH_4	0.6704	0.0124	0.0124	0.0087	0.4462
C_6H_6	0.3210	0.9870	0.9870	0.6956	0.0377
C_7H_8	0.0000	0.0005	0.0005	0.2639	0.4660
$C_{12}H_{10}$	0.0000	0.0000	0.0000	0.0316	0.0000

Table A.1 Continued

Stream name	bp1out	cHE1out	tot	v11out	cHE1in
Temperature [C]	63.1408	603.8906	621.1698	45.3982	63.4554
Pressure [bar]	37.5695	37.5695	33.4989	33.4989	39.6449
MolarFlow[kgmole/hr]	97.5531	1903.2682	2050.3764	49.5230	1903.2679
H_2, mole fraction	0.0501	0.0501	0.0381	0.0001	0.0501
CH_4	0.4462	0.4462	0.4564	0.0087	0.4462
C_6H_6	0.0377	0.0377	0.3624	0.6956	0.0377
C_7H_8	0.4660	0.4660	0.1283	0.2639	0.4660
$C_{12}H_{10}$	0.0000	0.0000	0.0147	0.0316	0.0000

Stream name	Rgas	cHEin	Rout	conden	gas
Temperature [C]	70.1096	63.4554	665.8514	45.0001	45.0000
Pressure [bar]	39.6449	39.6449	33.4989	31.2918	31.2918
MolarFlow[kgmole/hr]	1604.5511	2000.8211	2000.8534	2050.3774	1825.9545
H_2, mole fraction	0.0713	0.0501	0.0425	0.0381	0.0713
CH_4	0.8466	0.4462	0.5077	0.4564	0.8466
C_6H_6	0.0720	0.0377	0.3242	0.3624	0.0720
C_7H_8	0.0102	0.4660	0.1128	0.1283	0.0102
$C_{12}H_{10}$	0.0000	0.0000	0.0128	0.0147	0.0000

Stream name	Toltot	toquench	toC1	dischg	V9out
Temperature [C]	58.3730	45.2575	45.2575	70.1096	143.4763
Pressure [bar]	39.6449	36.5413	36.5413	39.6449	2.1144
MolarFlow[kgmole/hr]	171.9899	49.5230	174.8998	1604.5511	42.6818
H_2, mole fraction	0.0000	0.0001	0.0001	0.0713	0.0000
CH_4	0.0000	0.0087	0.0087	0.8466	0.0000
C_6H_6	0.0002	0.6956	0.6956	0.0720	0.0007
C_7H_8	0.9998	0.2639	0.2639	0.0102	0.8923
$C_{12}H_{10}$	0.0000	0.0316	0.0316	0.0000	0.1070

Stream name	b1	V5out	d2	b2	V8out
Temperature [C]	200.5129	45.8101	56.2455	153.1833	90.6646
Pressure [bar]	7.0404	6.9269	6.1410	6.1516	1.3790
MolarFlow[kgmole/hr]	42.6818	174.8998	9.0659	123.1519	123.1519
H_2, mole fraction	0.0000	0.0001	0.0086	0.0000	0.0000
CH_4	0.0000	0.0087	0.6704	0.0000	0.0000
C_6H_6	0.0007	0.6956	0.3210	0.9995	0.9995
C_7H_8	0.8923	0.2639	0.0000	0.0005	0.0005
$C_{12}H_{10}$	0.1070	0.0316	0.0000	0.0000	0.0000

Stream name	hHEout	hHEin	Rin	bp1
Temperature [C]	121.6053	621.1699	621.1113	63.4554
Pressure [bar]	31.4298	33.4989	34.6719	39.6449
MolarFlow[kgmole/hr]	2050.3774	2050.3764	2000.8213	97.5531
H_2, mole fraction	0.0381	0.0381	0.0501	0.0501
CH_4	0.4564	0.4564	0.4462	0.4462
C_6H_6	0.3624	0.3624	0.0377	0.0377
C_7H_8	0.1283	0.1283	0.4660	0.4660
$C_{12}H_{10}$	0.0147	0.0147	0.0000	0.0000

APPENDIX B

EQUIPMENT AND DATA SPECIFICATION OF HEAT-INTEGRATED PROCESS OF HDA PROCESS

Table B.1 Equipment data and Specifications of heat-integrated plant of HDA process

Equipments	Specifications	Heat-integrated process of HDA process			
		HIP1	HIP2	HIP3	HIP4
Reactor	Diameter (m)	17.374	17.374	17.374	17.374
	Length (m)	2.905	2.905	2.905	2.905
	Number of tube	1	1	1	1
Furnace	Tube volume (m^3)	8.5	8.5	8.5	8.5
Cooler	Tube volume (m^3)	8.5	8.5	8.5	8.5
Separator	Liquid volume (m^3)	1.13	1.13	1.13	1.13
FEHE1	Shell volume (m^3)	14.16	14.16	14.16	14.16
	Tube volume (m^3)	14.16	14.16	14.16	14.16
	UA (kJ/C-h)	1.9×10^6	2.29×10^6	7.82×10^5	4.41×10^5
FEHE2	Shell volume (m^3)			14.16	14.16
	Tube volume (m^3)	-	-	14.16	14.16
	UA (kJ/C-h)			1.47×10^5	2.09×10^4
Reboiler Column 1 (RC1)	Shell volume (m^3)				14.16
	Tube volume (m^3)	-	-	-	14.16
	UA (kJ/C-h)				1.89×10^5
Reboiler Column 2 (RC2)	Shell volume (m^3)		14.16		
	Tube volume (m^3)	-	14.16	-	-
	UA (kJ/C-h)		1370.97		
Reboiler Column 3 (RC3)	Shell volume (m^3)			14.16	
	Tube volume (m^3)	-	-	14.16	-
	UA (kJ/C-h)			24945.57	
Tank Bottom C1(TB1)**	Vessel volume (m^3)		-	-	5.6174
Tank Bottom C2(TB2)**	Vessel volume (m^3)	-	1.6308	-	-
Tank Bottom C3(TB3)**	Vessel volume (m^3)		-	0.4439	-

Table B.2 Column Specifications of the typical of HDA process Alternative 1 compared with the Heat-integrated plant of HDA process HIP1

