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และโครงสร้างการควบคุมของโรงงานไดเมทิลอีเทอร์

นายเลอศักดิ์ บุญณะชัย

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คณะวิศวกรรมศาสตร์ จุฬาลงกรณ์มหาวิทยาลัย

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PERFORMANCE EVALUATION OF HEAT INTEGRATION AND CONTROL STRUCTURES OF DIMETHYL ETHER PLANT

Mr.Lersak Bunnachai

A Thesis Submitted in Partial Fulfillment of the Requirements
for the Degree of Master of Engineering Program in Chemical Engineering
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By	Mr.Lersak Bunnachai
Field of study	Chemical Engineering
Advisor	Assistant Professor Montree Wongsri, D.Sc.

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

.... Dean of the Faculty of Engineering

(Associate Professor Boonsom Lerdhirunwong, Dr.Ing.)

THESIS COMMITTEE

.... Chairman

(Associate Professor Tharathon Mongkhonsi, Ph.D.)

.... Advisor

(Assistant Professor Montree Wongsri, D.Sc.)

.... Examiner

(Soorathep Kheawhom, Ph.D.)

.... External Examiner

(Phisit Jaisathaporn, Ph.D.)

เลอศักดิ์ บุญณะชัย : การประเมินสมรรถนะโครงสร้างการเบ็ดเสร็จพลังงานและโครงสร้างการควบคุมของโรงงานไดเมทิลอีเทอร์. (PERFORMANCE EVALUATION OF HEAT INTEGRATION AND CONTROL STRUCTURES OF DIMETHYL ETHER PLANT) อ.ที่ปรึกษาวิทยานิพนธ์หลัก: ผศ. ดร. มนตรี วงศ์ศรี, 172 หน้า.

ไดเมทิลอีเทอร์เป็นพลังงานที่สะอาดและยังเป็นทางเลือกหนึ่งทางด้านเศรษฐ์กิจซึ่งสามารถผลิตได้จากก๊าซธรรมชาติผ่านกระบวนการสังเคราะห์ โดยคุณสมบัติของไดเมทิลอีเทอร์มีความใกล้เคียงกับก๊าซหุงต้มทำให้มีความสามารถในการนำไป ใช้งานหลายด้าน เช่นการผลิตไฟฟ้า หรือทางด้านการขนส่ง ซึ่งในการวิจัยนี้ได้นำเสนอการเบ็ดเสร็จด้านพลังงานสำหรับกระบวนการผลิตไดเมทิลอีเทอร์สามทางเลือกโดยใช้หลักการออกแบบการเบ็ดเสร็จด้านหลังงานและการส่งผ่านความแปรปรวนตามหลักการของ Wongsri (1990) และสามโครงสร้างการควบคุมใหม่ซึ่งทำการออกแบบโครงสร้างการควบคุมกระบวนการตามหลักการออกแบบของ Lyben (1998) และหลักการออกแบบแบบฟิกเจอร์พอล์ท โดยนำมาเปรียบเทียบกับกระบวนการผลิตตัวอย่างเพื่อหาทางเลือกที่มีการใช้พลังงานที่น้อยที่สุดและมีความสามารถในการควบคุมมากที่สุด จากผลการทดลอง พบว่าการเบ็ดเสร็จด้านพลังงานทางเลือกที่ 2 เหมาะสมที่สุด เพราะ สามารถประหยัดพลังงานได้ถึง 58.8% และโครงสร้างการควบคุมแบบ 3 มีความสามารถในการควบคุมมากที่สุดเพราะ ให้ค่า IAE น้อยที่สุด ซึ่งแบบจำลองกระบวนการให้นำโปรแกรมไฮซิส (HYSYS 3.1) มาใช้ทั้งสภาวะคงที่และสภาวะพลวัต และทำการควบคุมกระบวนการโดยใช้ตัวควบคุมแบบพีไอดี

ภาควิชา........วิศวกรรมเคมี......... ลายมือชื่อนิสิต......เลอศักดิ์ บุญณะชัย..........
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LERSAK BUNNACHAI:PERFORMANCE EVALUATION OF HEAT INTEGRATION AND CONTROL STRUCTURES OF DIMETHYL ETHER PLANT. ADVISOR: ASST. PROF. MONTREE WONGSRI, D.Sc., 172 pp.

Dimethyl ether (DME) is a clean and economical alternative fuel which can be produced from natural gas through synthesis gas. The properties of DME are very similar to those of LP gas. DME can be used for various fields as a fuel such as power generation, transportation, etc. In this research we present heat integrate plant (HIP) for three alternatives using disturbance propagation method is provided by Wongsri (1990) and three new plantwide control structure of DME process using Lyben (1998)'s heuristics and fixture point method to compare with Base Case of DME process to minimize energy usage and best control structure. From the result, HIP2 is the best alternative because it can be saved energy 58.8% and control structure 3 is the best plantwide control structure to give the minimize IAE score. HYSYS 3.1 was used to simulate the DME production process in both steady state and dynamic modes, PID controller are provided to control in this process.

Department:.........Chemical Engineering.... Student's Signature: Lersak Bunnachai
Field of Study:.....Chemical Engineering.... Advisor's Signature.........................
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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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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

A chemical industry has become even more competitive as companies try to improve their profits and reduce production times to shorten the supply chain to the customer. Part of this competitiveness has lead to the design of highly complex processes. The increased complexity has been justified on the bases of improving energy recovery and unused raw material, and reducing the environmental impact of the process.

Dimethyl ether (DME) is a clean and economical alternative fuel which can be produced from natural gas through synthesis gas that is easy to liquefy and transport. The properties of DME are very similar to those of LP gas. DME can be used for various fields as a fuel such as power generation, transportation, etc. It contains no sulfur or nitrogen. It is not corrosive to any metal and not harmful to human body.

Significant potential in three major markets; power generation is already approved by manufacturers such as Mitsubishi, Hitachi and General Electric as a fuel for their gas turbines, DME is an efficient alternative to other energy sources for medium-sized power plants especially on islands or in isolated regions where is can be difficult to transport natural gas and where the construction of liquefied natural gas (LNG) regasification terminals would not be viable. DME is transported at a temperature of -25^0C, making it easier to handle than LNG, which is shipped at -163^0C. Its use would reduce costs across the supply chain because existing LPG infrastructure could be utilized. Domestic LPG substitute; likely to have a generally more attractive price structure than LPG, DME can be blended in a proportion of 15 to 20% in LPG, without necessitating modifications to equipment or distribution networks. Automotive Fuel; often described as "diesel LPG", DME is a future automotive fuel solution. Promoting its use in captive corporate and public fleets would initially reduce the problems of developing a clean distribution network, while taking advantage of its environmental benefits such as no particulate or sulfur emissions. In addition, few engine modifications would be required.

An innovative process of direct synthesis of DME from synthesis gas has been developed to minimum energy usage by heat integration process, but it increase the interaction between unit so a plantwide control strategy for DME plant is present.

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.

The main objective of this study is to use plantwide control strategies to develop the new control structures for the DME 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 Objectives of the Research

The objectives of this work are listed below:

1. To design Heat-integrated processes structures for DME plant.

2. To design control structures for heat exchanger networks (HENs) with heat-integrated process structures in DME plant.

3. To evaluate performance of design heat-integrated process structures for DME plant.

1.3 Scopes of the Research

1. Description and data of DME plant are obtained from Analysis, Synthesis, and Design of Chemical Process, Prentic Hall, Richard Turton (2003).

2. The heat exchanger network with control structures of the DME plant are programmed using a commercial process simulator HYSYS for control structure performance tests.

3. The design heat-integrated process structures for DME plant for 3 alternatives.

4. The design control structures for heat-integrated process DME plant are design using Luyben's heuristics method for 1 alternative and using fixture point method for 2 alternatives.

1.4 Contributions of the Research

The contributions of this work are as follows:

1. The new plantwide control structures for typical of DME process.

2. The new heat-integrated processes structures for DME process.

3. The new plantwide control structures with heat-integrated processes structures for DME process.

4. Process flow diagrams of DME process with heat-integration process have been simulated.

1.5 Research Procedures

The procedures of this research are as follows:

1. Study plantwide process control theory.

2. Study DME process and related information.

3. Study and design heat-integrated processes structures for DME process by using HEN heuristics.

4. Steady state simulation of heat-integrated processes structures of DME process.

5. Dynamic simulation of heat-integrated processes structures of DME process.

6. Development of the new design plantwide heat-integrated process structures for DME process.

7. Dynamic simulation for the heat-integrated process structures for DME process alternative.

8. Evaluation and analysis of the dynamic performance of the heat-integrated processes structures.

9. Conclusion of the thesis.

1.6 Research Contents

This thesis is divided into six chapters.

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

Chapter II reviews the work carried out on heat exchanger networks (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 the DME plant.

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

Chapter VI the overall conclusions and recommendations of this thesis are discussed.

CHAPTER II

LITERATURE REVIEWS

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 (1988). 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 (1988) 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 (1988) 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.

Terrill and Douglas (1987) 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 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 Design and Control Structure

Vasbinder and Hoo (2003) present the concept of plantwide control structure in a DME process. They use method base on a modified version of the decision-making methodology of the analytic hierarchical process (AHP). The decomposition utilizes a series of steps to select among a set of competing modules. The control structure for each of the individual modules was developed using Luyben' nine-step approach. The decomposition serves to make the plantwide control problem tractable by reducing the size of the problem, while the modified analytic hierarchical process guarantees consistency. The modular decomposition approach was applied to the dimethyl ether (DME) process, and the results were compared to a traditional plantwide design approach. Both methods produced the same control structure that was shown to be adequate for the process. Satisfactory disturbance rejection was demonstrated on the integrated flowsheet. Future work will include demonstrating the approach on a more complex flowsheet and employing a model-based centralized control structure.

Van der Lee, Young and Svrcek (2002) present a background on Dimethyl Ether (DME) that included it's current uses, potential as an alternative fuel and the current and future production processes. A detailed steady state and dynamic simulation of the current methanol dehydration production process was performed. Two control strategies were examined using the dynamic simulation of the plant, a base control strategy which utilized PI and PID controllers exclusively, and an alternative strategy, which incorporated a DMC controller to control the methanol distillation column. It was found that the base control scheme showed a good response for both circulation flow rate and feed composition change however the response to a composition controller set points changes was poor for the methanol column. The implemented DMC controller resulted in slightly better result foe the methanol column bottoms composition set point changes when compared to the base case controller scheme, however if failed to effectively reject circulation rate and feed composition disturbance.

Handogo and 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 fuid 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.

Luyben and 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.

Luyben, Tyreus and 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.

Luyben (2000) studied the process had the exothermic, irreversible, gas-phase reaction A + B $\rightarrow$ 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.

Kietawarin (2002) presented a comparison among 4 control structures designed for withstanding disturbances that cause production rate change of HDA process. The changes had been introduced to the amount of toluene and feed temperature before entering the reactor. Compared with the reference control structure using a level control to control toluene quantity in the system, the first control scheme measured toluene 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, Tyreus and Luyben 1998, the result was performance of these structures higher than reference.

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 (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.

2.3 Heat Exchanger Networks (HENs)

Linhoff and Hindmarsh (1983) presented a novel method for the design of HEN. The method is the first to combine sufficient simplicity to be used by hand with near certainty to identify "best" designs, even for large problems. Best design features the highest degree of energy recovery possible with a given number of capital items. Moreover, they feature network patterns required for good controllability, plant layout, intrinsic safety, etc. Typically, 20-30 % energy savings, coupled with capital 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, Dunford and Smith (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 (1984) 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 (1986) 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 ?Tmin 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 (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 (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.

CHAPTER III

THEORY

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 Plantwide Control Design Procedures

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.

3.1.1 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.

3.1.2 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).

3.1.3 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.

3.1.4 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.

3.1.5 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)

3.1.6 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.

3.1.7 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.

3.1.8 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.

3.1.9 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.

3.2 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.

3.2.1 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 3.1 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.

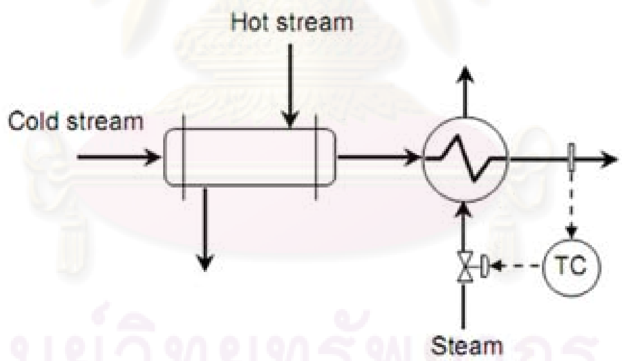

Figure 3.1 Control of process-to-process heat exchanger using the auxiliary utility

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.

3.2.2 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 3.2 shown 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 3.2.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.

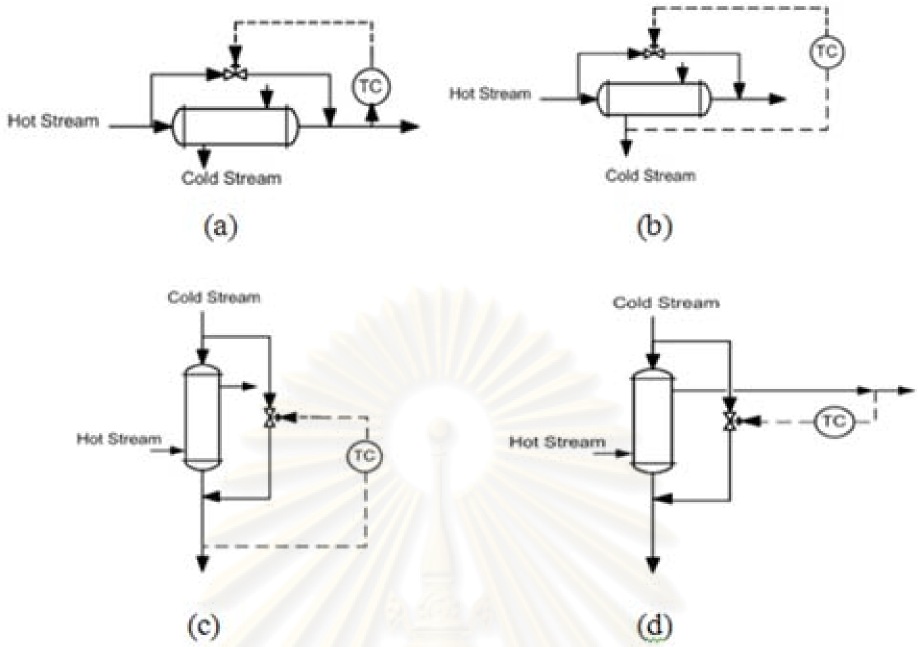

Figure 3.2 Bypass control of process-to-process heat exchangers. (a) Controlling and bypassing hot stream; (b) controlling cold stream and bypassing hot stream; (c) controlling and bypassing cold stream; (d) controlling hot stream and bypassing hot stream.

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

3.3 Heat Exchanger Network

It is generally accepted that an optimal network must feature a minimum number of units that reflects on a capital cost and minimum utility consumption that reflects on operating costs. A good engineering design must exhibit minimum capital and operating costs. For Heat Exchanger Network (HEN) synthesis, other features that are usually considered in design are operability, reliability, safety, etc. in recent years the attention in HEN synthesis has been focused on the operability features of a HEN, e.g. the ability of a HEN to tolerate unwanted changes in operating conditions. It has been learned that considering only a cost objective in synthesis may lead to a worse network, i.e. a minimum cost network may not be operable at some neighboring operating conditions. The design must not only feature minimum cost, but also be able cope with a fluctuation or changes in operating conditions. The ability of a HEN to tolerate unwanted changes is called *resiliency*. It should be note that the ability of a HEN to tolerate wanted changes is called *flexibility*.

The resiliency property of a design becomes an important feature to be accounted for when the extent of integration of a design introduces significant interactions among process components. The energy integration of a HEN generates a quite complex interaction of process streams, despite the fact that transfer of heat from hot to cold process streams is the only activity of the network. The goal of a network is to deliver the process streams to their target temperatures by using most of their heating and cooling availability and a minimum of heating and cooling utilities. The process streams are coupled through a net of heat exchangers. Changes in conditions of one stream in the network may affect the performances of many heat exchanges and the conditions of several process streams. Since resiliency is a property of a network structure.

3.3.1 Definition of HEN Resiliency

In the literature, resiliency and flexibility have been used synonymously to describe the property of HEN to satisfactorily handle variations in operating conditions. These two terms have difference in meaning.

The resiliency of a HEN is defined as the ability of a network to tolerate or remain feasible for disturbances in operating conditions (e.g. fluctuations of input temperatures, heat capacity flowrate, etc.). As mentioned before, HEN flexibility is closed in meaning to HEN resiliency, but HEN flexibility usually refers to the wanted changes of process conditions, e.g. different nominal operating conditions, different feed stocks, etc. That is, HEN flexibility refers to the preservation of satisfactory performance despite varying conditions, while flexibility is the capability to handle alternate (desirable) operating conditions.

A further distinction between resiliency and flexibility is suggested by Colberg el al. (1989). Flexibility deals with planed, desirable changes that often have a discrete set of values, resilience deal with unplanned, undesirable changes that naturally are continuous values. Thus a flexibility is a 'multiple period' type of problem. A resilience problem should be a problem with a continuous range of operating conditions in the neighborhood of nominal operating points.

Wongsri (1990) developed the heuristic and procedures for resilient heat exchanger network synthesis. The heuristics are used to develop basic and derived match patterns which were classified according to their (1) resiliency (2) chances that they are in solution and (3) the matching rules like the pinch method, and the thermodynamics law etc. Furthermore the same author developed for synthesize heat exchanger network called "The Disturbance Propagation Method". This method will find a resiliency network structure directly from the resiliency requirement and also feature minimum number of units and maximum energy recovery.

3.3.2 Design Conditions

There are several design conditions for resilient HEN synthesis. Usually, these are specified at extreme operating conditions. The following conditions (Wongsri, 1990) are:

1. *Nominal Operating Condition.* This is an operating condition that is obtained from a steady state heat and mass balance of a process. In a good design, a network must be operated at this condition most of the time. In general, a fluctuation in operating condition is plus and minus from this point.

2. *Maximum Heat Load Condition.* This is a condition where all process stream are at their maximum heat loads. For example inlet temperatures of hot streams are the highest and of cold streams are the lowest. This is also known as the largest maximum energy recovery condition.

3. *Maximum Cooling Condition.* This is a condition where hot process streams are at their maximum heat loads whereas cold process streams are at their minimum heat loads. For example inlet temperatures of hot and cold streams are the highest.

4. *Minimum Heating Condition.* This is a condition where hot process streams are at their minimum heat loads whereas cold process streams are at their maximum heat loads. For example inlet temperatures of hot and cold streams are the lowest.

5. *Minimum Heat Load Condition.* This is a condition where all process streams are at their minimum heat loads. For example inlet temperatures of hot streams are the lowest and of cold streams are the highest. This is also known as the lowest maximum energy recovery condition.

3.3.3 Match Patterns

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

By using the 'tick off rule' there are four match patterns for a pair of hot and cold streams according to the match position and the length of streams. The four patterns are considered to the basic match pattern classes. The members of these classes are the patterns where other configurations and properties are specified. The four match pattern classes are simply called A, B, C and D and are shown in Figure 3.3 to 3.6 respectively.

1. *Class A Match Pattern:* The heat load of a cold stream is greater than the heat load of a hot stream in a pattern, i.e. the hot stream is totally serviced. The match is positioned at the cold end of the cold stream. The residual heat load is on the hot portion of the cold stream. (See Figure 3.3) A match of this class is a first type match at cold end position and the heat load of the cold stream is greater than that of the hot stream. This is a upstream match. For a heating subproblem, a Class A match is favored, because it leaves a cold process stream at the pinch heuristics.

2. *Class B Match Pattern:* The heat load of a hot stream is greater than the heat load of a cold stream in a pattern, i.e. the cold stream is totally serviced. The match is positioned at the hot end of the hot stream. The residual heat load is on the cold portion of the hot stream. (See Figure 3.4) A match of this class is a second type match; a hot end match and the heat load of the hot stream are greater than that of the cold stream. This is an upstream match. For a cooling subproblem, a Class B match is favored, because it leaves a hot process stream at the cold end also follows the pinch heuristics.

3. *Class C Match Pattern:* The heat load of a hot stream is greater than the heat load of a cold stream in a pattern, i.e. the cold stream is totally serviced. The match is positioned at the cold end of the hot stream. The

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

4. *Class D Match Pattern:* The heat load of a cold stream is greater than the heat load of a hot stream in a pattern, i.e. the hot stream is totally serviced. The match is positioned at the hot end of the cold stream. The residual heat load is on the cold portion of the cold stream. (See Figure 3.6) A match of this class is a second type match; a hot end match and the heat load f the cold stream is greater than that of the hot stream. This is a downstream match.

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

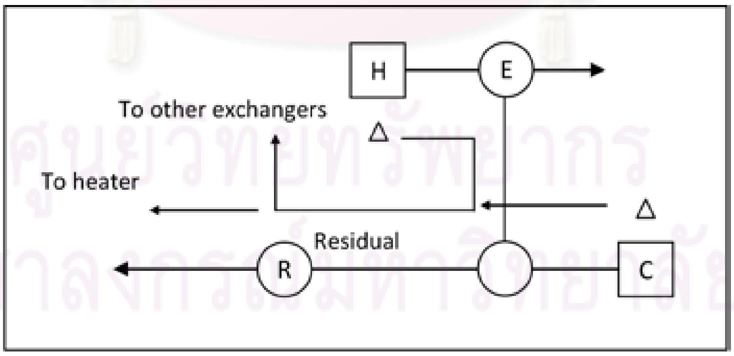

Figure 3.3 Class A Match Pattern.

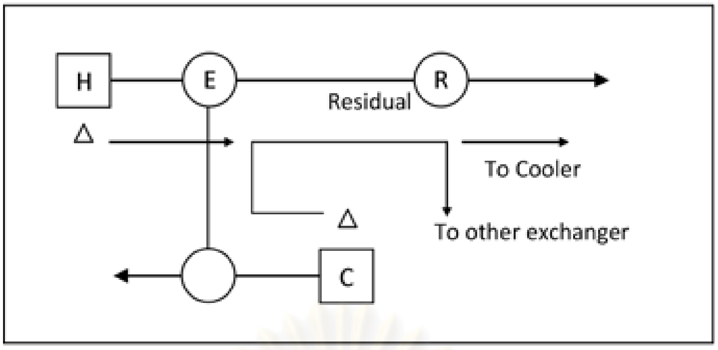

Figure 3.4 Class B Match Pattern.

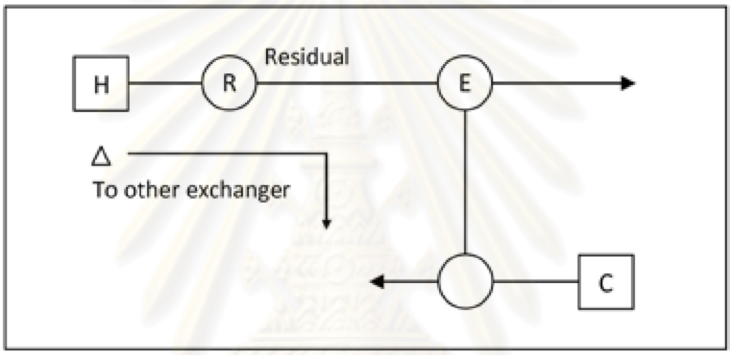

Figure 3.5 Class C Match Pattern.

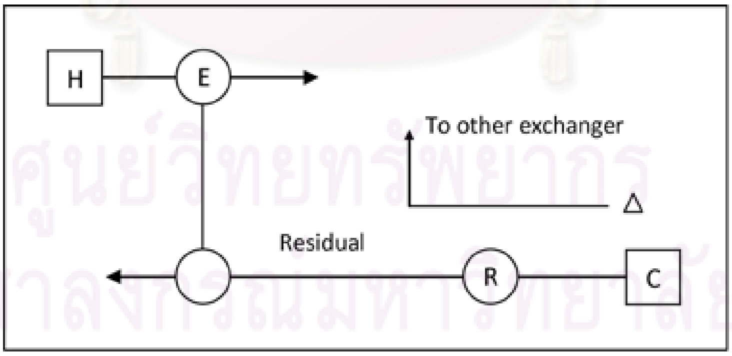

Figure 3.6 Class D Match Pattern.

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

3.3.4 Disturbance Propagation Design Method

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

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

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

By choosing the minimum heat load condition for the design, the new input temperature of a residual stream to its design condition according to the propagated disturbance. The propagated disturbance will proportionally cause more temperature variation in the residual stream and the range of temperature variation of the residual stream will be larger than its original range. The propagated disturbance of a stream is the disturbance caused by a variation in heat load of 'up-path' streams to which such a stream is matched. Only a residual stream will have a propagated disturbance. The new disturbance load of a residual stream will be the sum of its own disturbance (if any) and the propagated disturbance. See Figure 3.7 and 3.8.

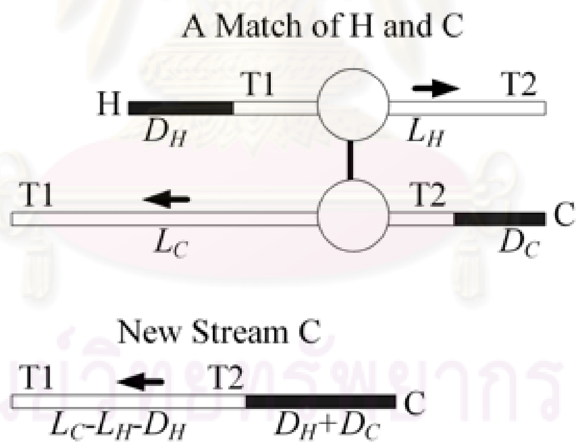

Figure 3.7 A Concept of Propagated Disturbance

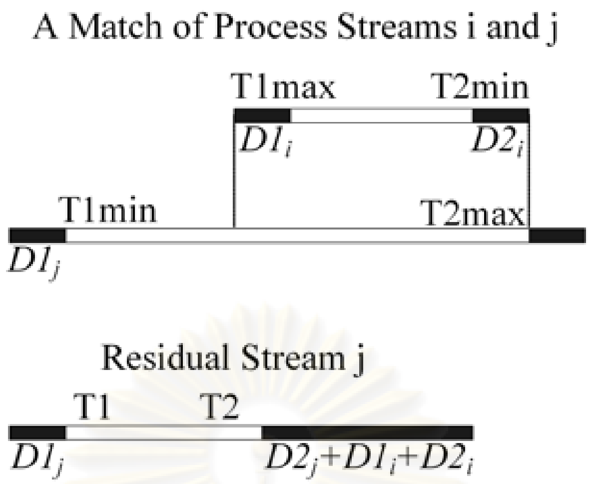

Figure 3.8 A General Concept of Propagated Disturbance

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

CHAPTER IV

DIMETHYL ETHER (DME) PROCESS

4.1 Process Description

Dimethyl ether (DME) is used primarily as propellant. DME is miscible with most organic solvents, it has a high solubility in water, and it is completely miscible in water and 6% ethanol. Recently, the use of DME as a fuel additive for diesel engines has been investigated due to its high volatility (desirable for cold starting) and high cetane number. The production of DME is via the catalytic dehydration of methanol over an acid zeolite catalyst. The main reaction is as follows:

$$2CH_3OH \rightarrow (CH_3)_2O + H_O$$

In the temperature range of normal operation, there are no significant side reactions.

A preliminary process flow diagram for a DME process is shown in Finger 4.1, in which 50,000 metric tons per year of 99.5 wt% purity DME product is produced. Due to the simplicity of the process, a stream factor of 0.95 (8375 h/yr) is used.

Fresh methanol is combined with recycle reactant, and vaporized prior to being sent to a fixed-bed reactor operating between 250^0C and 368^0C. The single-pass conversion of methanol in the reactor is 80%. The reactor effluent is then cooled prior to being sent to the first of two distillation columns, T-100 and T-101. DME product is taken overhead from the first column. The second column separates the water from the unused methanol. The methanol is recycled back to the front end of the process, while the water is sent to waste water treatment to remove trace amounts of organic compounds.

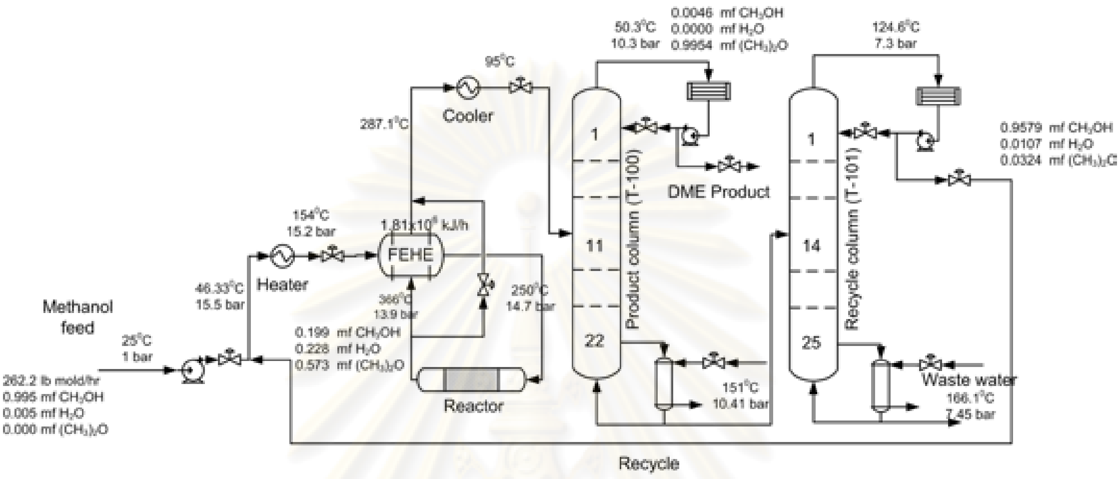

Figure 4.1 The flow sheet of DME process (Turton, 1998.)

The reaction taking place is mildly exothermic with a standard heat of reaction, $\Delta H_{reac}(25^0C)$ = -11,770 kj/kmol. The equilibrium constant for this reaction at three different temperatures below:

Table 4.1 The equilibrium constant (Kp) for reaction

T	K_P
473 K (200^0C)	92.6
573 K (300^0C)	52.0
673 K (400^0C)	34.7

The corresponding equilibrium conversions for pure methanol feed over the above temperature range are greater than 92%. The equilibrium constants reported above appear to be higher that those calculated by method using standard Gibbs free energy and heat of formation data. The single-pass conversion of 90%

used above may not be attainable due to equilibrium constants. A singer-pass conversion of 80% may be more realistic goal for this design.

The reaction takes place on an amorphous alumina catalyst treated with 10.2% silica. There are no significant side reactions below 400^0C. Above 250^0C, the rate equation is given as:

$$-r_{methanol} = k_0 \exp\,[-E_a/\mathrm{RT}]p_{methanol}$$

Where k_0 = 1.21 x 106 kmol/(m^3 catalyst filled reactor h kpa), E_a = 80.84 kJ/mol, and $P_{methanol}$ = partial pressure of methanol (kPa)

4.2 Design of Heat Exchanger Networks of DME Process

Heat exchanger networks for DME process are designed by Wongsri (1990) method. 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 DME process. The information for design is shown in the following Table 4.2

Table 4.2 The information of DME Process

Stream Name	Tin^0C	$Tout^0C$	W $(kW/^0C)$	Duty (kW)
H1 : Reactor effluence stream	287.15	95.00	19.37	3722.08
H2 : Waste Water Stream	150.82	49.94	3.26	328.85
H3 : Product Column Condenser	50.30	46.35	235.42	927.88
H4 : Recycle Column Condenser	124.61	121.69	496.19	1452.56
C1 : Vaporizer Feed Stream	46.80	154.00	35.67	3823.49
C2 : Product Column Reboiler	142.84	151.07	119.86	985.99
C3 : Recycle Column Reboiler	163.60	166.26	535.52	1421.35

4.2.1 Resilient Heat Exchanger Network of DME process alternative 1

The Table 4.3 is the process stream data that is chosen for 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 46.8^0C in cold stream and 56.8^0C in hot stream. The minimum utility requirements have been predicted 101.52 kW of hot utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.4. Figure 4.2 shows a design of resilient heat exchanger network for DME process alternative 1. In our case as shown in Figure 4.2, the minimum temperature difference in the process-to-process-heat-exchangers ΔT_{min} is set to be 10^0C.

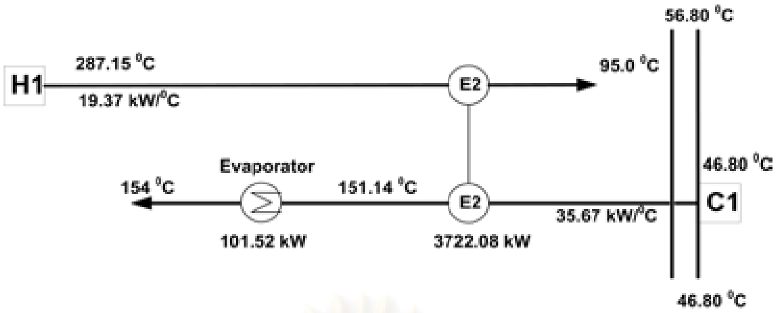

Figure 4.2 The resilient heat exchanger network alternative 1

table 4.3 Process stream data for alternative 1

Stream Name	Tin^0C	$Tout^0C$	W $(kW/^0C)$	Duty (kW)
H1	287.15	95.00	19.37	3722.08
C1	46.80	154.00	35.67	3823.49

Table 4.4 Problem table for alternative 1

W $(kW/^0C)$		T hot	T cold	ΣW	ΔT	Required	Interval	Cascade	Sum
H1	C1	(^0C)	(^0C)	(kW/^oC)	(^0C)	Heat(kW)	(kW)	Heat(kW)	Interval(kW)
0.00	0.00	87.15	277.15	0.00	0.00	Qh	0.00		0.00
19.37	0.00	164.00	154.00	19.37	123.15	101.52	2385.52	2487.03	2385.52
19.37	35.67	95.00	85.00	-16.30	69.00	2487.03	-1124.52	1362.51	1260.99
0.00	35.67	56.80	46.80	-35.67	38.20	1362.51	-1362.51	0.00	-101.52
								Qc	

synthesis table for hot end of alternative 1

state 1	Load	W	T1	T2	D1	D2	Action
H1	3625.22	19.37	282.15	95.00	193.70	0.00	selected
C1	3645.15	35.67	154.00	51.80	0.00	356.68	selected
state 2	Load	W	T1	T2	D1	D2	Action
H1	0.00	0.00	0.00	0.00	0.00	0.00	Matched to C1
C1	19.92	35.67	154.00	153.44	0.00	550.38	to Heater

4.2.2 Resilient Heat Exchanger Network of DME process alternative 2

The Table 4.5 is the process stream data that is chosen for alternative 2. There are two streams in the network. We do not find pinch temperature using problem table method as shown in Table 4.6. The minimum utility requirements have been predicted 2735.55 kW of cold utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.6. Figure 4.3 shows a design of resilient heat exchanger network for DME process alternative 2. 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 10^0C.

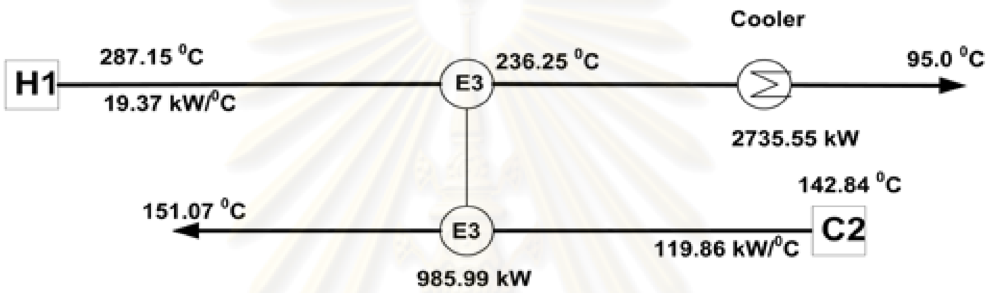

Figure 4.3 The resilient heat exchanger network alternative 2

Table 4.5 Process stream data for alternative 2

Stream Name	Tin^0C	$Tout^0C$	W ($kW/^0C$)	Duty (kW)
H1	287.15	95.00	19.37	3722.08
C2	142.84	151.07	985.99	119.86

Table 4.6 Problem table for alternative 2

W ($kW/^0C$)		T hot	T cold	ΣW	ΔT	Required	Interval	Cascade	Sum
H1	C2	(^0C)	(^0C)	(kW/^oC)	(^0C)	Heat(kW)	(kW)	Heat(kW)	Interval(kW)
0.00	0.00	287.15	277.15	0.00	0.00	Qh	0.00	0.00	0.00
19.37	0.00	161.07	151.07	19.37	126.08		2442.17		2442.17
19.37	119.86	152.84	142.84	-100.49	8.23		-827.03		1615.14
19.37	0.00	95.00	85.00	19.37	57.84		1120.36		2735.50
								Qc	

synthesis table for hot end of alternative 2

state 1	Load	W	T1	T2	D1	D2	Action
H1	3528.25	19.37	277.15	95.00	193.70	0.00	Selected
C2	986.45	119.86	151.07	142.84	0.00	119.86	Selected
state 2	Load	W	T1	T2	D1	D2	Action
H1	2421.94	19.37	277.15	152.11	313.56	0.00	To cooler
C2	0.00	0.00	0.00	0.00	0.00	0.00	Matched to H1

4.2.3 Resilient Heat Exchanger Network of DME process alternative 3

The Table 4.7 is the process stream data that is chosen for 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 46.8^0C in cold stream and 56.8^0C in hot stream. The minimum utility requirements have been predicted 1088.40 kW of hot utilities.

The synthesis procedure using the disturbance propagation method and math pattern is shown in Table 4.8. Figure 4.4 shows a design of resilient heat exchanger network for DME process alternative 3. In our case as shown in Figure 4.4, the minimum temperature difference in the process-to-process-heat-exchangers ΔT_{min} is set to be 10^0C.