Parameters	The typical of HDA process Alternative.1			Heat-integrated process of HDA process HIP1		
	Stabilizer Column	Product Column	Recycle Column	Stabilizer Column	Product Column	Recycle Column
F	174.90	166.21	42.44	175.15	132.66	42.49
D	8.69	123.77	39.56	132.66	9.06	39.60
B	166.21	42.44	2.88	42.49	123.60	2.89
Tdi	51.05	93.33	133.33	154.79	55.56	137.62
Tbi	189.64	137.78	260.00	200.67	153.26	291.92
Pdi	10.34	2.068	2.068	6.84	6.14	2.07
Pbi	10.38	2.221	2.108	7.05	6.16	2.11
Rm reflux rate	14.76	42.44	10.14	360.07	98.23	8.76
R/Rm reflux ratio	1.70	3.04	0.26	2.71	10.84	0.22
Nt	6	27	7	36	3	7
Nrt	3	15	5	20	3	4
Qci (kw)	176.72	4061.26	437.40	2664.23	1097.69	425.58
Qri (kw)	1273.01	3461.76	481.49	4494.11	175.97	321.34
Total Qc (kw)		4675.38			4187.51	
Total Qr (kw)		5216.26			4991.42	
Condenser						
Diameter	0.62	1.19	1.34	2.22	3.39	3.76
Length	0.93	1.79	2.01	3.34	5.08	5.64
Volumn	0.28	2.00	2.83	12.95	45.86	62.71
Reboiler						
Diameter	1.82	1.19	1.06	5.43	1.18	0.81
Length	2.73	1.79	1.59	8.15	1.76	1.21
Volumn	7.08	2.00	1.42	188.94	1.91	0.62
Specification of heavy / light key						
mole fac. In dis	0.042000	0.0003	0.00002	0.00034	0.08000	0.00002
mole fac. In bot.	0.000001	0.0006	0.00026	0.00090	0.000001	0.00026
Tray sections						
Diameter	1.07	5.16	0.76	5.16	1.50	0.76
tray/pack space	0.61	0.61	0.61	0.61	0.55	0.61
tray/pack volumn	0.54	12.75	0.28	12.75	0.97	0.28
weir H	0.05	0.05	0.05	0.05	0.05	0.05
weir L	0.85	4.13	0.61	4.13	1.20	0.61
DC volumn	0.09	0.09	0.09	0.09	0.09	0.09

Table B.3 Column Specifications for calculate cost of distillation using by CAPCOST; Alternative.1

Parameters columns	The typical of HDA process Alternative.1		
	Stabilizer Column	Product Column	Recycle Column
Materials	Carbon Steel	Carbon Steel	Carbon Steel
Materials of Construction	Starinless Steel	Starinless Steel	Starinless Steel
Tray	Sieve Tray	Sieve Tray	Sieve Tray
Tray MOC	Carbon Steel	Carbon Steel	Carbon Steel
Number of tray	6.00	27.00	7.00
Diameter of Vessel (m)	1.067	5.161	0.762
Tray space (m)	0.6096	0.6096	0.6096
Height of Vessel (m)	3.66	16.46	4.27
Maximum Pressure (barg)	9.38	1.221	1.108
Condenser (Utility Cooler)			
Pressure shell or tube (barg)	9.34	1.068	1.068
Heat transfer area (m2)	60.32	60.32	60.32
Type of Exchanger	Multiple Pipe	Multiple Pipe	Multiple Pipe
Materials shell or tube	CS/CS	CS/CS	CS/CS
Number of shells	1.00	1.00	1.00
Duty of Condenser (Qci, MJ/h)	633.60	14590.00	1575.00
Range of Tempareture (C)	51.05	93.33	133.30
Range of Pressure (barg)	9.34	1.068	1.068
Reboiler (Utility Heater)			
Pressure shell or tube (barg)	9.38	1.221	1.108
Heat transfer area (m2)	60.32	60.32	60.32
Type of Exchanger	Multiple Pipe	Multiple Pipe	Multiple Pipe
Materials shell or tube	CS/CS	CS/CS	CS/CS
Number of shells	1.00	1.00	1.00
Duty of Reboiler (Qri, MJ/h)	4583.00	12460.00	1733.00
Range of Tempareture (C)	189.60	137.80	260.00
Range of Pressure (barg)	9.38	1.221	1.108
Total Duty of Condenser (Qc, MJ/h)		16798.60	
Total Duty of Reboiler (Qr, MJ/h)		18776.00	
Total Duty (Q, MJ/h)		35574.60	

Table B.4 Column Specifications for calculate cost of distillation using by CAPCOST; HIP1

Parameters Columns	Heat-integrated process of HDA process HIP1		
	Stabilizer Column	Product Column	Recycle Column
Materials	Carbon Steel	Carbon Steel	Carbon Steel
Materials of Construction	Starinless Steel	Starinless Steel	Starinless Steel
Tray	Sieve Tray	Sieve Tray	Sieve Tray
Tray MOC	Carbon Steel	Carbon Steel	Carbon Steel
Number of tray	36.00	3.00	7.00
Diameter of Vessel (m)	5.161	1.500	0.762
Tray space (m)	0.6096	0.5500	0.6096
Height of Vessel (m)	21.95	1.65	4.27
Maximum Pressure (barg)	6.05	5.161	1.108
Condenser (Utility Cooler)			
Pressure shell or tube (barg)	5.84	5.141	1.068
Heat transfer area (m2)	60.32	60.32	60.32
Type of Exchanger	Multiple Pipe	Multiple Pipe	Multiple Pipe
Materials shell or tube	CS/CS	CS/CS	CS/CS
Number of shells	1.00	1.00	1.00
Duty of Condenser (Qci, MJ/h)	9591.00	3952.00	1532.00
Range of Tempareture (C)	154.80	55.56	137.60
Range of Pressure (barg)	5.84	5.141	1.068
Reboiler (Utility Heater)			
Pressure shell or tube (barg)	6.05	5.161	1.108
Heat transfer area (m2)	60.32	60.32	60.32
Type of Exchanger	Multiple Pipe	Multiple Pipe	Multiple Pipe
Materials shell or tube	CS/CS	CS/CS	CS/CS
Number of shells	1.00	1.00	1.00
Duty of Reboiler (Qri, MJ/h)	16180.00	633.50	1157.00
Range of Tempareture (C)	200.70	153.30	291.90
Range of Pressure (barg)	6.05	5.161	1.108
Total Duty of Condenser (Qc, MJ/h)		15075.00	
Total Duty of Reboiler (Qr, MJ/h)		17970.50	
Total Duty (Q, MJ/h)		33045.50	

APPENDIX C

IMPLEMENTATION OF HEAT PATHWAY MANIPULATOR IN HDA PROCESS HIP 1

Figure 6.7.1 shows the plantwide control structure of HDA process HIP 1. The major loops in HDA process HIP 1 are the same as those used in Luyben et al. (1999), except for the outlet temperature control in FEHE and the tray temperature control in the recycle column (C3).

Based on the heat pathway heuristics for plantwide control, a selective controller with low selector switch (LSS) for FEHE is now employed in the current study to select an appropriate heat pathway (Fig. 6.7.1). This control system involves one manipulated variable and two controlled variables and works as follows: The hot outlet temperature of FEHE is controlled at its nominal set point by manipulating the valve on the bypass line (VBP1). At the same time, the cold outlet temperature of FEHE 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 FEHE drops below the allowable limit due to, for example, a disturbance load entering the process, the LSS switches the control action from the hot temperature control (ThE1) to the cold temperature control (TcE1), and closes the valve VBP1.

As a result, the cold outlet temperature of FEHE will rise to its normal temperature and the hot outlet temperature of FEHE will be further decreased, so the cooler duty will also be decreased. Whenever the cold outlet temperature of FEHE 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 from TcE1 to ThE1.

Consequently, the hot outlet temperature of FEHE will drop to its normal temperature and the cold outlet temperature of FEHE will be further increased, so the furnace duty will also be decreased. In order to apply this LSS control strategy in HYSYS, a selector block can be added to the process flow diagram. On the connections page tab, all input signals are specified. On the parameter page tab, the mode of selector is set; in this case the minimum of all input should be selected.

Since the temperature profile in the recycle column is very sharp because of the large difference in boiling point between toluene and diphenyl, this produces large 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 Figure 6.7.1) 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.