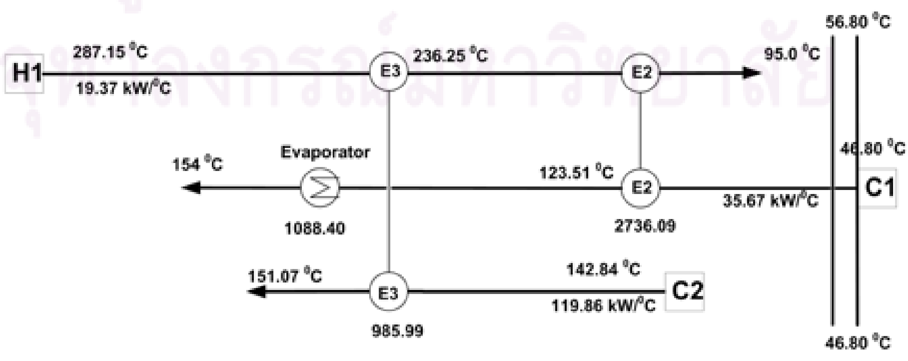

Figure 4.4 The resilient heat exchanger network alternative 3

Table 4.7 Process stream data for alternative 3

Stream Name	Tin^0C	$Tout^0C$	W $(kW/^0C)$	Duty (kW)
H1	287.15	95.00	19.37	3722.08
C1	46.80	154.00	35.67	3823.49
C2	142.84	151.07	985.99	119.86

Table 4.8 Problem table for alternative 3

W $(kW/^0C)$			T hot	T cold	ΣW	ΔT	Required	Interval	Cascade	Sum
H1	C1	C2	(^0C)	(^0C)	(kW/^oC)	(^0C)	Heat(kW)	(kW)	Heat(kW)	Interval(kW)
0.00	0.00	0.00	287.15	277.15	0.00	0.00	Qh	0.00	0.00	0.00
19.37	0.00	0.00	164.00	154.00	19.37	123.15	1088.04	2385.47	3473.51	2385.47
19.37	35.67	0.00	161.07	151.07	-16.30	2.93	3473.51	-47.75	3425.75	2337.71
19.37	35.67	119.86	152.84	142.84	-136.16	8.23	3425.75	-1120.60	2305.16	1217.12
19.37	35.67	0.00	95.00	85.00	-16.30	57.84	2305.16	-942.64	1362.51	274.47
0.00	35.67	0.00	56.80	46.80	-35.67	38.20	1362.51	-1362.51	0.00	-1088.04
									Qc	

synthesis table for hot end of alternative 3

state 1	Load	W	T1	T2	D1	D2	Action
H1	3625.10	19.37	282.15	95.00	193.70	0.00	Selected
C1	3645.47	35.67	154.00	51.80	0.00	356.70	
C2	163.45	19.86	151.07	142.84	0.00	19.86	Selected
state 2	Load	W	T1	T2	D1	D2	Action
H1	3441.79	19.37	277.15	99.46	213.56	0.00	Selected
C1	3467.12	35.67	154.00	56.80	0.00	356.70	Selected
C2	0.00	0.00	0.00	0.00	0.00	0.00	Matched H1
state 3	Load	W	T1	T2	D1	D2	Action
H1	0.00	0.00	0.00	0.00	0.00	0.00	Matched C1
C1	25.34	35.67	154.00	153.29	0.00	570.26	to Heater
C2	0.00	0.00	0.00	0.00	0.00	0.00	Matched H1

The various alternatives of heat exchanger network are designed for the DME process, the energy saved from the base case is shown in Table 4.9

Table 4.9 Energy integration for DME process

The DME process	Alternatives			
	Base-Case	AL 1	AL 2	AL 3
Vaporizer	3823.49	101.41	3823.49	1087.40
Cooler1	3722.08	0	2736.09	0
Cooler2	328.85	328.85	328.85	328.85
Product column reboiler	985.99	985.99	0	0
Recycle column reboiler	1421.35	1421.35	1421.35	1421.35
Hot utilities usage, (kW)	6230.83	2508.8	5244.8	2508.8
Product column condenser	927.88	927.88	927.88	927.88
Recycle column condenser	1452.56	1452.56	1452.56	1452.56
Cold utilities usage, (kW)	6431.37	2709.29	5445.38	2709.29
Total Hot &Cold utilities (kW)	12662.20	5218.04	10690.22	5218.04
Energy savings from RHEN,%	0	58.79	15.57	58.79

4.3 Alternative Structures of DME Process

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

Figure 4.5 shown the base case of the dimethyl ether process with simply energy integration, they used a feed-effluent heat exchanger (FEHE) to vaporize the reactor feed stream.

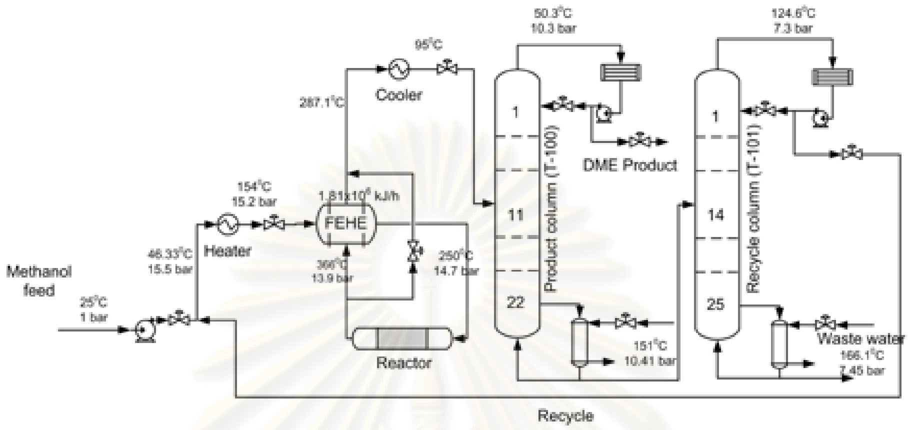

Figure 4.5 Dimethyl Ether process Base Case

In alternative 1 part of the heat in the reactor effluent stream is used to vaporize the feed stream but vaporizer is needed for the process because total heat from reactor is not enough to vaporize all feed stream as shown on Figure 4.6.

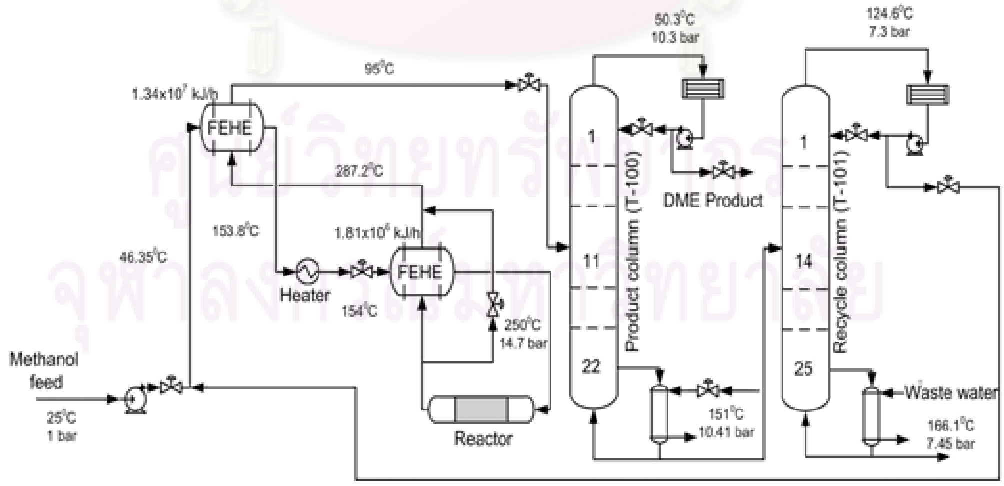

Figure 4.6 Dimethyl Ether process Alternative 1

In alternative 2 part of the heat in the reactor effluent stream is used to drive the product column reboiler that does not add the auxiliary reboiler because the total heat from reactor effluent stream is enough for reboiler as shown on Figure 4.7.

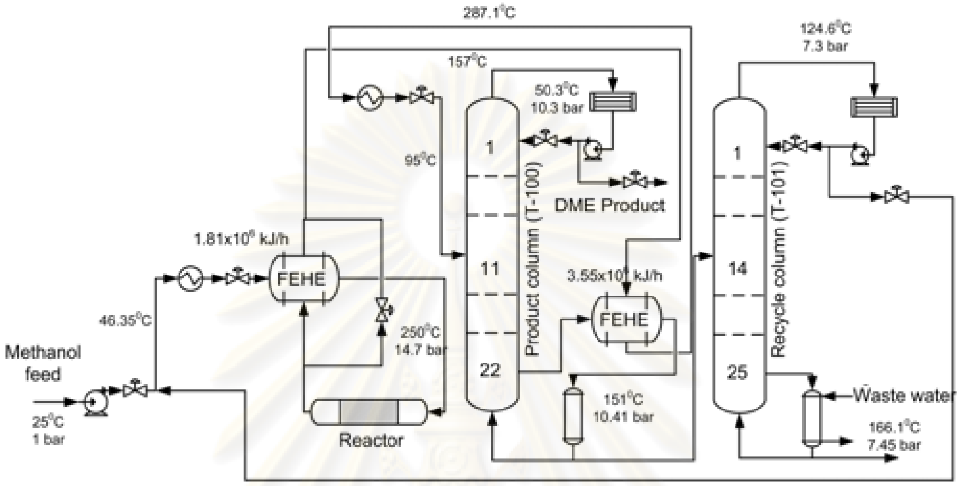

Figure 4.7 Dimethyl Ether process Alternative 2

In alternative 3 part of the heat in the reactor effluent stream is used to drive the product column reboiler and vaporizer at feed stream. The auxiliary vaporizer is need to added at feed stream because total heat from effluent stream not enough for both of reboiler and vaporizer as shown on Figure 4.8.

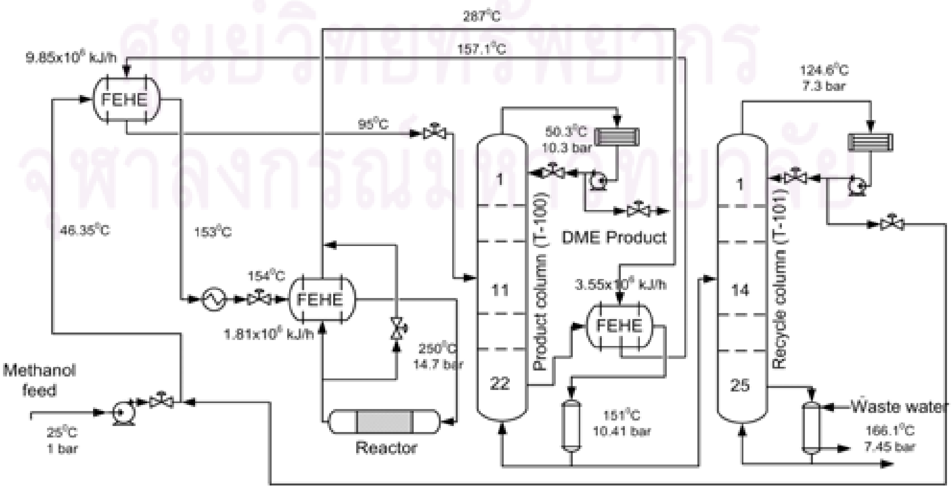

Figure 4.8 Dimethyl Ether process Alternative 3

4.4 Steady State Modeling of DME Process

First, a steady-state model is built in HYSYS.PLANT, using the flowsheet and equipment design information from Turton(1988). Figures 4.11 to 4.14 show the HYSYS flowsheets of the DME process with energy integration schemes for alternatives 1, 2, and3, respectively. For our simulation, UNIQUAC model is selected for physical property calculations because it can be applied to a wide range of mixture containing water, alcohols, nitriles, amines, esters, ketones, aldehydes, halogenated hydrocarbons and hydrocarbons, so it can be applied with dimethyl ether process. The reaction kinetics of reaction is modeled with standard Arrhenius kinetic expressions available in HYSYS.PLANT, and the kinetic data are taken from Bondiera and Naccache.

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 distillate rate and overhead dimethly ether mole fraction for the product column. For the remaining columns, distillate rate and bottom water mole fraction are selected. 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.

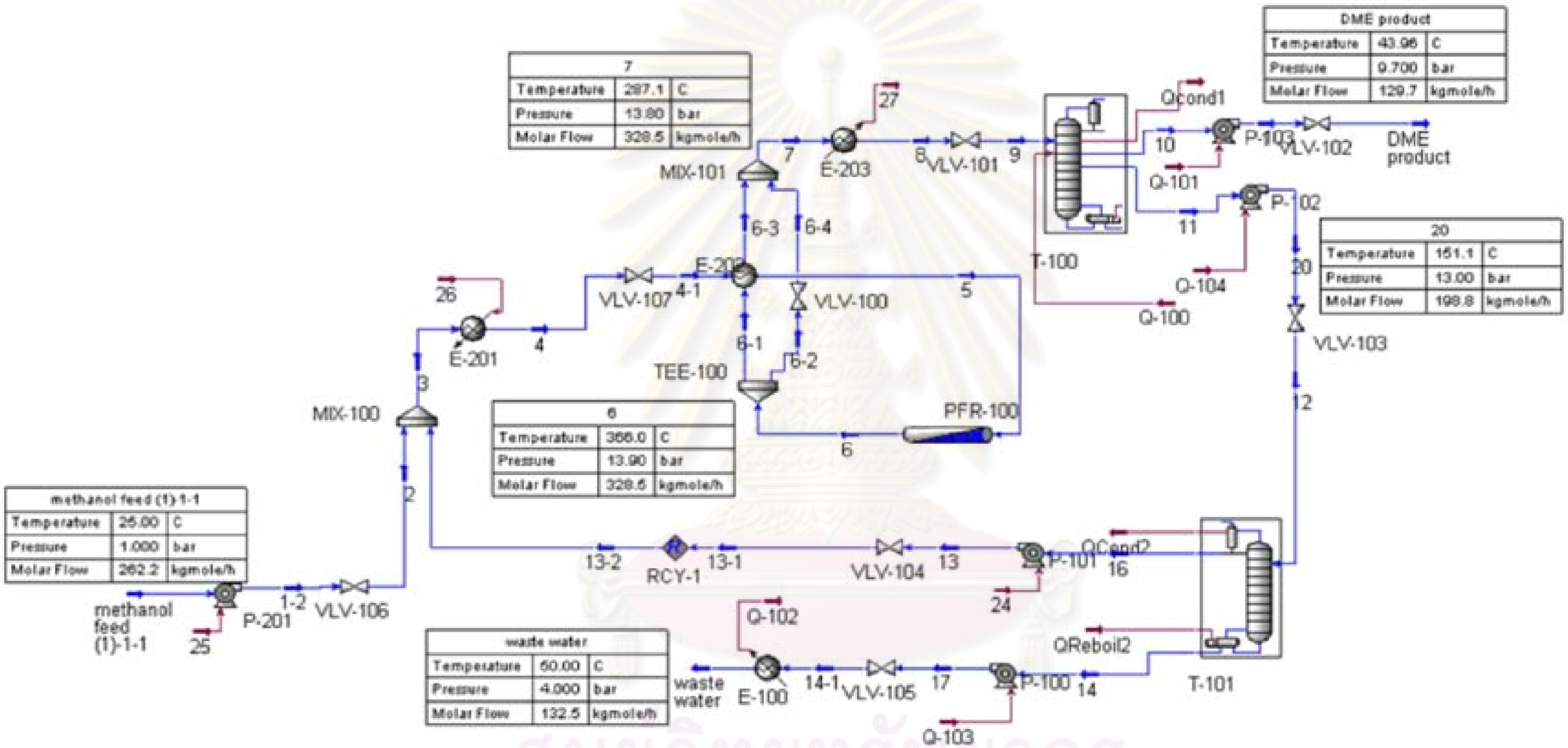

Figure 4.9 HYSYS Flowsheet of the steady state modeling of DME process base-case

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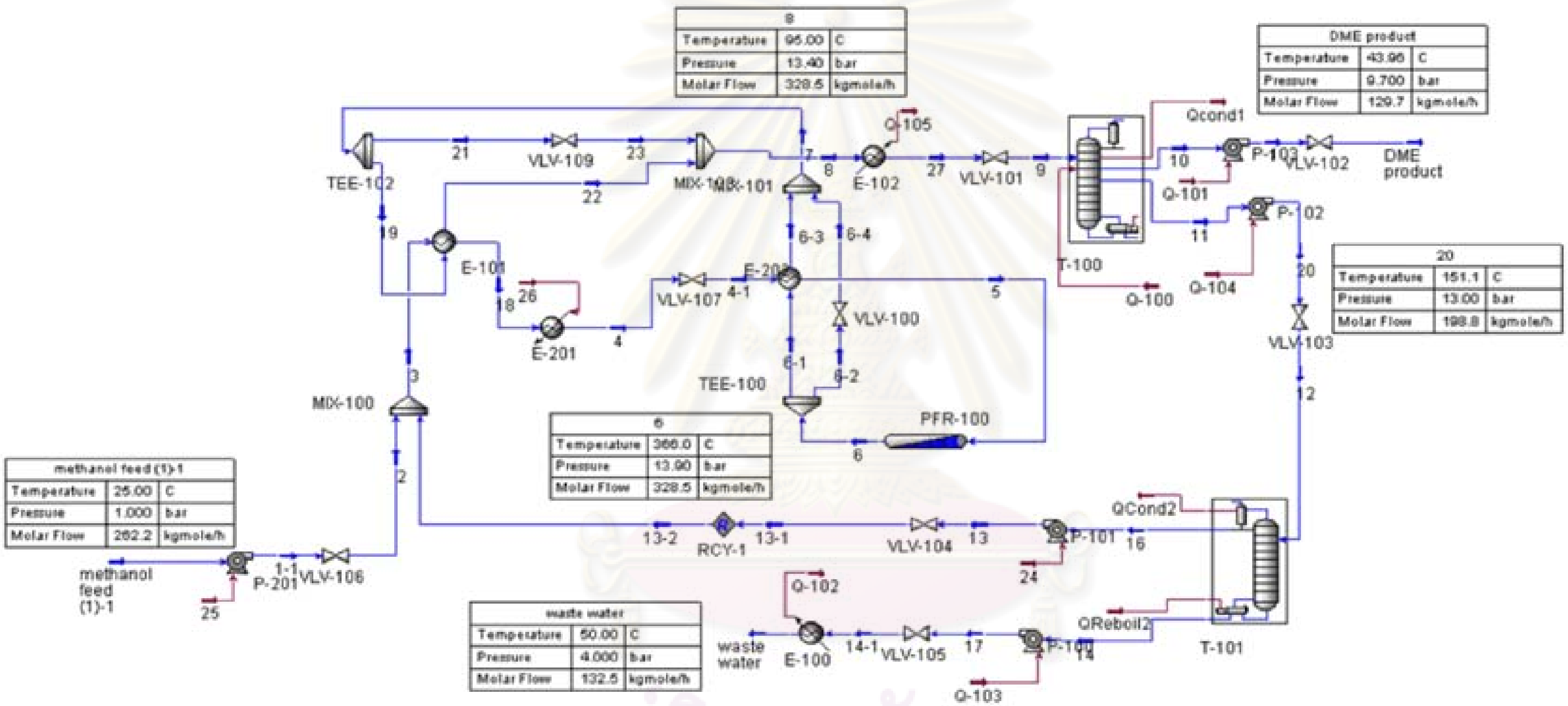

Figure 4.10 HYSYS Flowsheet of the steady state modeling of DME process alternative 1

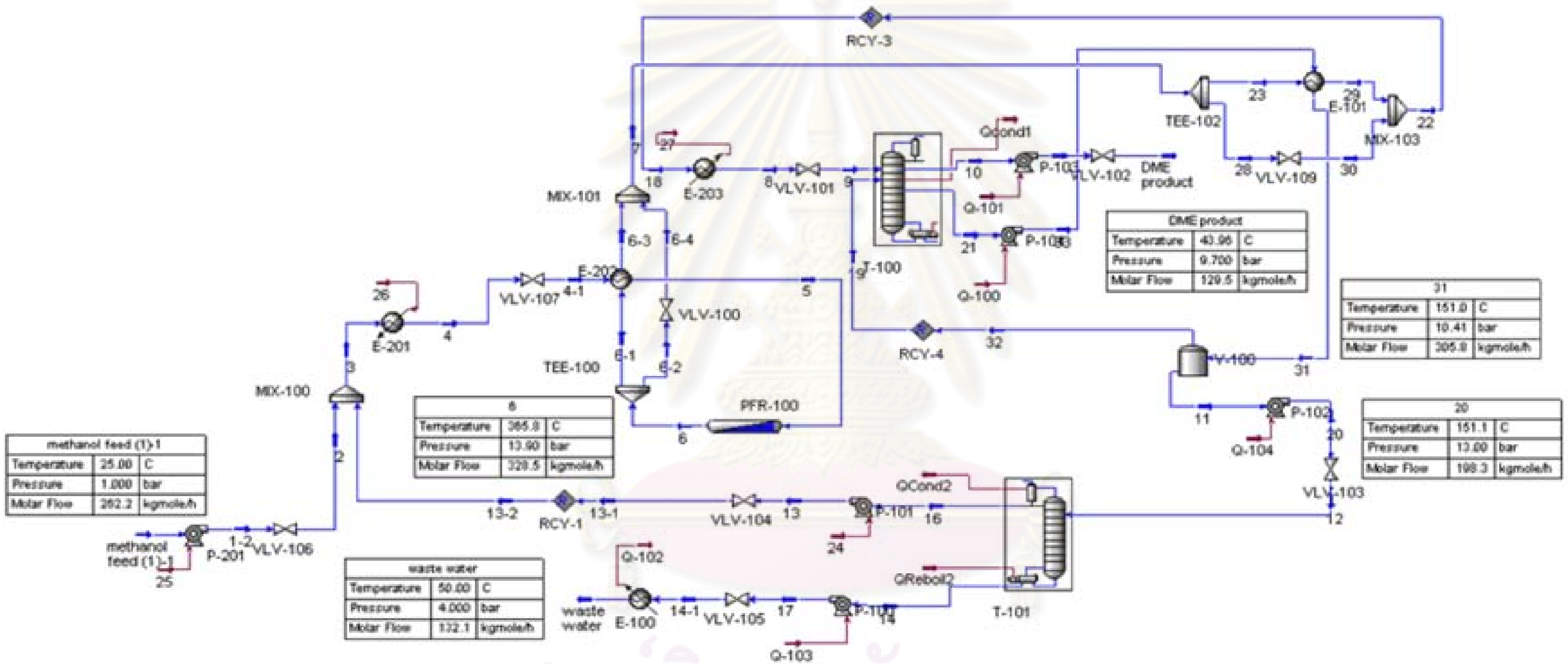

Figure 4.11 HYSYS Flowsheet of the steady state modeling of DME process alternative 2

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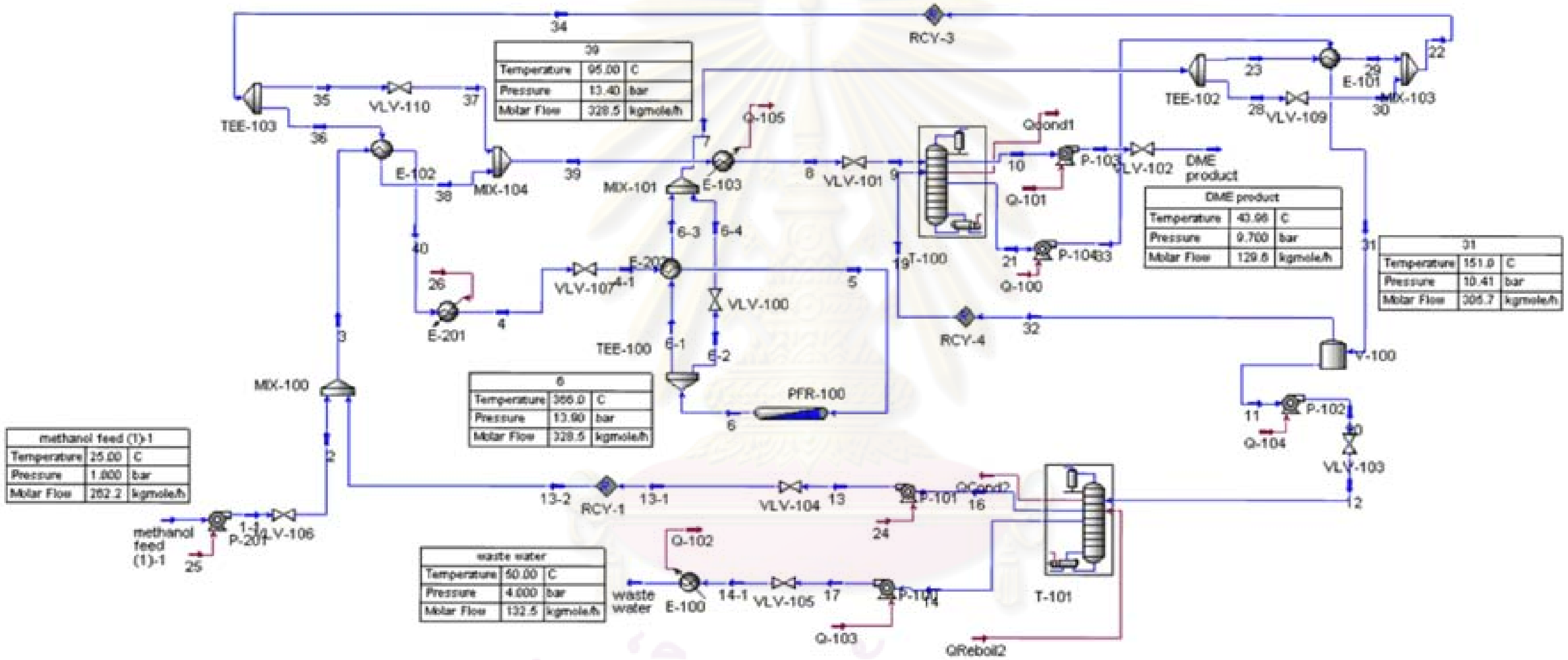

Figure 4.12 HYSYS Flowsheet of the steady state modeling of DME process alternative 3

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4.5 Energy Integration for Steady State Simulation of DME Process

From steady state simulation results by HYSYS, the energy saved from the base case heat consumption as shown in Table 4.10.

Table 4.10 Energy integration for DME process (Steady State Simulation)

The DME process	Alternatives			
	Base-Case	AL 1	AL 2	AL 3
Vaporizer	3828.29	106.22	3828.34	1091.52
Cooler1	3721.62	0.00	2735.32	0.00
Cooler2	328.47	328.69	328.65	328.56
Product column reboiler	986.78	985.98	0.00	0.00
Recycle column reboiler	1422.72	1421.35	1423.86	1422.70
Hot utilities usage, (kW)	6237.80	2513.54	5252.19	2514.22
Product column condenser	927.85	927.88	928.07	927.63
Recycle column condenser	1454.33	1452.56	1455.06	1453.98
Cold utilities usage, (kW)	6432.27	2709.12	5447.10	2710.18
Total Hot & Cold utilities (kW)	12670.07	5222.67	10699.30	5224.40
Energy savings from RHEN,%	0.00	58.78	15.55	58.77

CHAPTER V

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 and energy integrated process. Moreover, the three new designed control structures and energy integrated process are also compared with base case of DME process that given by Turton(1998) based on rigorous dynamic simulation by using the commercial software HYSYS.

5.1 Plantwide Control Strategies

The plantwide control structure tool can be applied to the modules. Here, the nine-step approach of Luyben and Fixture point theorem are selected for demonstration on each of the DME module the module and discussion below.

5.1.1 Nine-step approach of Luyben

Step1. Establish the Design and Dynamic Control Objectives for the Module

The design and dynamic objectives for the separations module are to regulate the DME purity, maintain the production rate of DME, be flexible, and be profitable. The last two objectives are implicit and must be translated into directly measurable process variables. For the separations module, flexibility is associated with ability of the column to maintain the purity constraints, and profit is assigned to the product (DME) distillate rate.

Step2. Determine Control Degree of Freedom

Usually, this means counting the number of control valves, but other devices that qualify include the fan, mixer, etc. For the DME module, there are 17 control degrees of freedom, 10 control valves, and 7 utility streams. They include: methanol feed valve, heat exchanger feed valve, bypass for heat exchanger valve, stream inlet product column valve vaporizer and cooler utility streams: product column are included reflux valve, distillate and bottom product valves, condenser and reboiler utility streams: recycle column are included reflux valve, distillate and bottom product valves, condenser and reboiler utility streams, and cooler utility stream for waste water.

Step3. Establish Energy management system

The product DME is produced from the exothermic reaction of methanol at 250^0C. The reactor operates adiabatically, so for a given reactor design the exit temperature depends upon the reactor inlet temperature, and reactor conversion. Heat from the adiabatic reactor is carried in the effluent stream and it remove to stream inlet of reactor for base case.

The alternative way is using of the heuristic laws; Montree (1990) 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". Three alternatives were designed; first, we are integration with utility stream of vaporizer at feed stream; second, at the separations module we are integration with reboiler utility stream of distillation T100 and third, we are integration both of vaporizer and reboiler utility stream of distillation T100.

Step4. Set Production Rate

There are two places within the module where the production rate of DME can be controlled: the distillate valve or the feed valve of the DME column. The latter is selected by the base case to set the production rate on the basis of the process knowledge that control of the inventory in the condenser is better achieved by using the distillate flow of the DME column.

Step5. Control Product Quality and Address Safety, Operational, and Environmental Concerns

The DME product purity (99.5 percent) can be controlled using DME column reflux valve. The strategy is to employ an inferential measurement that is based on a temperature reading on stage 20. (The stage temperature was identified as being the most sensitive of temperature changing.)

The environmental constraint on the effluent water purity can be controlled using either the vapor boilup rate or the reboiler steam flow rate of the recycle column to maintain temperature in the recycle column like the DME 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 liquid levels must be controlled in this process. For the pressures, there are in the two distillation columns. DME column and recycle colume are pressure-controlled to the cooling utility stream at condenser.

The DME column condenser level is flow-controlled to the DME distillate valve, the DME reboiler level is flow-controlled to DME column bottoms valve, the recycle column condenser level is flow-controlled to the recycle column distillate valve and the recycle reboiler level is flow-controlled to the recycle column bottoms valve.

Step7. Check Component Balances

Component balances consists of: Dimethyl ether is removed from the system at production column by distillate flow and it is used to control level of condenser. Water is removed from the system at recycle column by bottom flow and it is used to control level of reboiler and methanol that reactant is sent back to the system by distillate flow of recycle column and it convert to dimethyl ether and water in reactor so no accumulation of any component in the system.

Step8. Control Individual Unit Operations

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

Cooling utility of the cooler controls temperature inlet of the product column. Heating utility of the vaporizer controls temperature of stream before inlet heat exchanger and feed flow controlled.

Step9. Optimize Economics or Improve Dynamic Controllability

This step is not considered in this work

5.1.2 Fixture Point Theorem

The fixture point theorem define the control variable that the most sensitivity. Defined control variable should consider to control and pairing with manipulate variable (MV) in the first.

Fixture point theorem analysis

1. *C*onsideration in dynamic mode of simulation until process set up to steady state.

2. *C*ontrol variable (CV) can be arranged to follow the most sensibility of the process variable by step change the MV (change only one MV, the other should be fixed then alternate to other until complete). Study the magnitude of integral absolute error (IAE) of all process variables that deviates

form steady state. This thesis considers six process variables including temperature, pressure, flow rate, composition, tank level and stage temperature.

3. Consider CV that give the most deviation from steady state (high IAE score) to match with MV. CV and MV should be directing interactive together, after that will consider the next CV to match with other MV.

5.2 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.

5.2.1 Design of Heat Pathways and HEN Control Configuration for Alternative 1

The design of the heat pathways for alternative 1 is shown in Figure 5.1. Both the positive and negative disturbance loads of C1 are directed to vaporizer; the corresponding of vaporizer duty is increased or decreased accordingly. The positive and the negative disturbance loads of H1 are shifted to vaporizer through the FEHE 1. Thus, the positive disturbance load of a hot stream will result in decrease of the vaporizer duty which is good. The negative disturbance load of hot stream will result in increase of the vaporizer duty which is ruled by ΔTmin constraint.

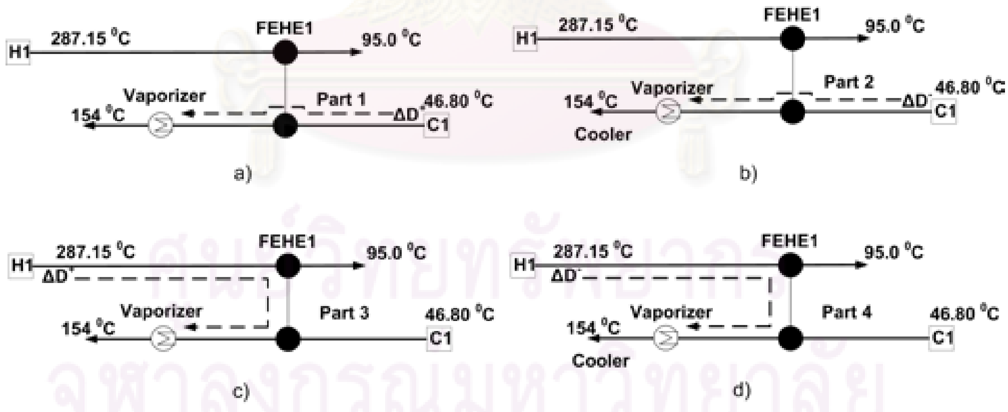

Figure 5.1 Heat pathways through alternative 1, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C1 to the vaporizer, (b) path 2 is used to shift the negative disturbance load of the cold stream C1 to the vaporizer, (c) path 3 is used to shift the positive disturbance load of the hot stream H1 to vaporizer, (d) path 4 is used to shift the negative disturbance load of the hot stream H1 to vaporizer.

From design the heat pathways for alternative 1, we can design the control configurations as show in Figure 5.2. These control systems involve two manipulated variable and two controlled variable and work as follows: the hot outlet temperature of FEHE1 is controlled at its nominal set point by manipulating the valve on the bypass line and the cold stream is nominal temperature controlled by manipulating the hot utility of vaporizer.

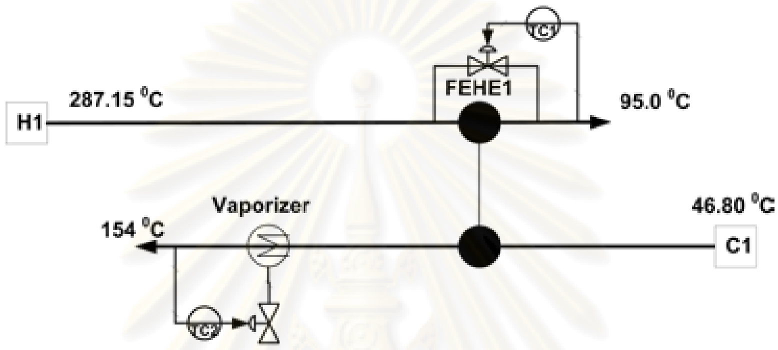

Figure 5.2 Control configurations of alternative 1

5.2.2 Design of Heat Pathways and HEN Control Configuration for Alternative 2

The design of the heat pathways for alternative 2 shown in Figure 5.3 shifts the positive and negative disturbance loads of C2 to cooler through the FEHE 2. Thus, the positive disturbance load of a cold stream will result in decrease of the cooler duty which is good. The negative disturbance load of a cold stream will result in increase of the heater duty which is ruled by ΔTmin constraint. The negative or positive disturbance load of H1 is directed to the cooler; the corresponding of cooler duty is increased or decreased accordingly.

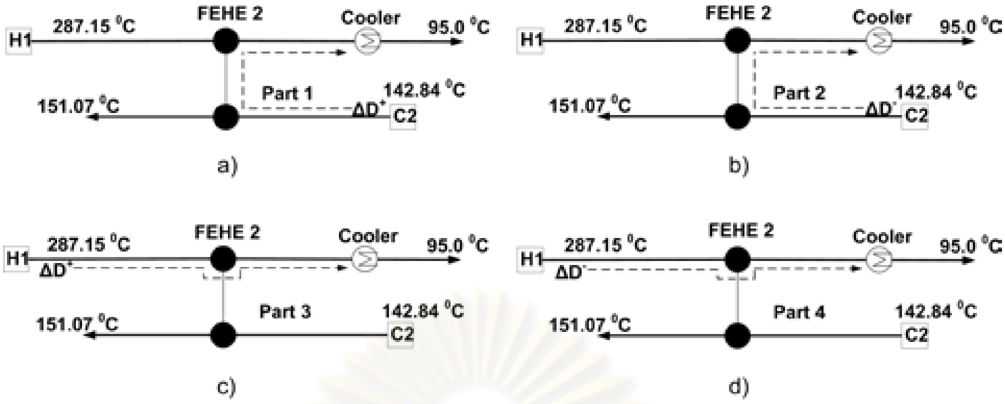

Figure 5.3 Heat pathways through alternative 2, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C2 to cooler, (b) path 2 is used to shift the negative disturbance load of the cold stream C2 to cooler, (c) path 3 is used to shift the positive disturbance load of the hot stream H1 to cooler, (d) path 4 is used to shift the negative disturbance load of the hot stream H1 to cooler.

From design the heat pathways for alternative 2, we can design the control configurations as show in Figure 5.4. These control systems involve two manipulated variable and two controlled variable and work as follows: the cold outlet temperature of FEHE2 is controlled at its nominal set point by manipulating the valve on the bypass line of hot stream and the hot stream is nominal temperature controlled by manipulating the cold utility of cooler.

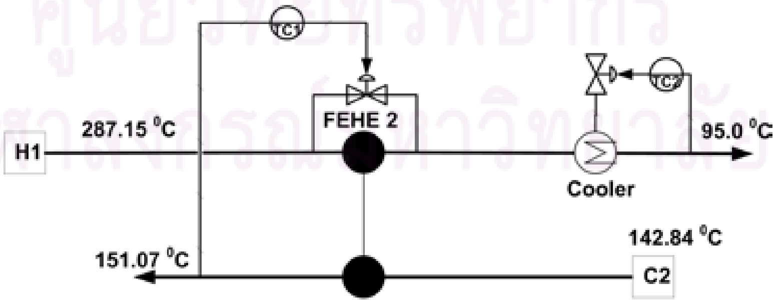

Figure 5.4 Control configurations of alternative 2

5.2.3 Design of Heat Pathways and HEN Control Configuration for Alternative 3

The design of the heat pathways for alternative 3 shown in Figure 5.5 the positive and negative disturbance loads of C1 are directed to vaporizer; the corresponding of cooler duty is increased or decreased accordingly, shifts the positive and the negative disturbance loads of H1 to the vaporizer through FEHE 4. Thus, the positive disturbance load of a hot stream will result in decrease of the vaporizer duty which is good. The negative disturbance load will result in increase of the vaporizer duty which is ruled by ΔTmin constraint and shift the positive and the negative disturbance loads of C2 to the vaporizer through FEHE 3 and FEHE 4. Thus, the positive disturbance load of a cold (C2) stream will result in increase of the vaporizer duty which is ruled by ΔTmin constraint. The negative disturbance load will result in decrease of the vaporizer duty which is good.