6.7.1 Control Structure 1 for the heat-integrated plant of HDA Process HIP 1

This control structure is shown in Figure 6.7.1. This control structure, the bypass of feed effluent heat exchangers (FEHE) is on cold side. Figure 6.7.1 shows the plantwide control structure for the heat-integrated plant of HDA process HIP 1.

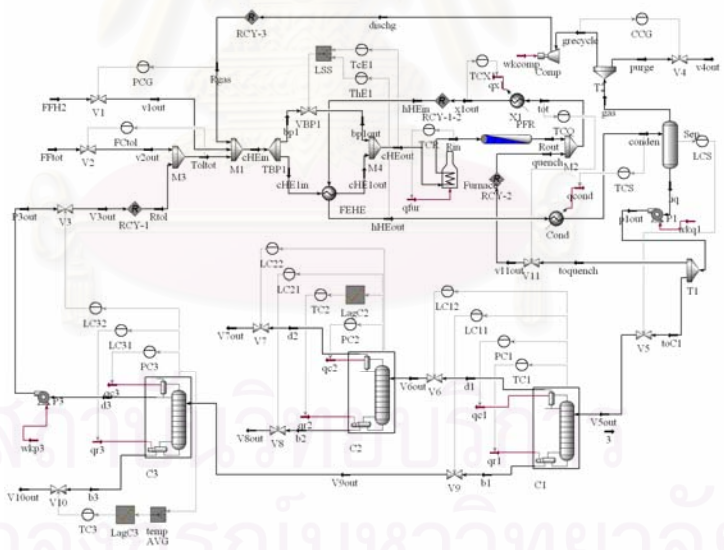

Figure 6.7.1 The plantwide control structure CS1 of the HDA process HIP 1

6.7.2 Control Structure 2 for the heat-integrated plant of HDA Process HIP 1

This control structure is shown in Figure 6.7.2. The major loops in this control structure are the same as CS1 except for control loop for FEHE. The all bypass of FEHE will be on hot side.

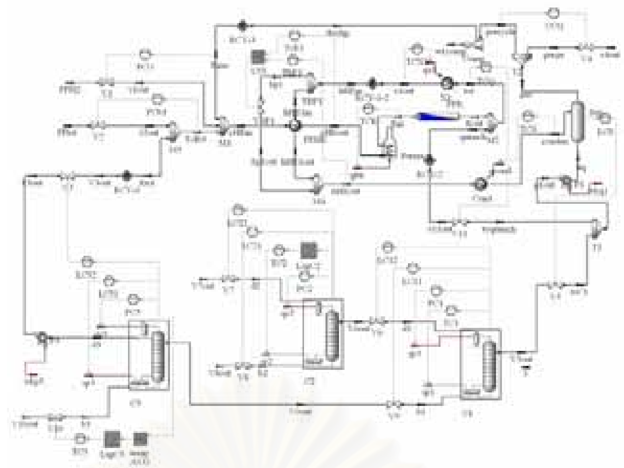

Figure 6.7.2 The plantwide control structure CS3 of the HDA process HIP 1

6.7.3 Control Structure 3 for the heat-integrated plant of HDA Process HIP 1

This control structure is shown in Figure 6.7.3. 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 bottom stage temperature and tray 2 temperature controls.

6.7.4 Dynamic Simulation Results

In order to illustrate the dynamic behaviors of the new control structure with LSS in HDA process HIP 1, several disturbance loads were made. Figures 6.7.4 to 6.7.12 show the dynamic responses of the control structures (CS1, CS2 and CS3) with LSS for the HDA process HIP 1. Results for individual disturbance load changes are as follows:

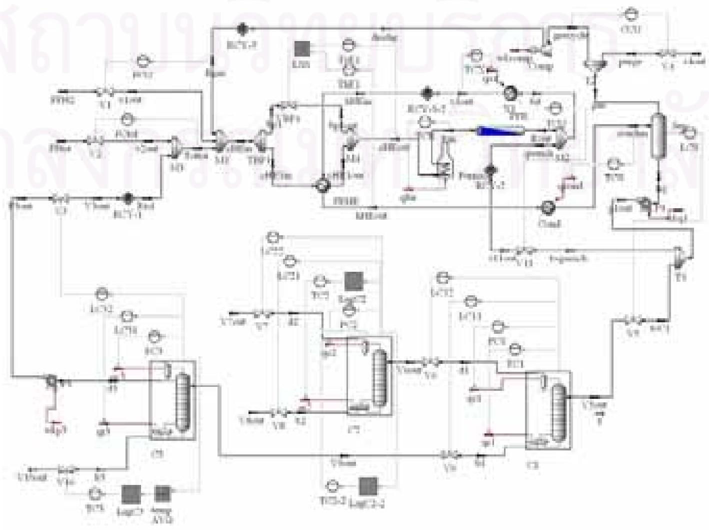

Figure 6.7.3 The plantwide control structure CS3 of the HDA process HIP 1

6.7.5 Change in the Disturbance Load of Cold Stream (Reactor Feed Stream)

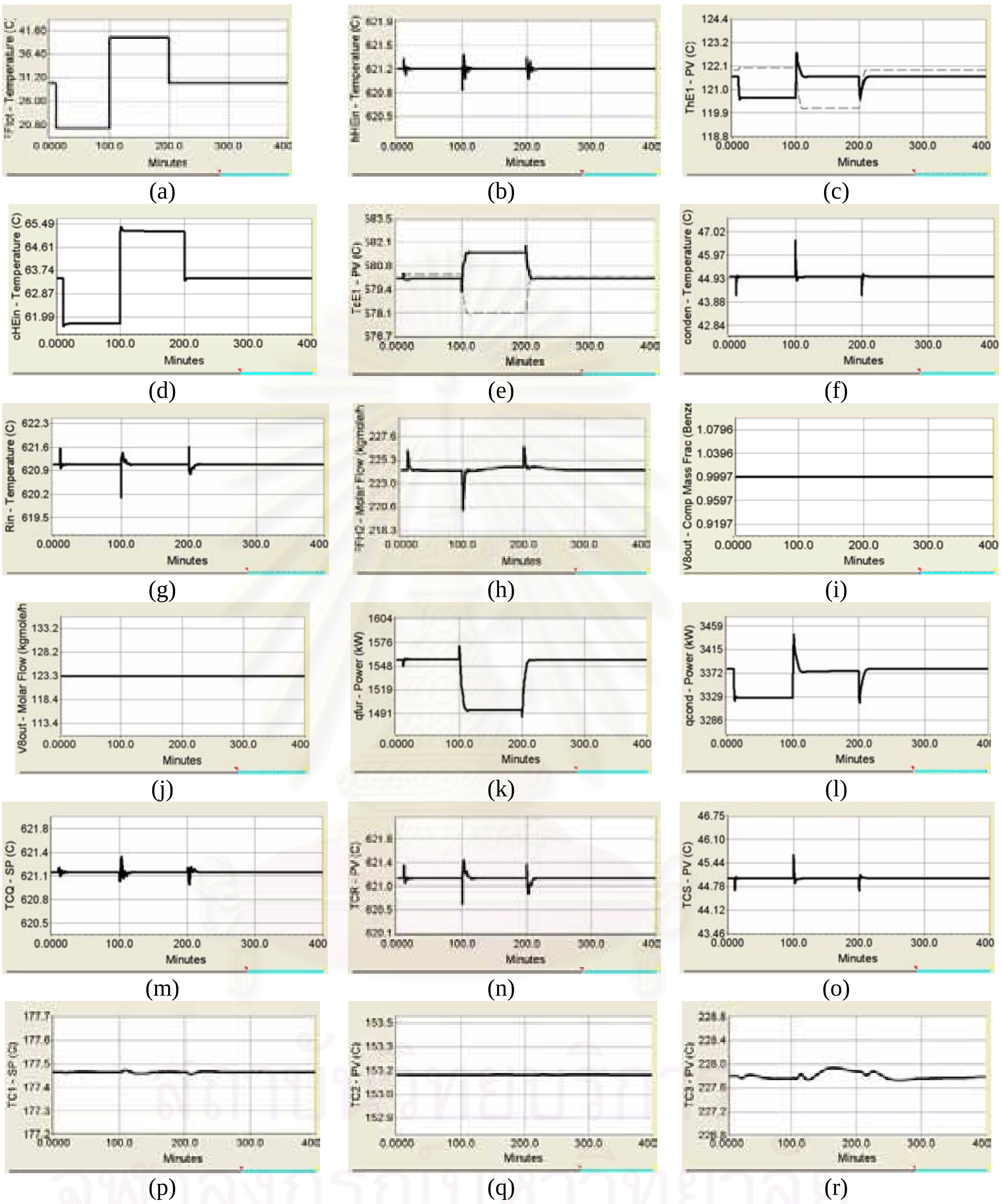

Figure 6.7.4 Dynamic Responses of the HDA Process of HIP 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

(Note ▬▬ Process variable (PV), ------ Manipulated variable)

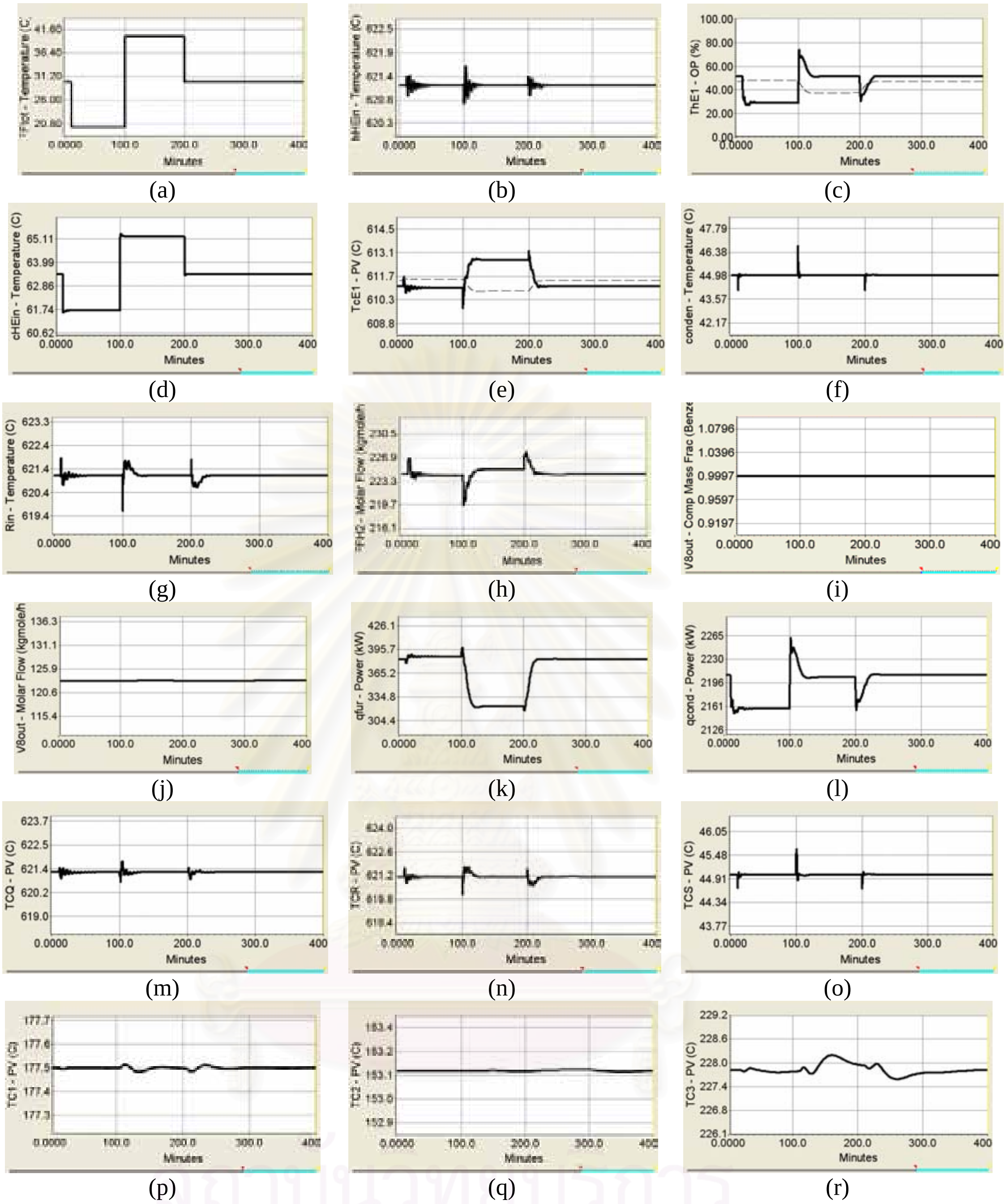

Figure 6.7.5 Dynamic Responses of the HDA Process of HIP 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

(Note ▬▬ Process variable (PV), ------ Manipulated variable)