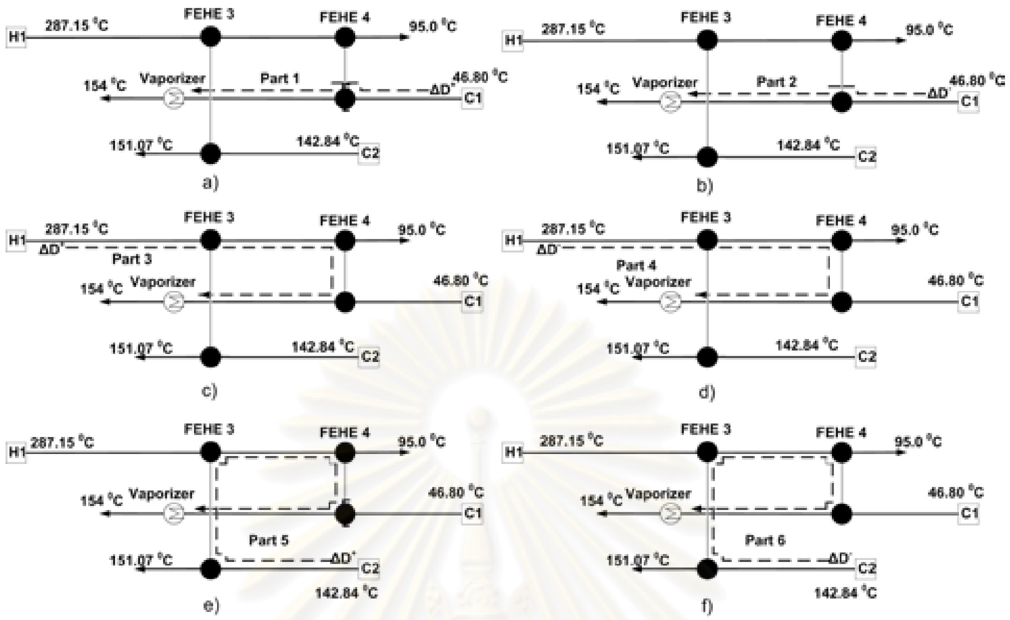

Figure 5.5 Heat pathways through alternative 3, where: (a) path 1 is used to shift the positive disturbance load of the cold stream C1 to the vaporizer, (b) path 2 is used to shift the negative disturbance load of the cold stream C1 to the vaporizer , (c) path 3 is used to shift the positive disturbance load of the hot stream H1 to vaporizer, (d) path 4 is used to shift the negative disturbance load of the hot stream H1 to vaporizer, (e) path 5 is used to shift the positive disturbance load of the cold stream C3 to vaporizer and (f) path 6 is used to shift the negative disturbance load of the cold stream C3 to vaporizer.

From design the heat pathways for alternative 3, we can design the control configurations as show in Figure 5.6. These control systems involve three manipulated variable and three controlled variable and work as follows: the hot stream outlet temperature of FEHE 4 is controlled at its nominal set point by manipulating the valve on the bypass line. The cold (C1) stream is nominal temperature controlled by manipulating the hot utility of vaporizer and the cold (C2) stream outlet temperature of FEHE3 is controlled at its nominal set point by manipulating the valve on the bypass line.

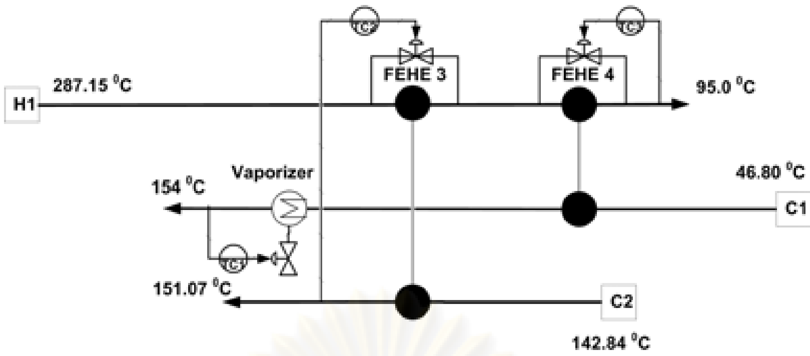

Figure 5.6 Control configurations of alternative 3

5.3 Design of Plantwide Control Structure

In this work, the DME process is designed by considering two control objectives (Turton, 1998); achieving a specified 50,000 metric tons per year and 99.5 wt% purity DME product is produced.

The major loops are the same as those used in Turton (1998), but we have designed three new heat exchanger alternatives and three new loops control structure, one Luyben's heuristic and two Fixture point method (Wongsri, 2008)

For all the heat integration units, the bypass streams are designed to control the outlet temperatures of FEHEs and the tray temperatures in the 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.

5.3.1 Design of Plantwide Control for the DME Process of the Base Case and Heat Integration Process Alternatives

Four control structures are present in this work. Control structure 1 (CS1) is design by Turton (1998) and three new control structures are designed following Luyben's heuristic in control structure 2 (CS2) and Fixture point method in control structure 3 and 4 (CS3 and CS4). The controller parameters are given in Appendices C.

5.3.1.1 Control Structure 1 (CS1) for Base Case of the DME Process

This control structure is shown in Figure 5.7. This control structure follow the fresh freed flow rate is controlled by valve, vaporizer outlet temperature is controlled by hot utility, heat exchanger outlet temperature for cold stream is controlled by bypass valve on hot side, feed temperature of the product column is controlled by cold utility. At separation section, the temperature of production column on stage 20 is controlled by hot utility of reboiler, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by the distillate valve and level of reboiler is controlled by bottom valve, the recycle column temperature on stage 24 is controlled by hot utility of reboiler, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by the distillate valve and level of reboiler is controlled by bottom valve. DME purity and recycle flow rate do not control at this control structure.

5.3.1.2 Control Structure 2 (CS2) for Base Case of the DME Process

This control structure is followed Luyben's heuristic that shown in Figure 5.8. This control structure follow; the fresh freed flow rate is controlled by valve, vaporizer outlet temperature is controlled by hot utility, heat exchanger outlet temperature for cold stream is controlled by bypass valve on hot side, feed temperature of the product column is controlled by cold utility. At separation section, the DME purity is controlled by reflux valve, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by the distillate valve and level of reboiler is controlled by reboiler utility, the recycle column; column pressure is controlled by the cold utility of condenser, level of condenser is controlled by the hot utility of reboiler and level of reboiler is controlled by bottom valve. Recycle flow is control by recycle valve but column temperature does not control at this control structure.

5.3.1.3 Control Structure 3 (CS3) for Base Case of the DME Process

This control structure is followed Fixture point method that is shown in Figure 5.9. Control variable and manipulation item can be selected as follow; the fresh freed flow rate is controlled by valve, stream pressure before feed in vaporizer is controlled by cascade with fresh feed flow controlled, vaporizer outlet temperature is controlled by hot utility, heat exchanger outlet temperature for cold stream is controlled by bypass valve on hot side, feed temperature of the product column is controlled by cold utility. At separation section, the DME purity is controlled by reflux vavle, product column temperature at stage 20 is controlled by hot utility of reboiler, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by the distillate valve and level of reboiler is controlled by cascade with flow control to recycle column. The recycle column; Recycle column temperature at stage 24 is controlled by hot utility of reboiler, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by reflux valve and level of reboiler is controlled by bottom valve. Recycle flow is control by recycle valve.

5.3.1.4 Control Structure 4 (CS4) for Base Case of the DME Process

This control structure is followed Fixture point method that is shown in Figure 5.10. Control variable and manipulation item can be selected as follow; the fresh freed flow rate is controlled by valve, stream pressure before feed in vaporizer is controlled by cascade with fresh feed flow controlled, vaporizer outlet temperature is controlled by hot utility, heat exchanger outlet temperature for cold stream is controlled by bypass valve on hot side, feed temperature of the product column is controlled by cold utility. At separation section, the DME purity is controlled by reflux vavle, product column temperature at stage 20 is controlled by hot utility of reboiler, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by the distillate valve and level of reboiler is controlled by feed valve in product column. The recycle column; Recycle column temperature at stage 24 is controlled by bottom valve, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by reflux valve and level of reboiler is controlled by hot utility of reboiler. Recycle flow is control by recycle valve.

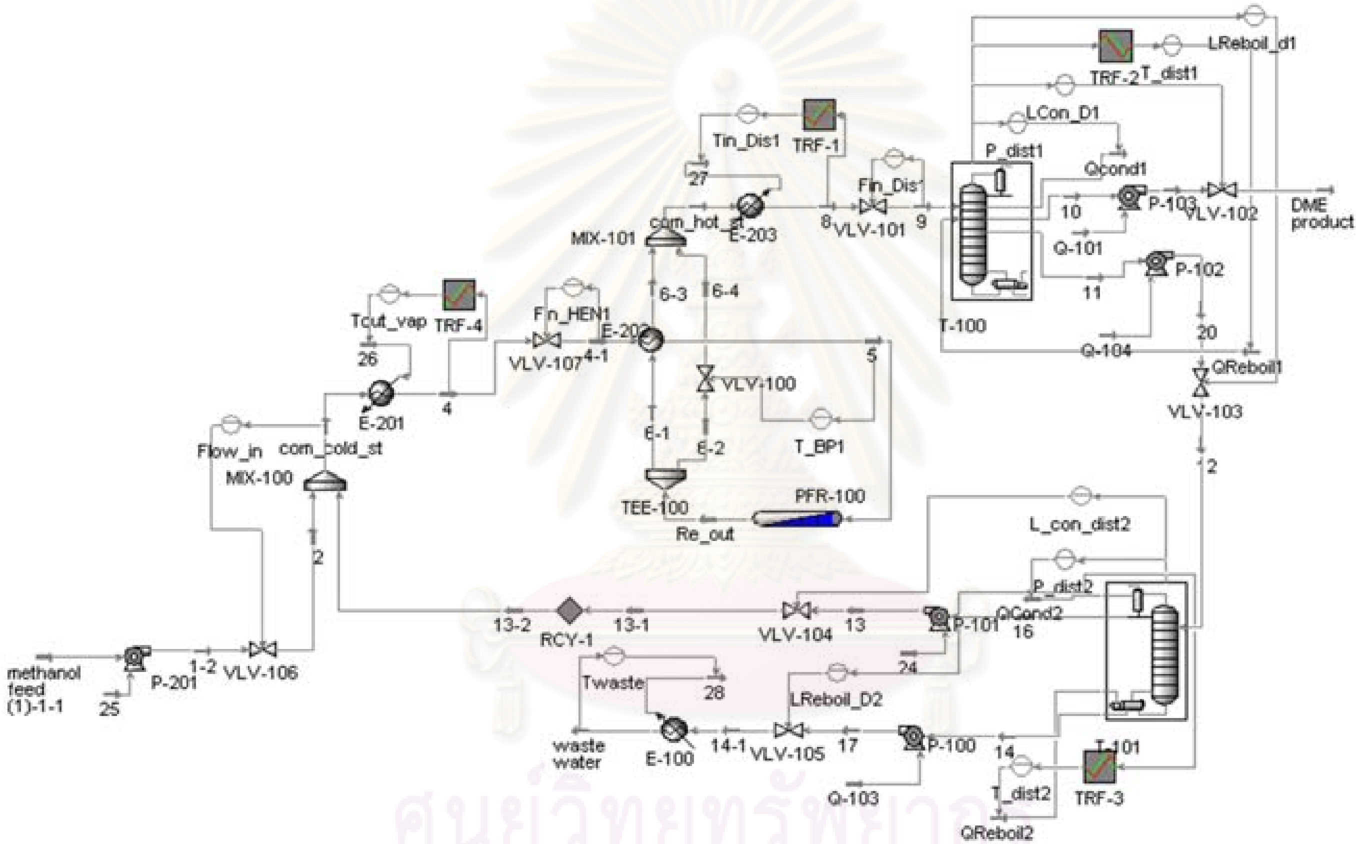

Figure 5.7 Application of control structure 1 (CS1) to the DME process base case (BC)

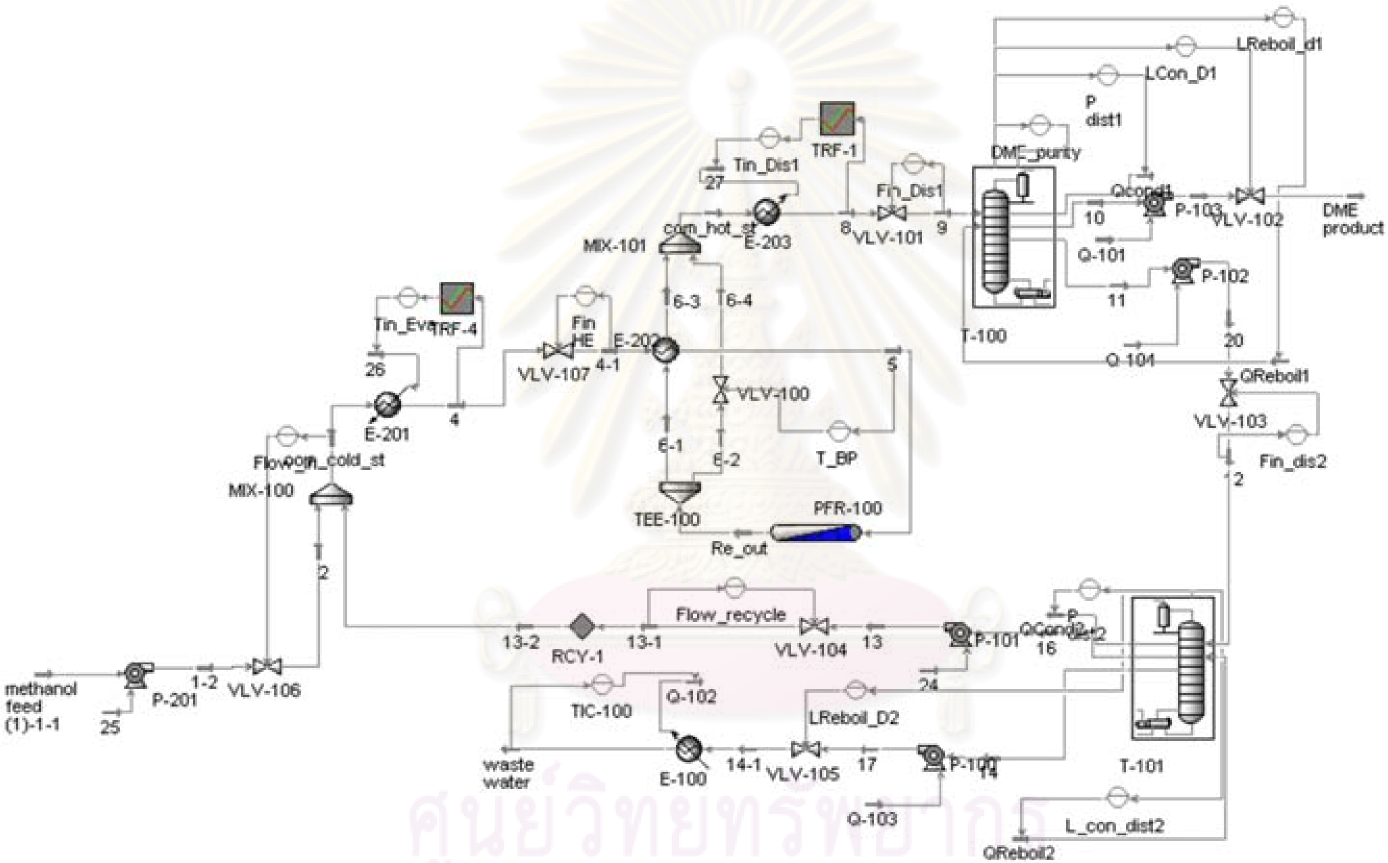

Figure 5.8 Application of control structure 2 (CS2) to the DME process base case (BC)

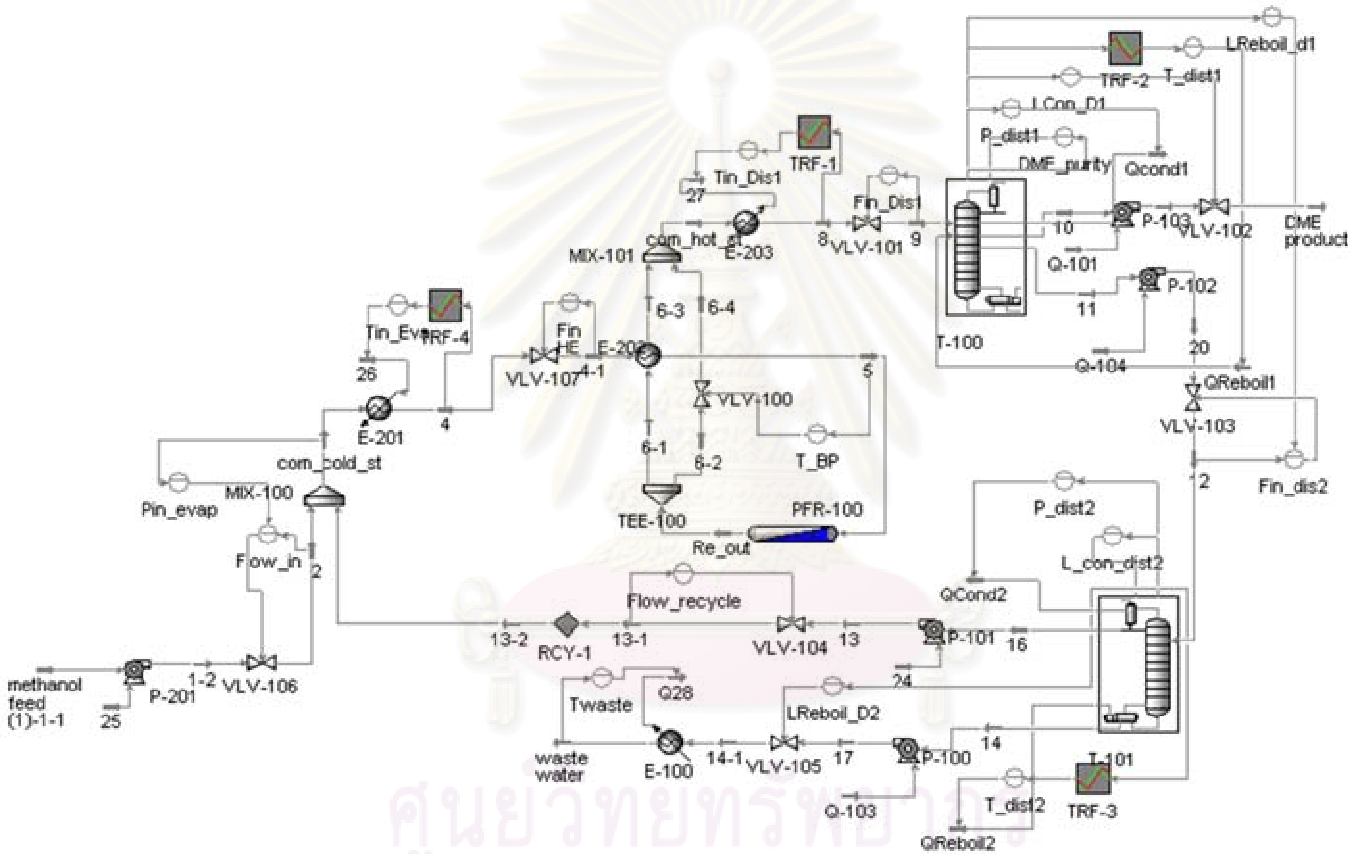

Figure 5.9 Application of control structure 3 (CS3) to the DME process base case (BC)

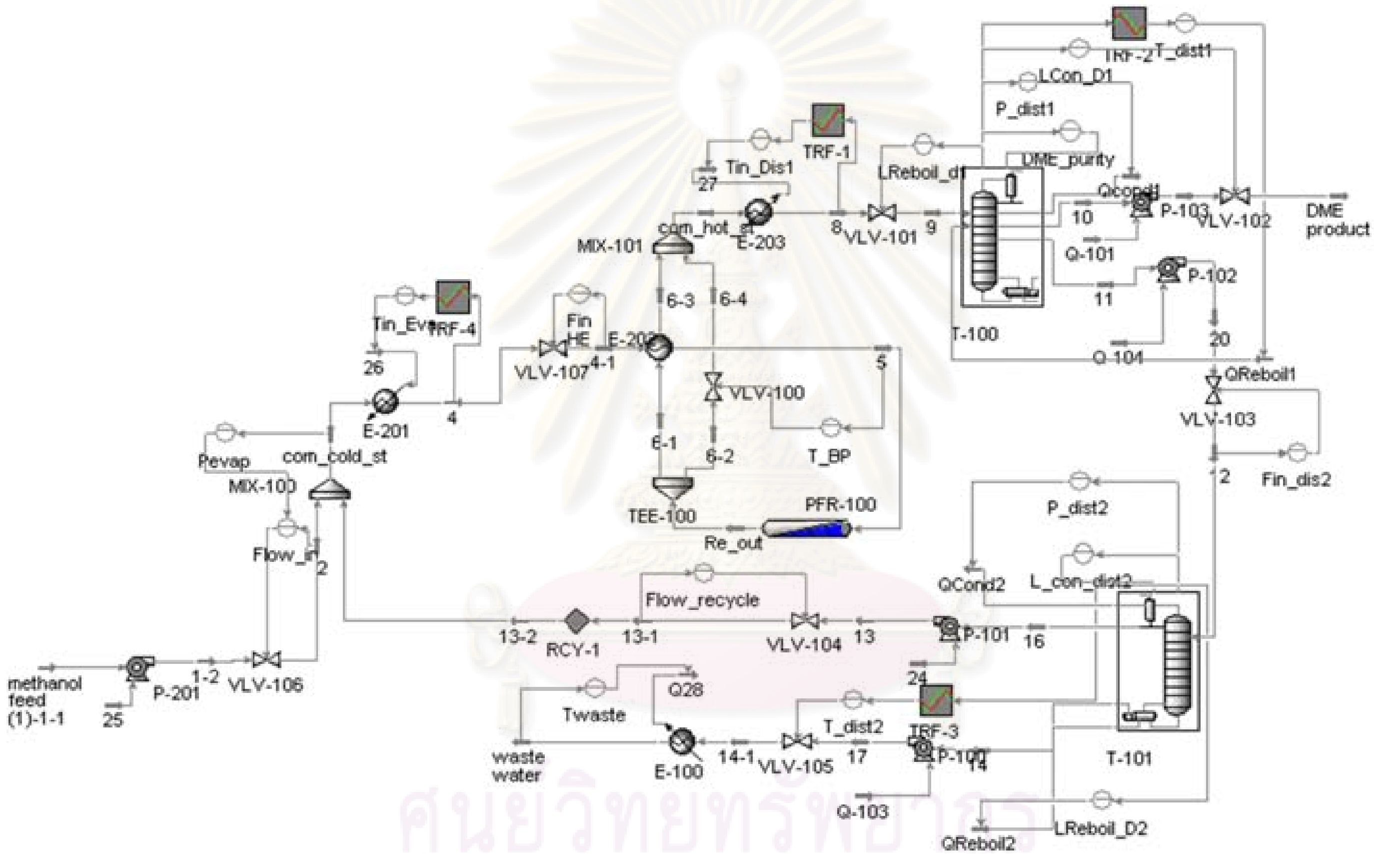

Figure 5.10 Application of control structure 4 (CS4) to the DME process base case (BC)

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5.3.1.5 Control Structure (CS1, CS2 CS3 and CS4) for the DME Process Alternative 1

This alternative show control structure 1 (CS1) in figure 5.11, control structure 2 (CS2) in figure 5.12, control structure 3 (CS3) in figure 5.13 and control structure 4 (CS4) in figure 5.14. The main control structure loop of CS1, CS2, CS3 and CS4 same as the base case that the fresh freed flow rate is controlled by valve, vaporizer outlet temperature is controlled by hot utility, heat exchanger outlet temperature for cold stream is controlled by bypass valve on hot side, but the alternative 1 added the heat exchanger network to integrate the reactor effluence stream for hot stream (H1) with the vaporizer feed stream for cold stream (C1). Temperature is controlled at hot stream outlet (product column feed temperature) by manipulating bypass valve of hot stream so this alternative not necessary to add cooler.

The separation loop control is the same as the base case that the product column; pressure is controlled by cold utility of condenser, temperature is controlled by hot utility of reboiler except CS2 is not controlled product column temperature, level of condenser is controlled by the distillate valve and level of reboiler is controlled by bottom valve for CS1, CS2 and control by cascade with flow control to recycle column for CS3 and control with column feed valve for CS4, DME purity is controlled by reflux valve of product column for CS2, CS3, CS4 but not controlled in CS1.The recycle column; temperature is controlled by hot utility of reboiler for CS1, CS2, CS3 but controlled by bottom valve for CS4, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by recycle valve for CS1, control by hot utility of reboiler for CS2 and controlled by reflux valve for CS3 and CS4, level of reboiler is controlled by bottom valve for CS1, CS2, CS3 and control by hot utility of reboiler for CS4.

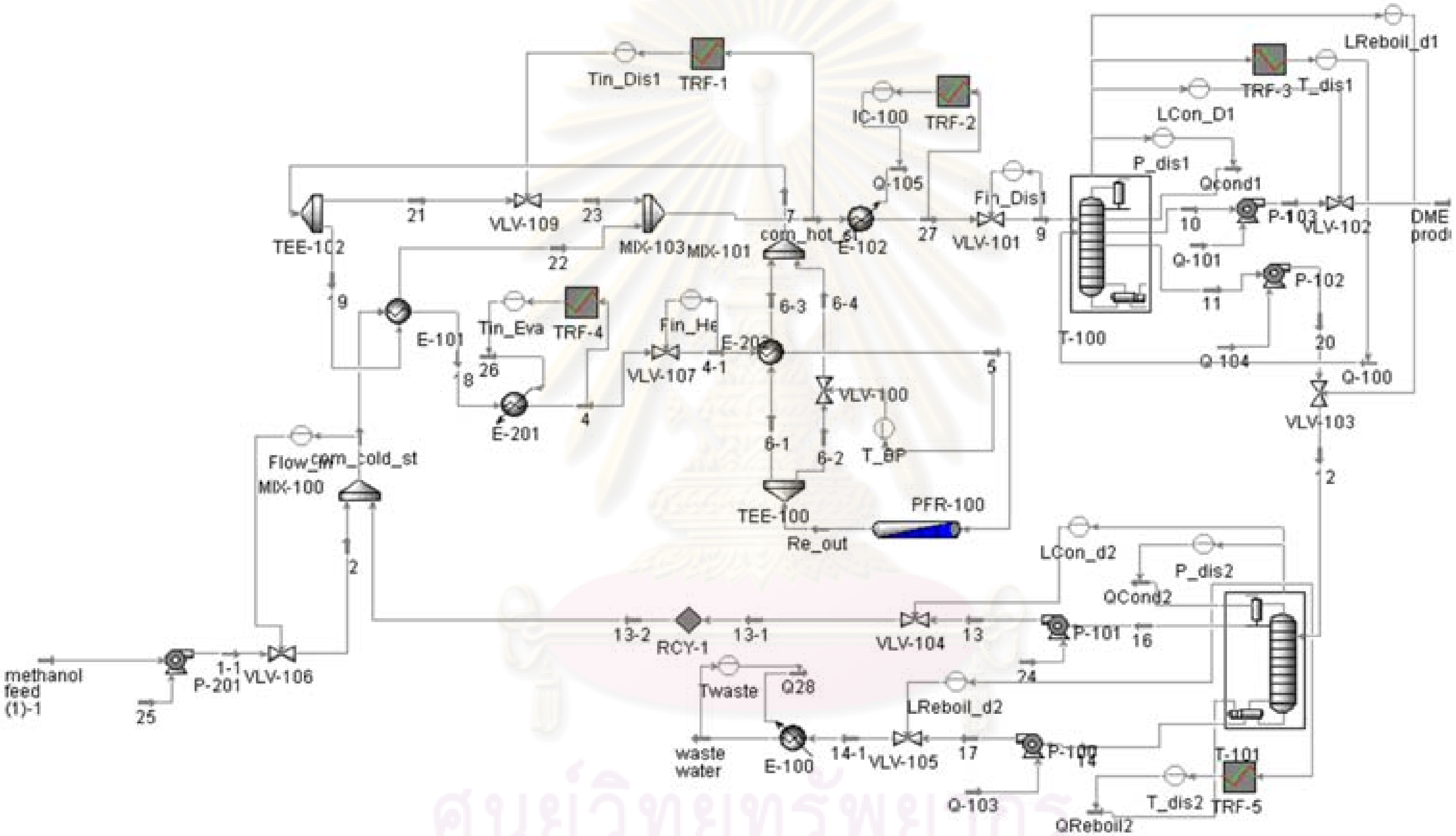

Figure 5.11 Application of control structure 1 (CS1) to the DME process alternative 1

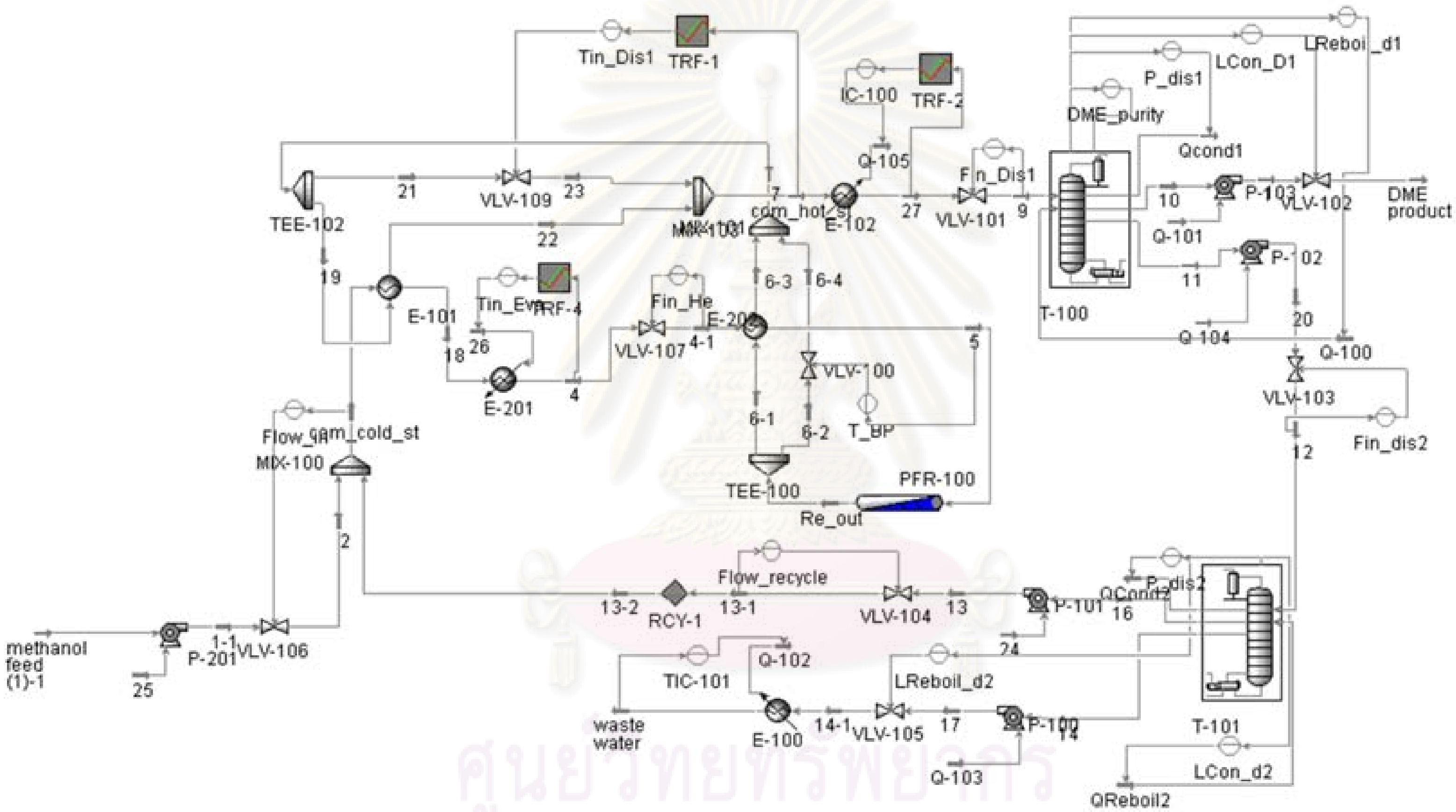

Figure 5.12 Application of control structure 2 (CS2) to the DME process alternative 1

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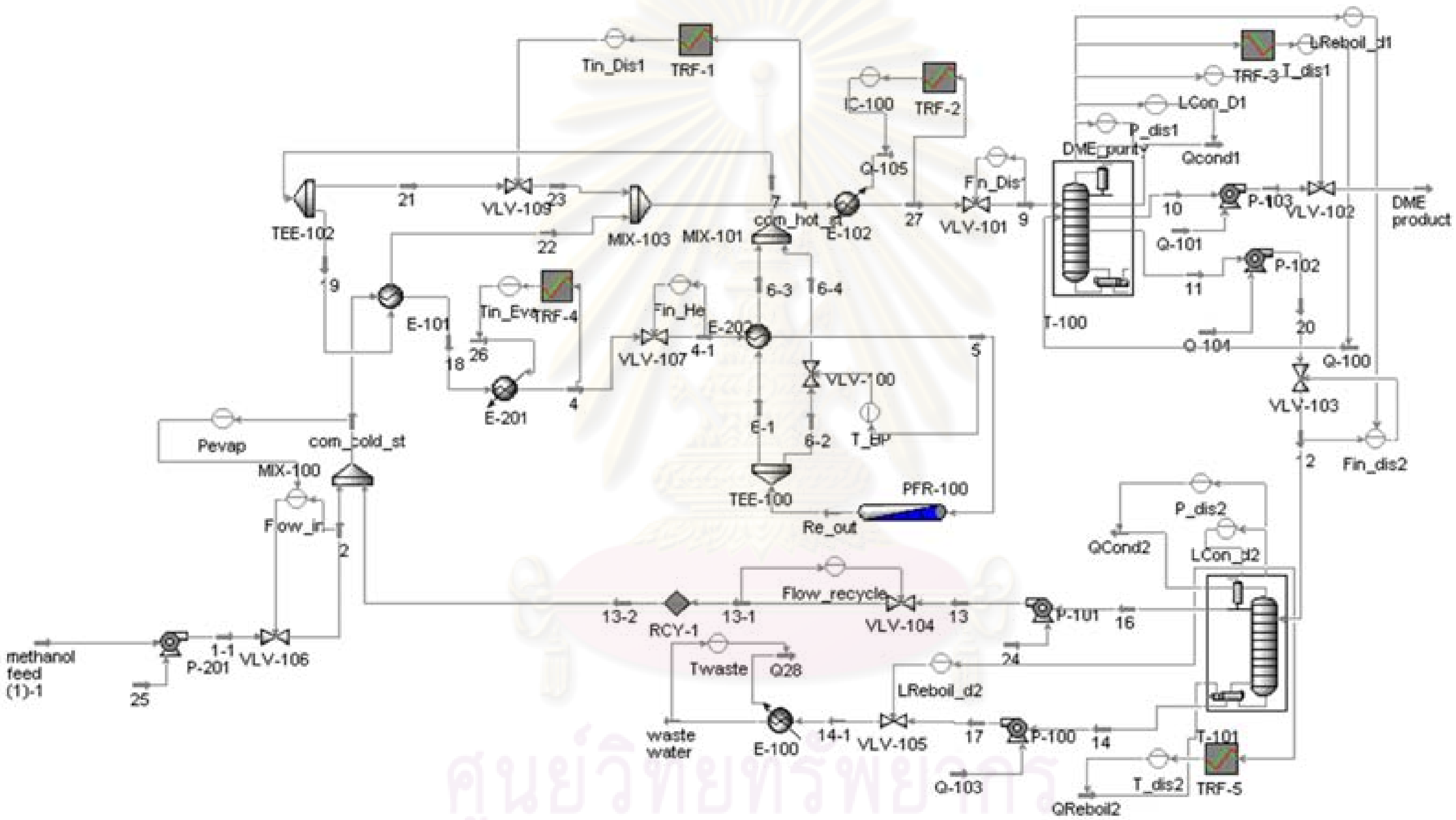

Figure 5.13 Application of control structure 3 (CS3) to the DME process alternative 1

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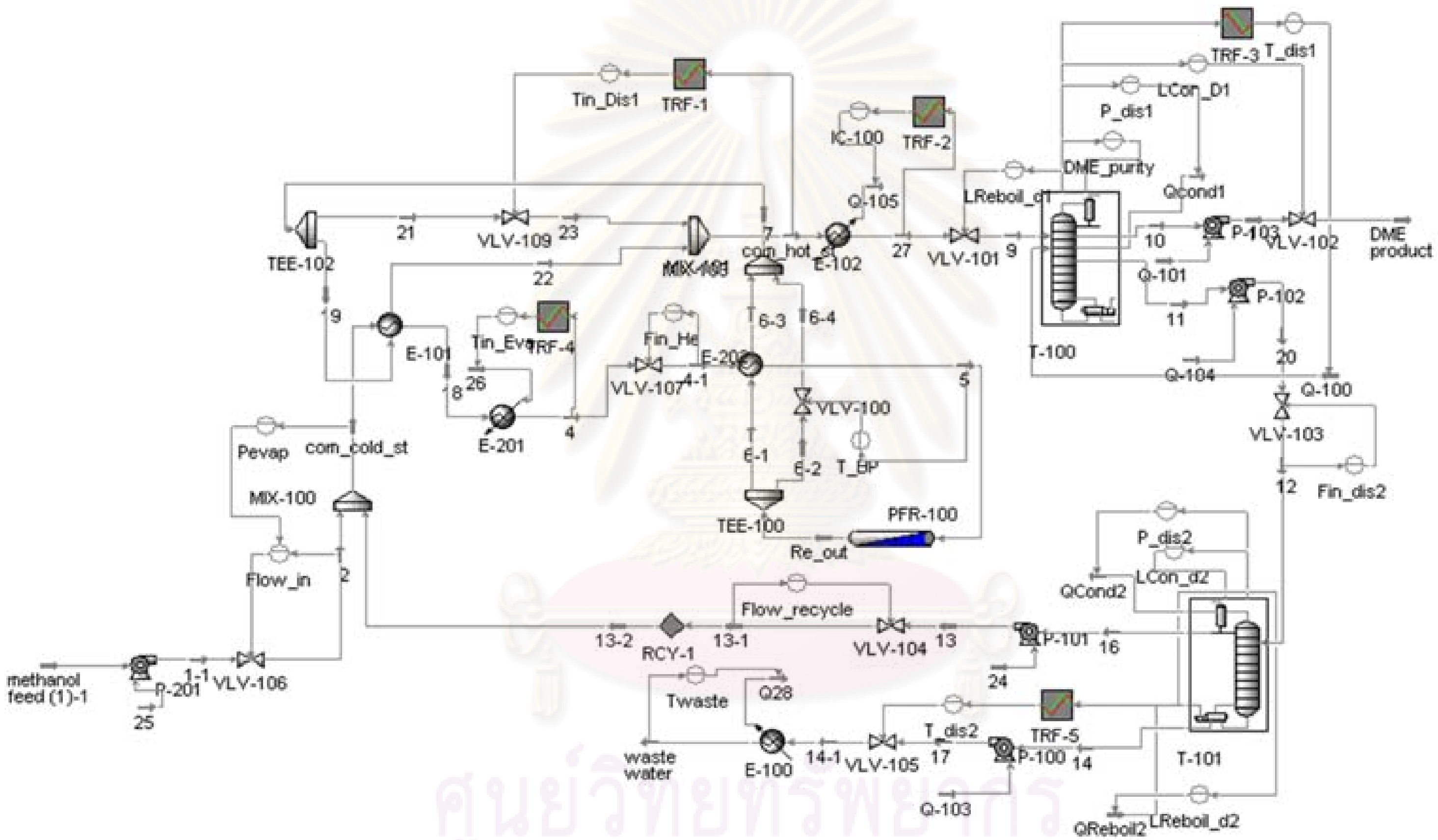

Figure 5.14 Application of control structure 4 (CS4) to the DME process alternative 1

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5.3.1.6 Control Structure (CS1, CS2 CS3 and CS4) for the DME Process Alternative 2

This alternative shows control structure 1 (CS1) in Figure 5.15, control structure 2 (CS2) in figure 5.16, control structure 3 (CS3) in figure 5.17 and control structure 4 (CS4) in figure 5.18. The main control structure loop of CS1, CS2, CS3 and CS4 is the same as the base case that the fresh freed flow rate is controlled by valve, pressure of stream inlet vaporizer is control by cascade with flow inlet control for CS3 and CS4, vaporizer outlet temperature is controlled by hot utility, heat exchanger outlet temperature for cold stream is controlled by bypass valve on hot side, product column inlet temperature is control by cooler utility but the alternative 2 added the heat exchanger network to integrate the reactor effluence stream for hot stream (H1) the product column reboiler for cold stream (C2). The temperature is controlled at heat exchanger out of cold stream (product column temperature) by manipulating bypass valve of hot stream so this alternative not necessary to add rebolier at product column.