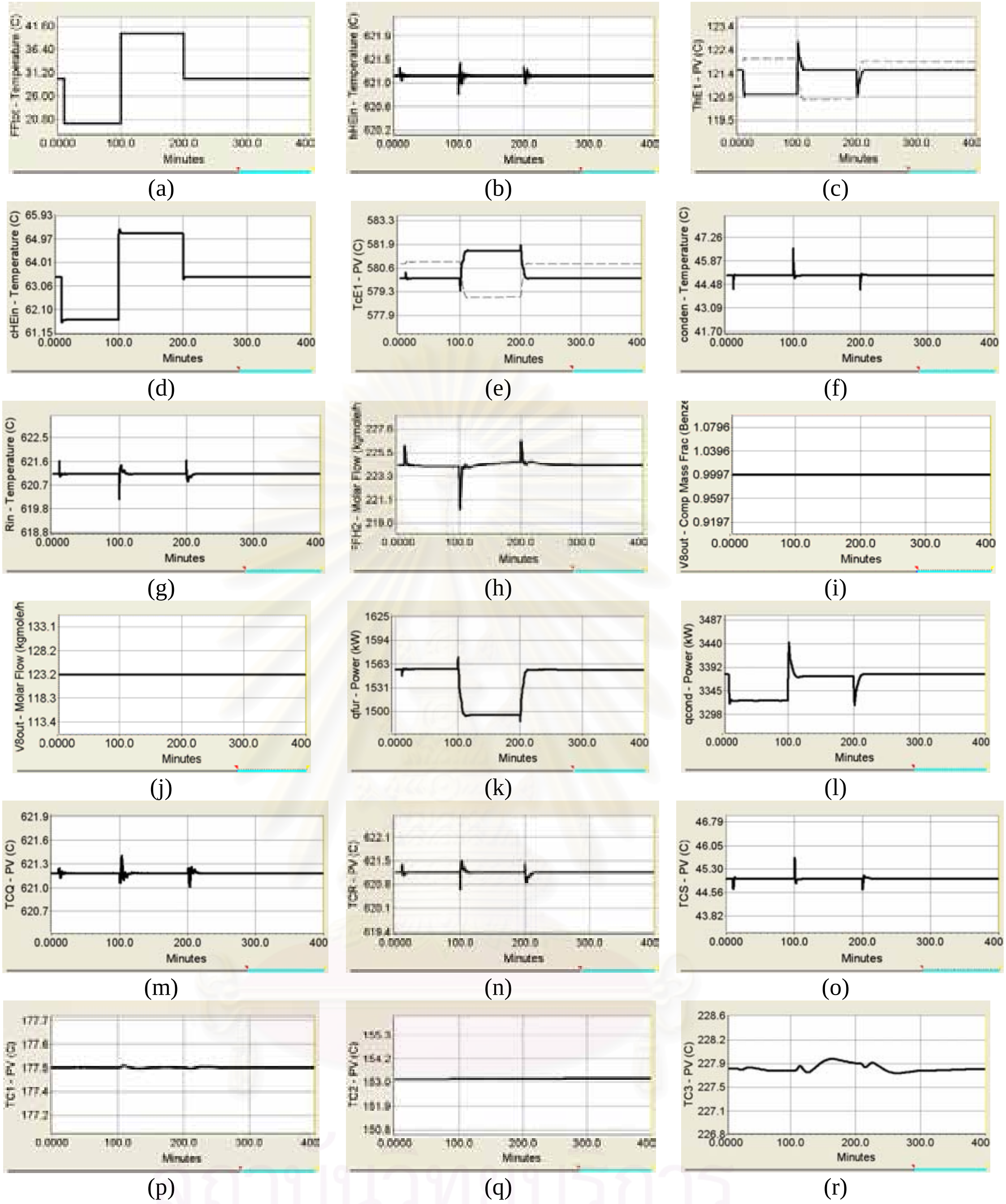

Figure 6.7.6 Dynamic Responses of the HDA Process of HIP 1 to A Change in the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) Fresh feed toluene temperature, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

(Note ▬▬ Process variable (PV), ------ Manipulated variable)

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

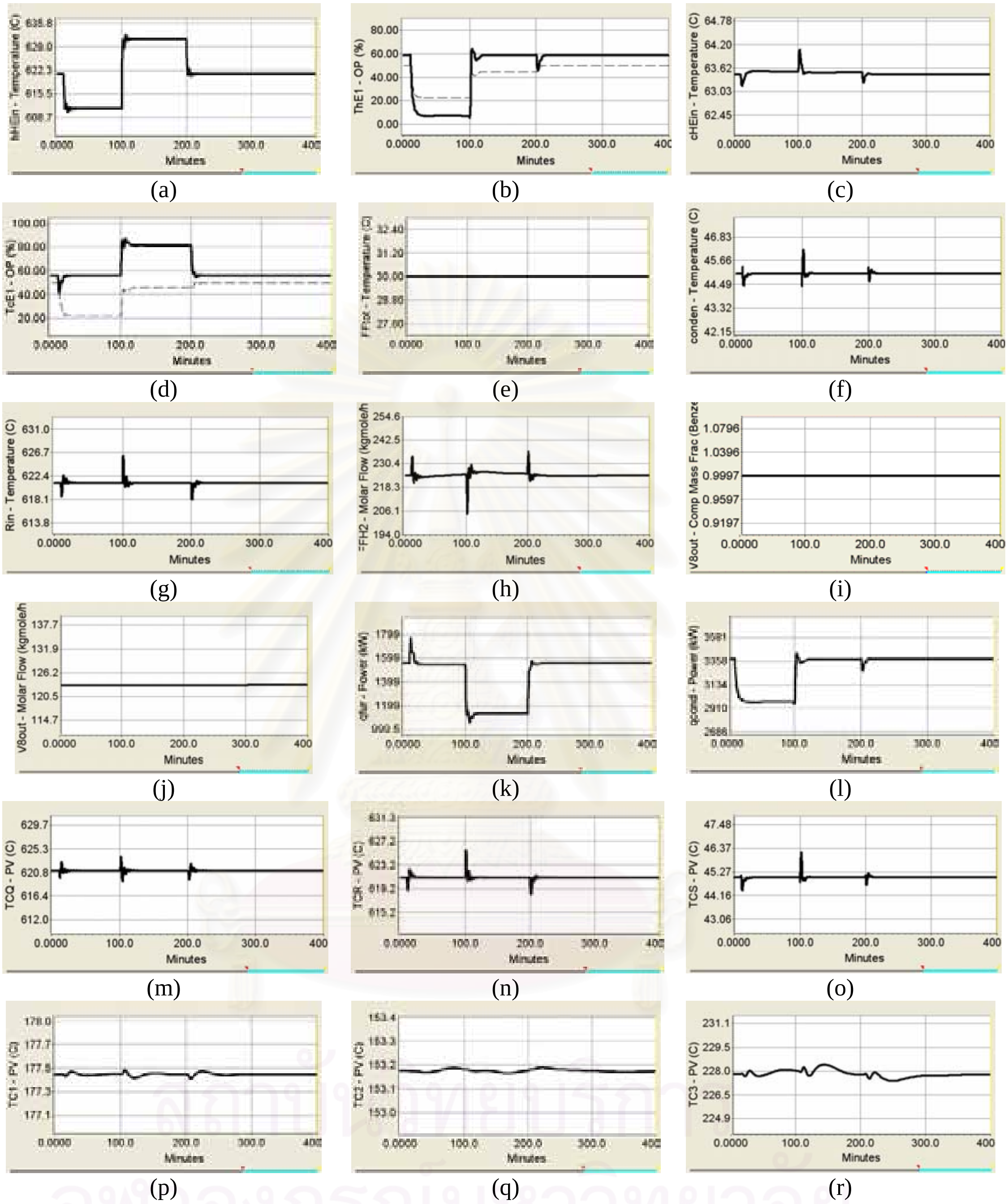

Figure 6.7.7 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS1, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