The separation loop control is the same as the base case that in the product column; pressure is controlled by cold utility of condenser, temperature is controlled by bypass valve adjusting except CS2 is not controlled product column temperature, level of condenser is controlled by the distillate valve and level of reboiler is controlled by bottom valve for CS1, CS2 and control by cascade with flow control to recycle column for CS3 and control with column feed valve for CS4, DME purity is controlled by reflux valve of product column for CS2, CS3, CS4 but not controlled in CS1.The recycle column; temperature is controlled by hot utility of reboiler for CS1, CS2, CS3 but controlled by bottom valve for CS4, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by recycle valve for CS1, control by hot utility of reboiler for CS2 and controlled by reflux valve for CS3 and CS4, level of reboiler is controlled by bottom valve for CS1, CS2, CS3 and control by hot utility of reboiler for CS4.

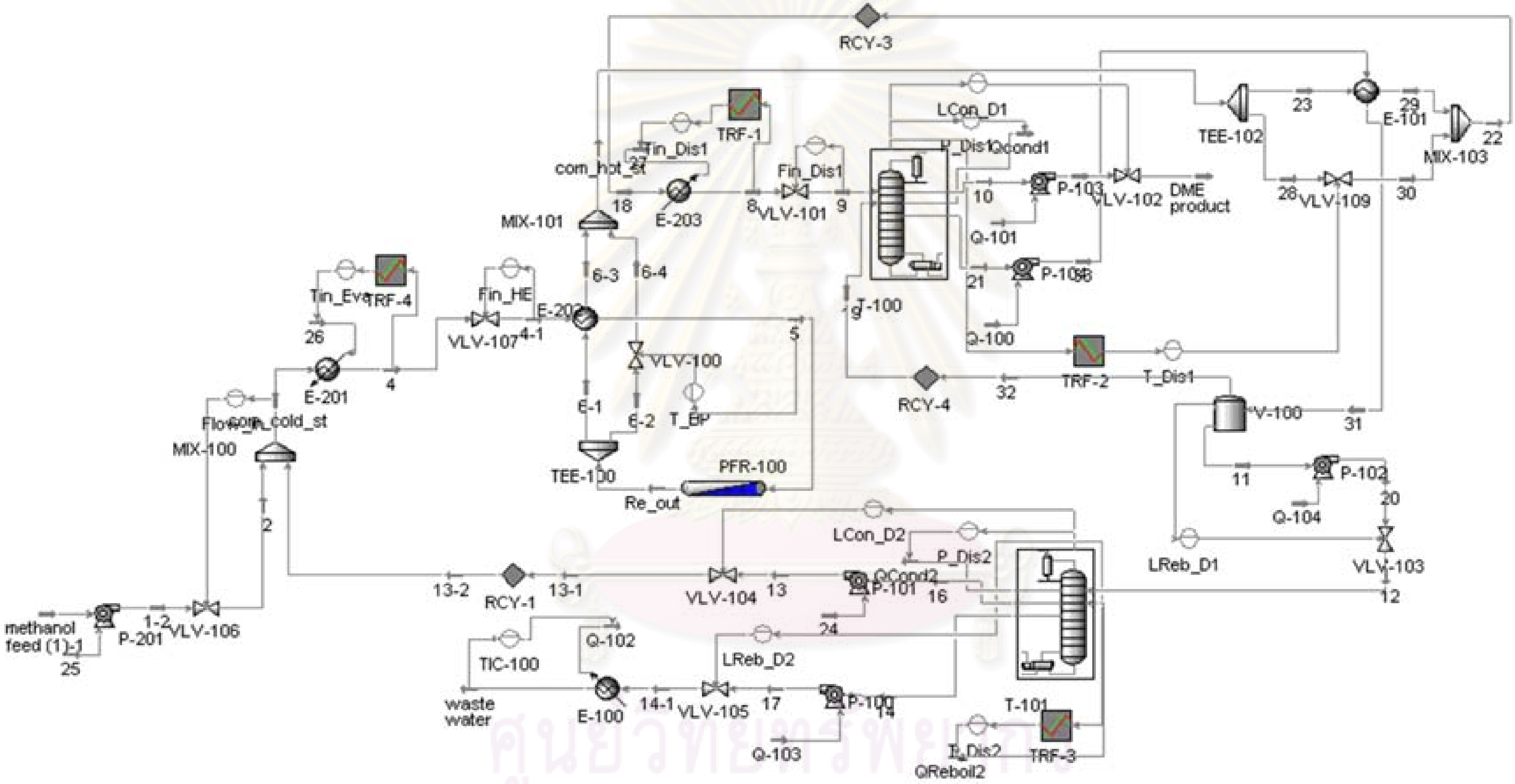

Figure 5.15 Application of control structure 1 (CS1) to the DME process alternative 2

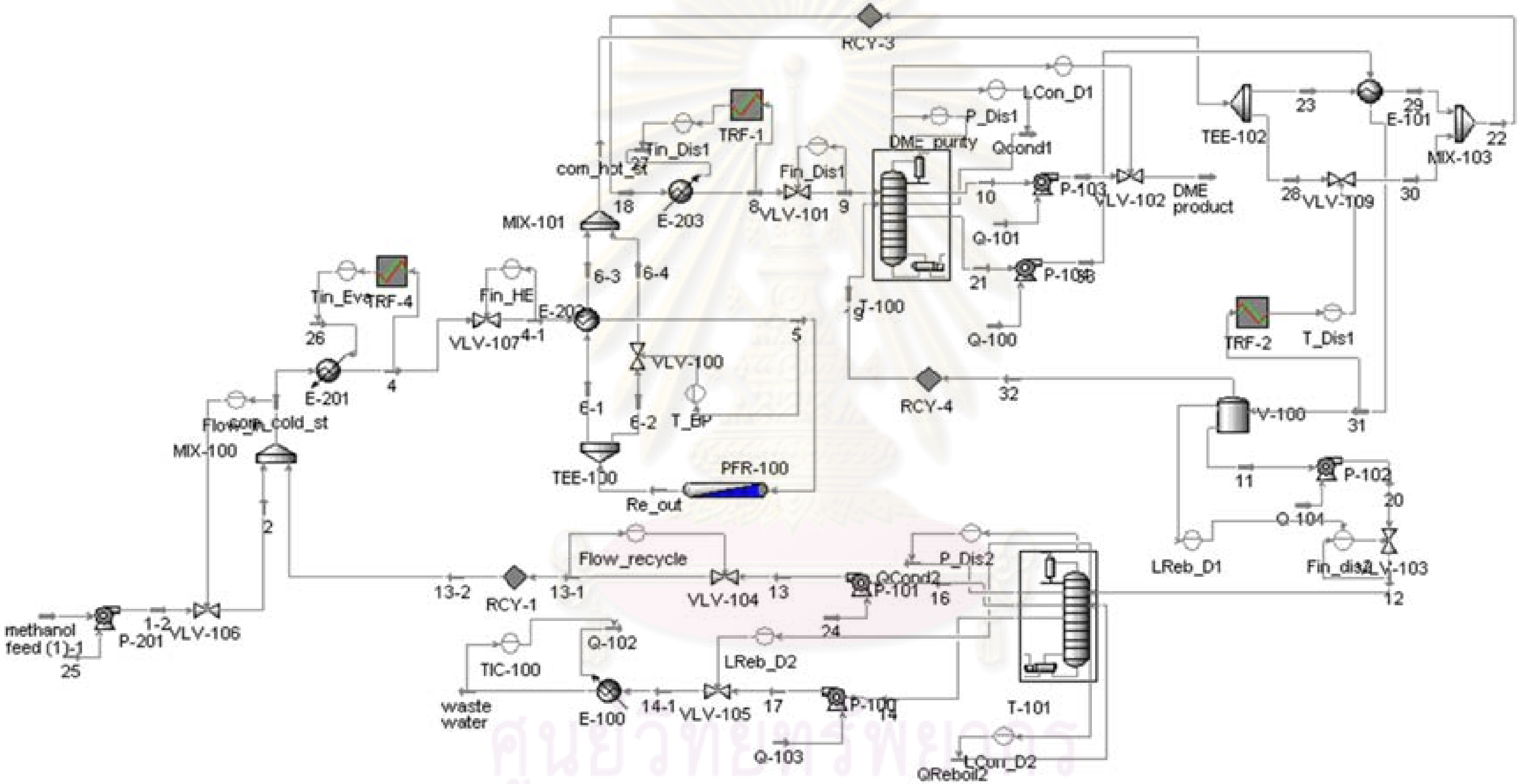

Figure 5.16 Application of control structure 2 (CS2) to the DME process alternative 2

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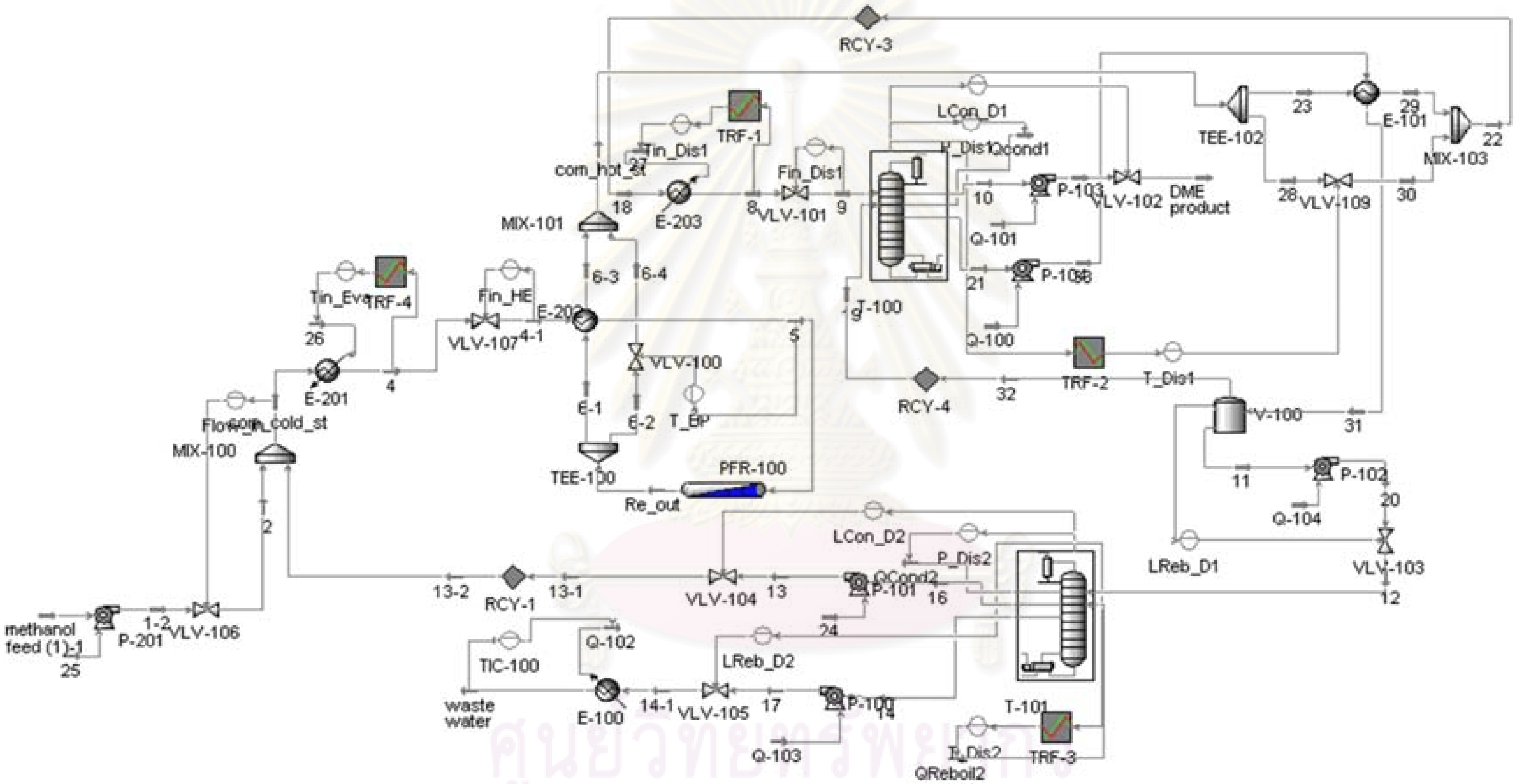

Figure 5.17 Application of control structure 3 (CS3) to the DME process alternative 2

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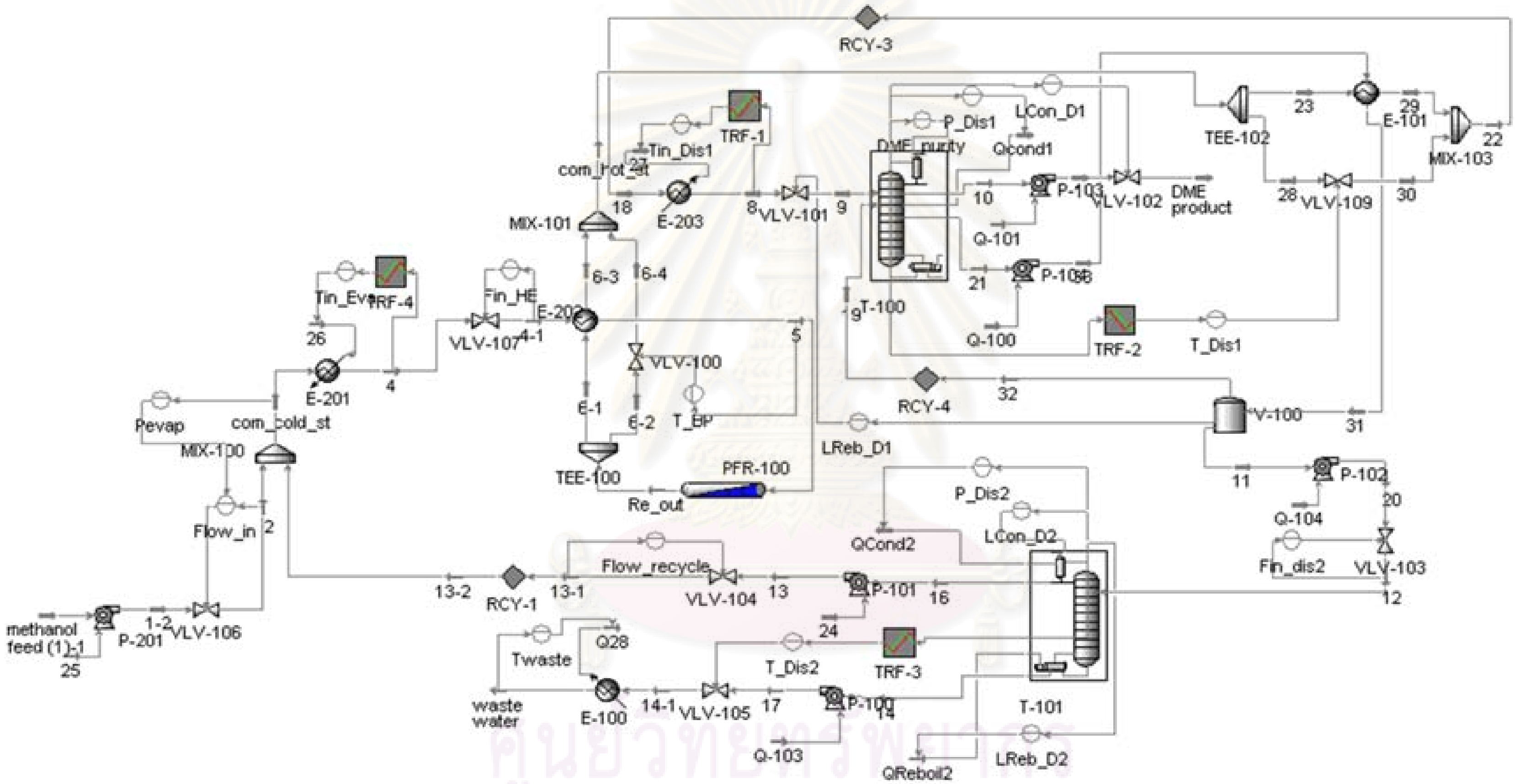

Figure 5.18 Application of control structure 4 (CS4) to the DME process alternative 2

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5.3.1.7 Control Structure (CS1, CS2 CS3 and CS4) for the DME Process Alternative 3

This alternative shows control structure 1 (CS1) in figure 5.19, control structure 2 (CS2) in figure 5.20, control structure 3 (CS3) in figure 5.21 and control structure 4 (CS4) in figure 5.22. The main control structure loop of CS1, CS2, CS3 and CS4 is the same as the base case that the fresh freed flow rate is controlled by valve, pressure of stream inlet vaporizer is control by cascade with flow inlet control for CS3 and CS4, vaporizer outlet temperature is controlled by hot utility, heat exchanger outlet temperature for cold stream is controlled by bypass valve on hot side, but alternative 3 is increased two heat exchanger network. First, integrate the reactor effluence stream for hot stream (H1) with the product column reboiler for cold stream (C2). The temperature is controlled at product column by manipulating bypass valve of hot stream. Second, integrate the reactor effluence stream for hot stream (H1) with the vaporizer feed stream for cold stream (C1). Temperature is controlled at hot stream outlet by manipulating bypass valve of hot stream.