(Note ▬▬ *Process variable (PV),* ------ *Manipulated variable)*

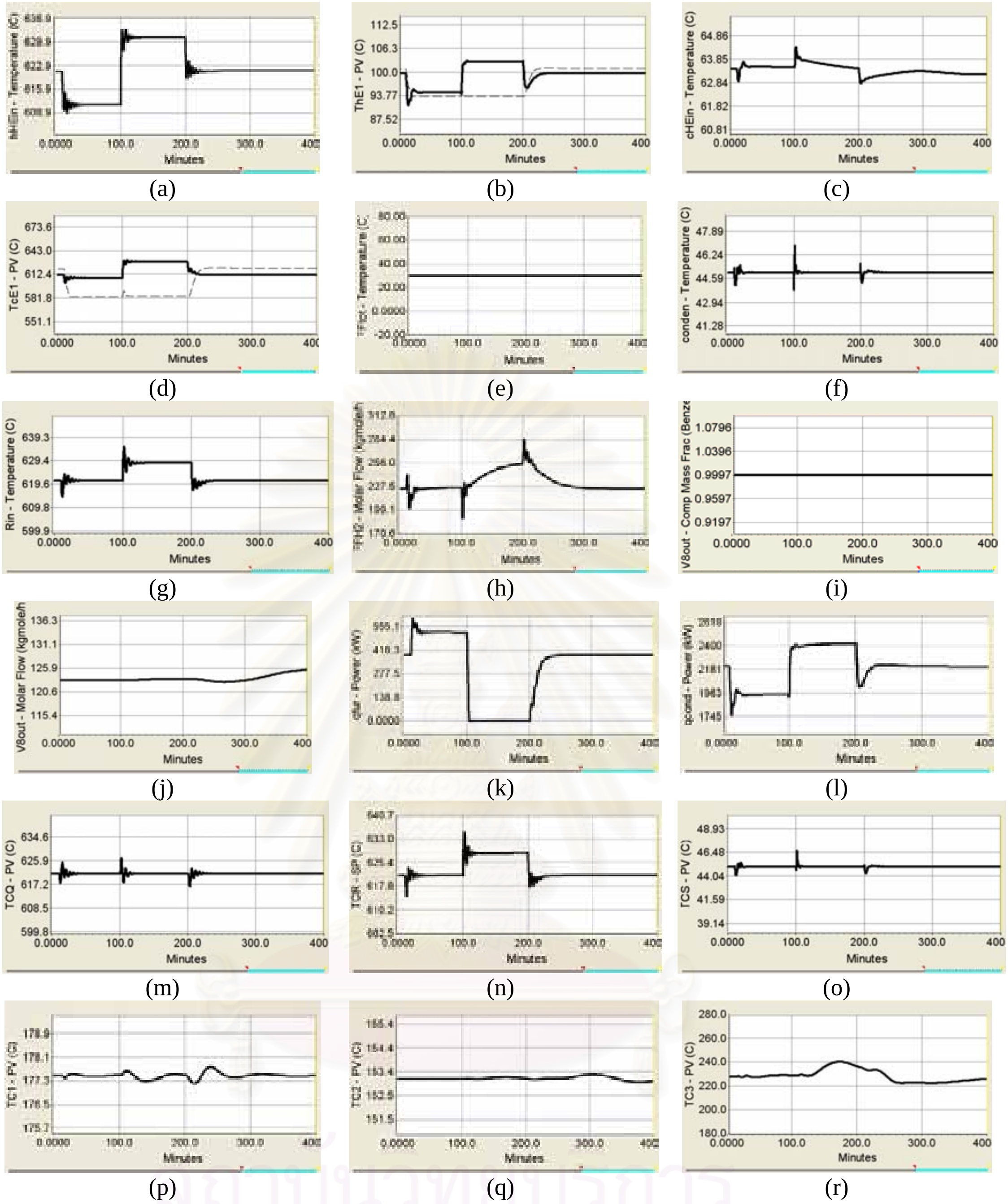

Figure 6.7.8 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS2, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

(Note —— Process variable (PV), ------ Manipulated variable)

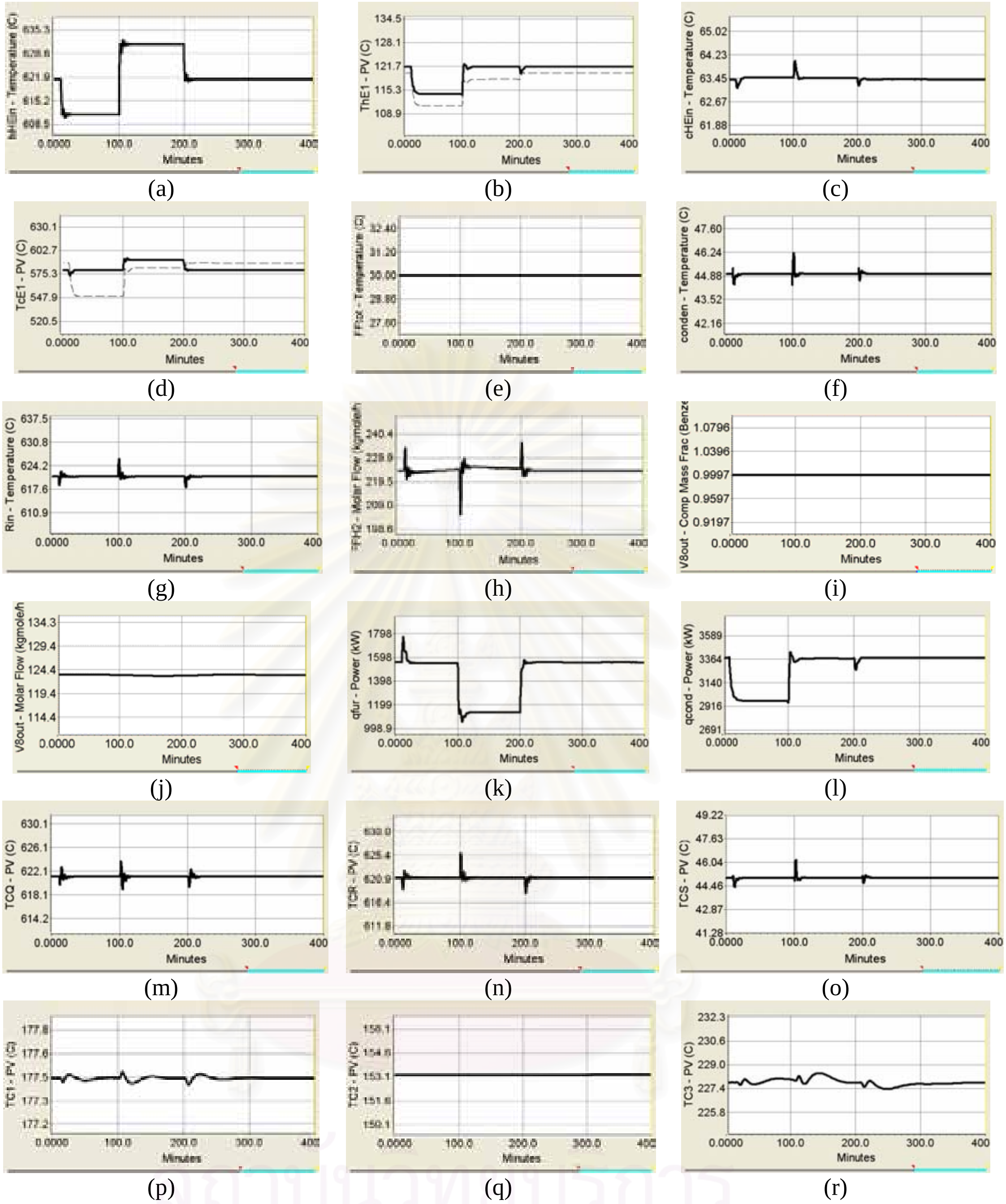

Figure 6.7.9 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Heat Load Disturbance of Hot Stream (Reactor Product Stream):CS3, where: (a) FEHE hot inlet temperature, (b) FEHE hot outlet temperature, (c) FEHE cold inlet temperature, (d) FEHE cold outlet temperature, (e) Fresh feed toluene temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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.

(Note ▬▬ *Process variable (PV),* ------ *Manipulated variable)*

6.7.4 Change in the Total Toluene Feed Flowrates

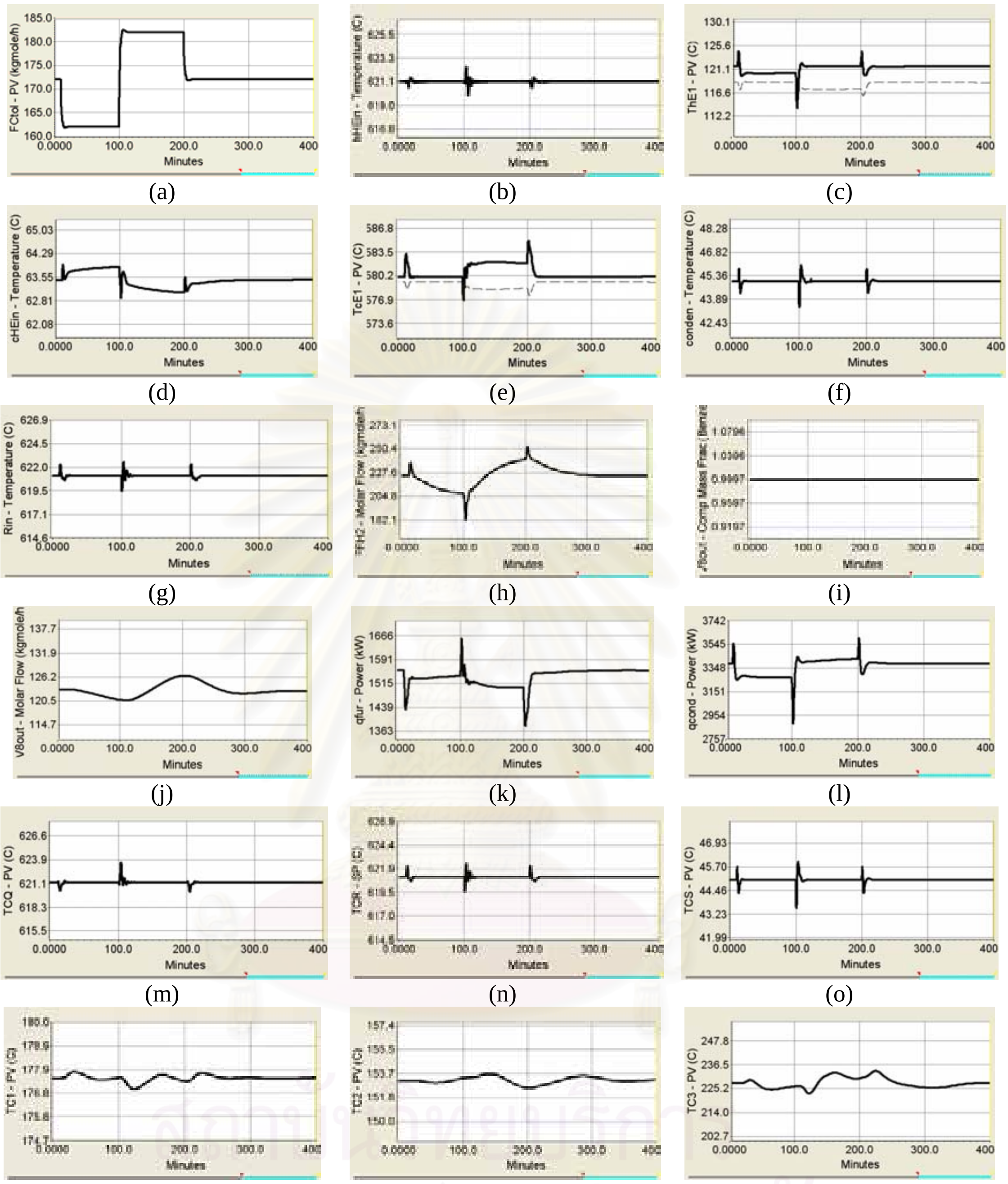

Figure 6.7.10 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Total Toluene Feed Flowrates:CS1, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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. *(Note* ▬▬ *Process variable (PV),* ------ *Manipulated variable)*

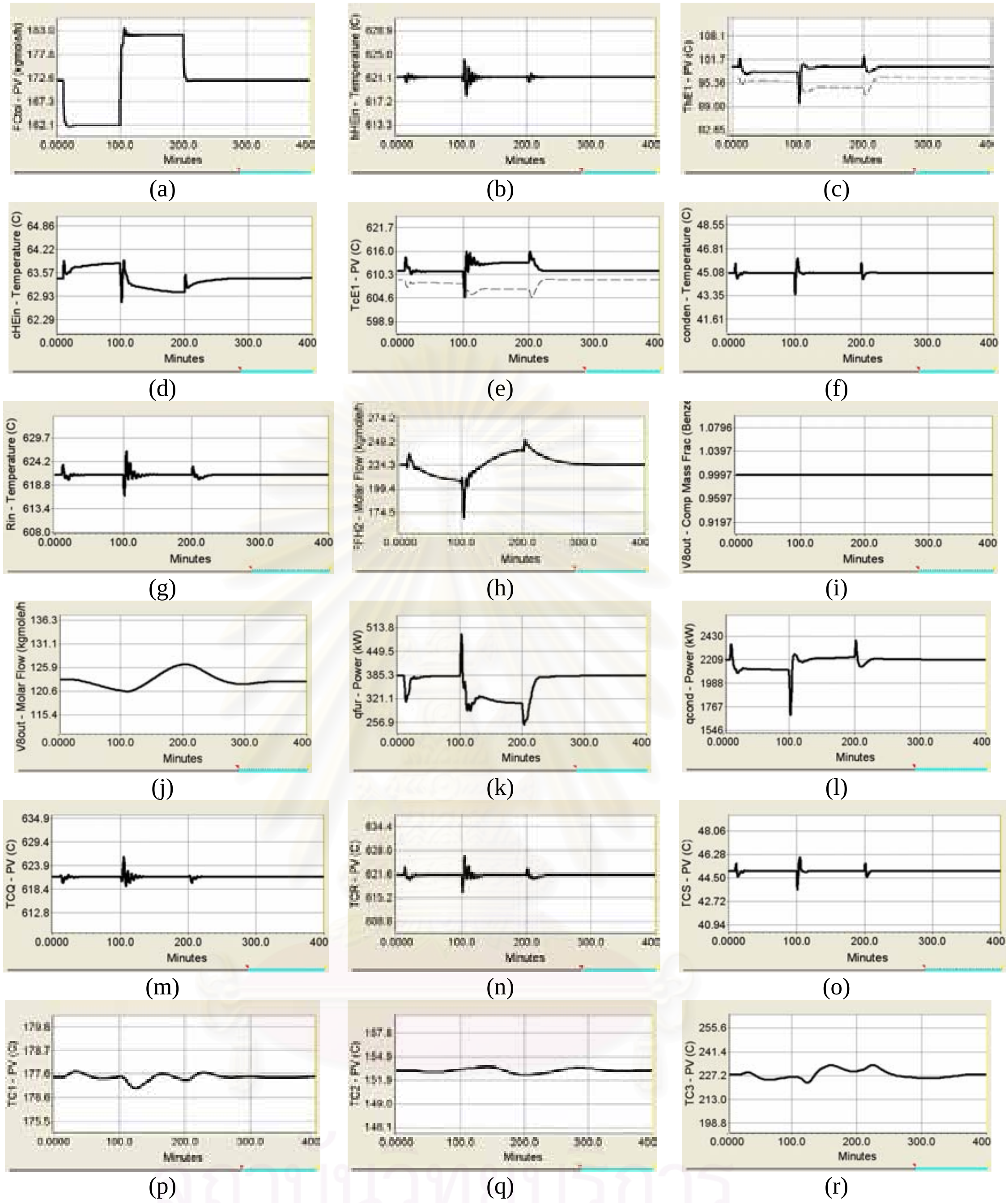

Figure 6.7.11 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Total Toluene Feed Flowrates:CS2, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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. *(Note* ▬▬ *Process variable (PV),* ------ *Manipulated variable)*

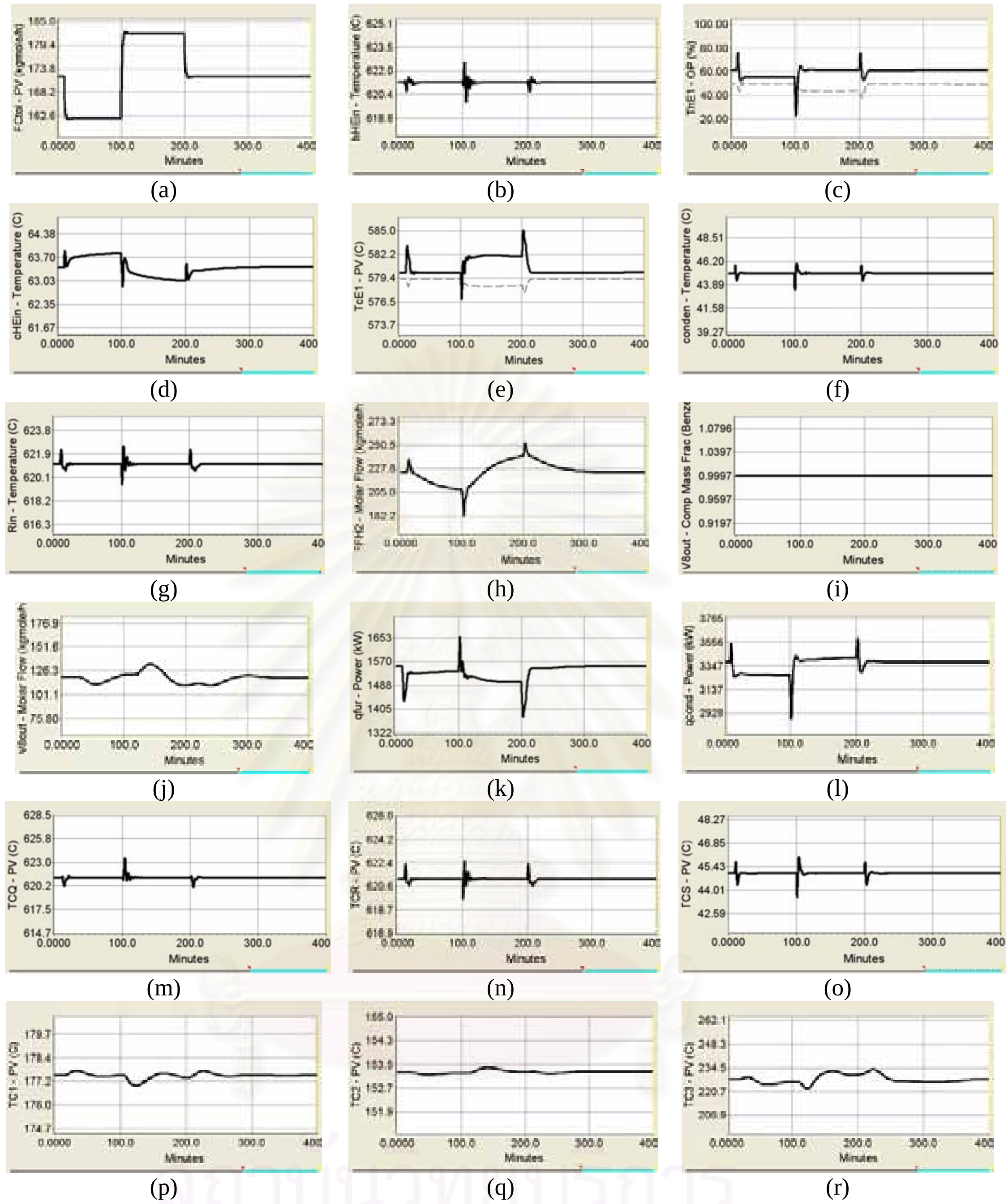

Figure 6.7.12 Dynamic Responses of the HDA Process of HIP 1 to a Change in the Total Toluene Feed Flowrates:CS3, where: (a) total toluene feed flowrates, (b) FEHE hot inlet temperature, (c) FEHE hot outlet temperature, (d) FEHE cold inlet temperature, (e) FEHE cold outlet temperature, (f) cooler outlet temperature, (g) furnace outlet temperature, (h) fresh feed hydrogen flowrates, (i) benzene purity, (j) benzene flowrate, (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. *(Note* ▬▬ *Process variable (PV),* ------ *Manipulated variable)*

APPENDIX D

SIZINZ AND COSTING OF DISTILLATION COLUMNS

D.1 Sizing and costing of distillation columns

For a given number of theoretical trays, HYSYS simulator calculates column diameter after converging for selected valve tray distillation column with 50.8-mm weir height. Valve trays of Glitsch type are considered. In order to estimate the actual number of trays, overall column effciency is calculated using simplied equation [4]

$$\log(E_o) = 1.67 - 0.25\ \log(\mu_{avg}\alpha_{avg}) + 0.30\ \log(L_m/V_m) + 0.30(h_l) \qquad (D.1)$$

$$N_{actual} = N_{theoretical}/E_o \qquad (D.2)$$

The height of the column for 0.6 m tray spacing and 6 m disengagement is given by

$$H = (N_{actual} - 1) \times 0.6 + 6.0 \qquad (D.3)$$

Where (Nactual – 1) x 0.6 = tray stack height (m). The costing of distillation columns (carbon steel construction) can be estimated by the following cost equations that are updated from mid-1968 to 1997 using the ratio of Marshall & Swift index (1056.8/280).

$$\text{Installed cost of the column shell,}\quad \$ = (\text{M\&S}/280)(937.61)D^{1.066}H^{0.802}(3.18) \qquad (D.4)$$

If the design pressure (P) is more than 345 kPa, a correction factor is applied.

$$[1 + 1.45 \times 10^{-4}(P - 345)]$$

$$\text{Installed cost of column,}\quad \$ = (\text{M\&S}/280)(136.14)D^{1.55}h \qquad (D.5)$$

$$\text{Total column cost} = \text{Installed cost of column shell} + \text{Installed cost of column trays} \qquad (D.6)$$

D.2 Annual capital cost

The capital cost (purchase plus installation cost) is annualized over a period which is often referred to as plant life time

TAC = Annual operating cost + Annual capital cost

Operating costs were assumed just utility cost (steam and cooling water).

Annual capital cost = Capital cost / Plant life time*

*Plant lifetime 10 years and Operating hours 8000 h/year

VITA

Mr. Kasin Pronpitakthum was born in Bangkok, Thailand on January 21, 1983. He graduated Bachelor Degree in department of Chemical Engineering from Prince of Songkla University in 2005. After that he entered the Graduate School of Chulalongkorn University to pursue the Master of Engineering in Chemical Engineering and completed in 2008.