The separation loop control is the same as the base case that in the product column; pressure is controlled by cold utility of condenser, temperature is controlled by bypass valve adjusting except CS2 is not controlled product column temperature, level of condenser is controlled by the distillate valve and level of reboiler is controlled by bottom valve for CS1, CS2 and control by cascade with flow control to recycle column for CS3 and control with column feed valve for CS4, DME purity is controlled by reflux valve of product column for CS2, CS3, CS4 but not controlled in CS1.The recycle column; temperature is controlled by hot utility of reboiler for CS1, CS2, CS3 but controlled by bottom valve for CS4, column pressure is controlled by the cold utility of condenser, level of condenser is controlled by recycle valve for CS1, control by hot utility of reboiler for CS2 and controlled by reflux valve for CS3 and CS4, level of reboiler is controlled by bottom valve for CS1, CS2, CS3 and control by hot utility of reboiler for CS4.

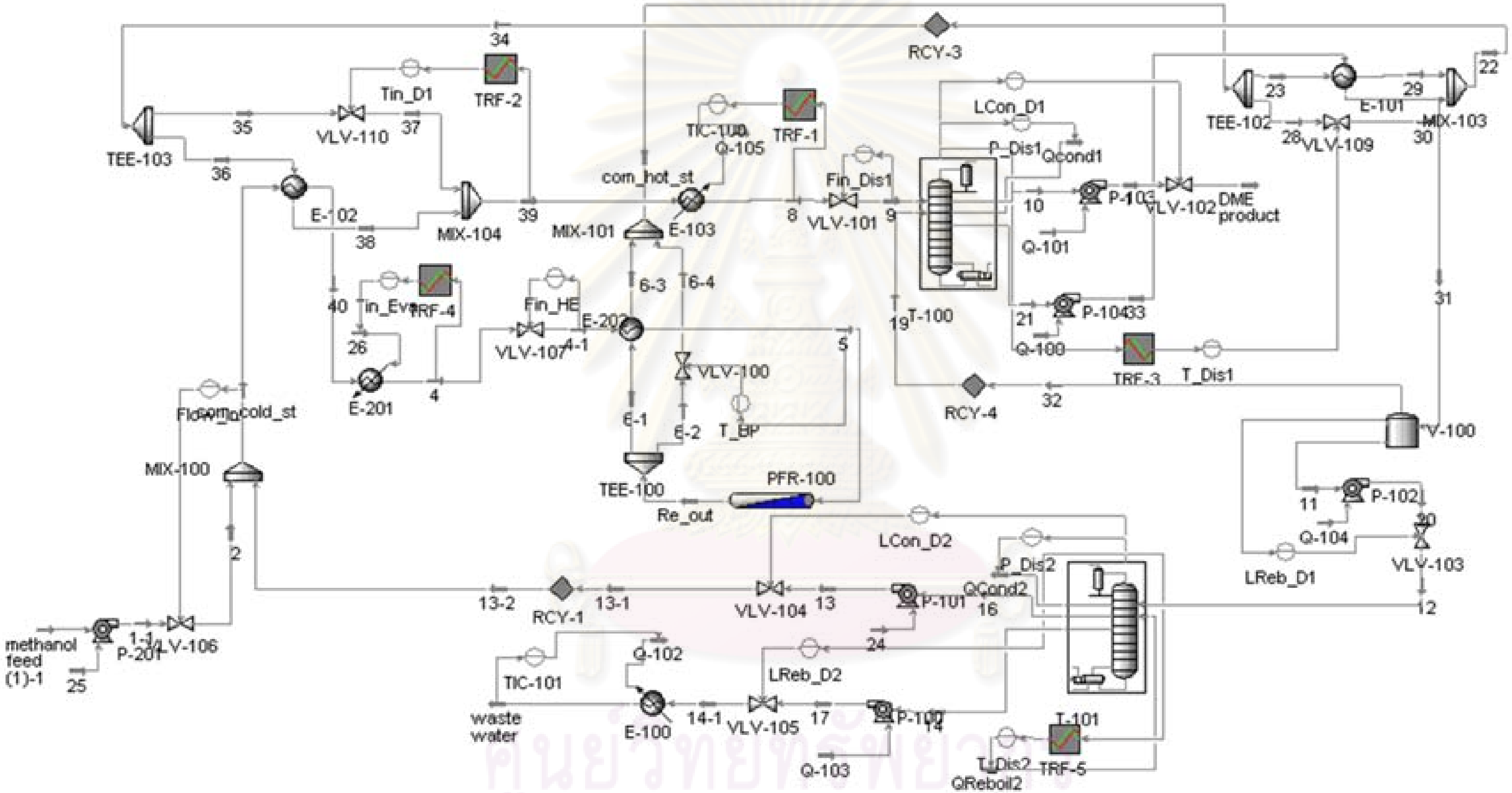

Figure 5.19 Application of control structure 1 (CS1) to the DME process alternative 3

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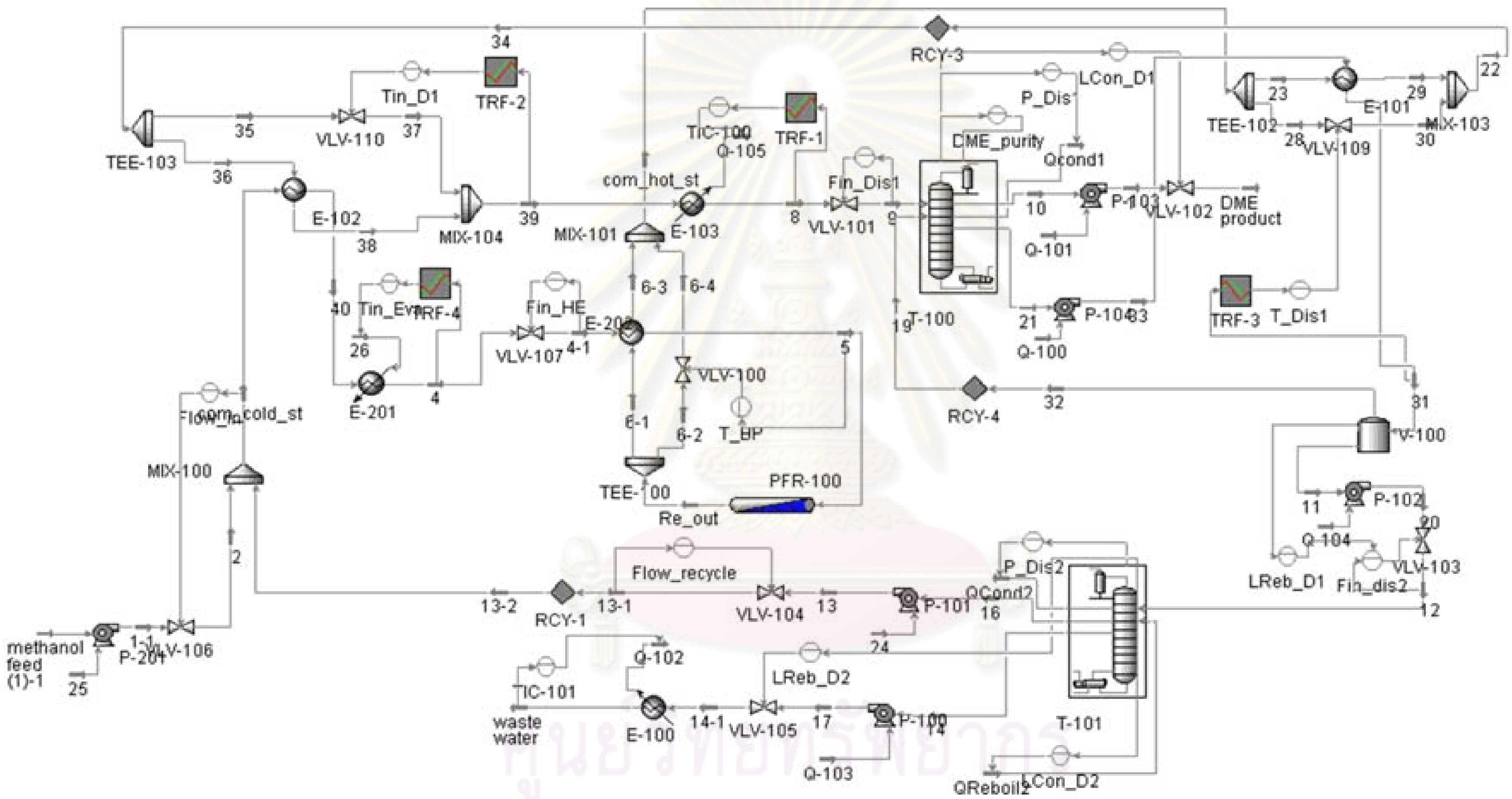

Figure 5.20 Application of control structure 2 (CS2) to the DME process alternative 3

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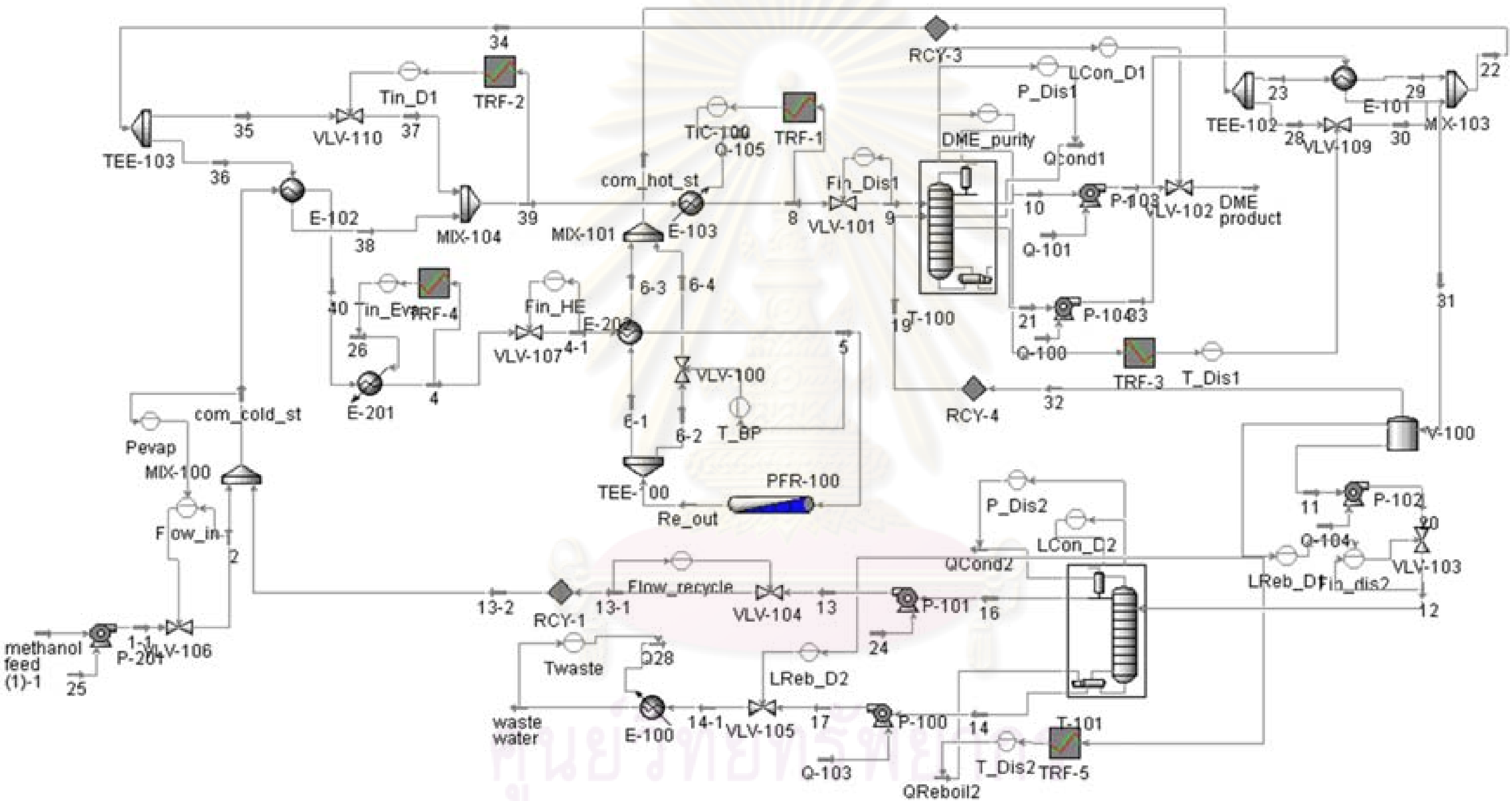

Figure 5.21 Application of control structure 3 (CS3) to the DME process alternative 3

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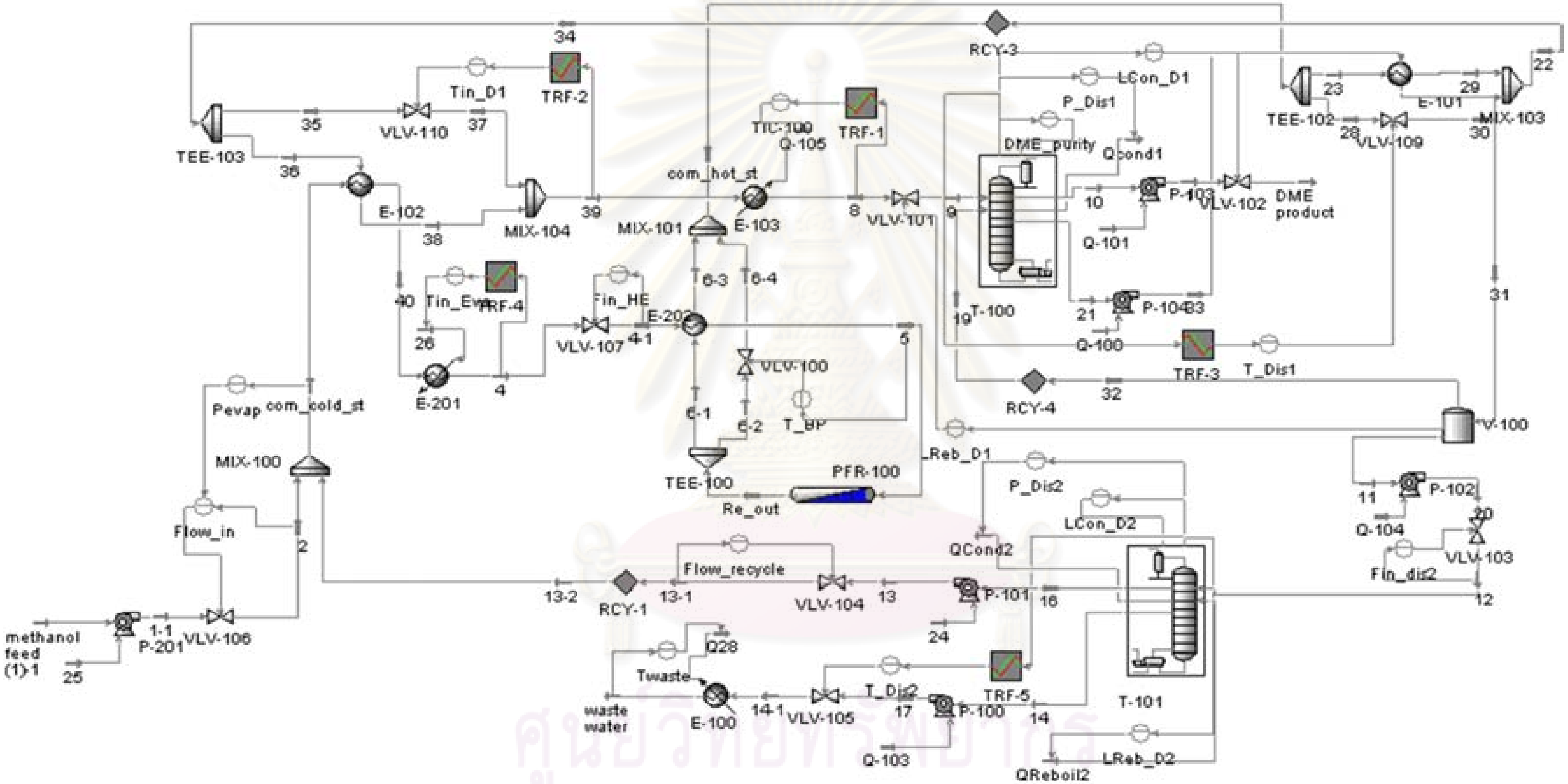

Figure 5.22 Application of control structure 4 (CS4) to the DME process alternative 3

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5.4 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 flow rate. Three disturbance loads are used to evaluate the dynamic performance of the base control structure (CS1) is provided by Turton (1998) and new control structures (CS2, CS3, and CS4) for the DME process.

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

In order to illustrate the dynamic behavior of the control structure in the DME process alternatives several disturbance loads are made. The dynamic results are explained in this part.

5.4.1 Dynamic Simulation Results of the DME Process Base case

Three disturbance loads are used to evaluate the dynamic performance of the base control structure CS1 and design control structures from Luyben's heuristic CS2 and fixture point method CS3 and CS4 for DME process base case.

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

Figures 5.23, 5.26, 5.29 and 5.32 show the dynamic responses of the control systems of DME process for base case to a change in the heat load disturbance of cold stream (reactor feed stream). In order to make this disturbance, first the temperature of combine stream between feed stream and recycle is decreased from 46.4^0C to 41.4^0C occurring at time equals 10 minutes, and the temperature is increased from 41.4^0C to 51.4^0C occurring at time equals 200 minutes, then its temperature is returned to its nominal value of 41.4^0C occurring at time equals 400 minutes (Figures 5.23a, 5.26a, 5.29a, and 5.32a).

All control structures including base control structure CS1 and new control structure CS2, CS3 and CS4 give the same result to reject disturbance by vaporizer to adjust duty as cold negative disturbance is effected to vaporizer duty decrease and cold positive disturbance is effected to vaporizer duty increase, shown in figures 5.23h, 5.26h, 5.29h, and 5.32h. Not only that heat load disturbance of cold stream make vapor fraction at vaporizer outlet not equal one at short period so effect to FEHE because this FEHE exchange heat of streams in vapor phase, however it can be adjusted by bypass valve to control temperature outlet of cold stream as shown in figure 5.23c, 5.26c, 5.29c, and 5.32c. This disturbance dose not affect product purity and a little affect production flow.

As can be see, this disturbance load has a little bit effect to the tray temperatures in the product column and product purity is slightly well however when consider outlet temperature of FEHE of cold stream CS3 and CS4 is better than CS1, CS2 because CS3 and CS4 control pressure of steam before feed to FEHE, it makes temperature control smoothly at vaporizer outlet.

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

Figures 5.24, 5.27, 5.30 and 5.33 show the dynamic responses of the control systems of DME process of Base Case to a change in the heat load disturbance of hot stream (Reactor Product stream).

In order to make this disturbance, first the temperature of the reactor outlet is decreased from 365.7^0C to 360.7^0C occurring at time equals 10 minutes, and the temperature is increased from 360.7^0C to 370.7^0C occurring at time equals 200 minutes, then its temperature is returned to its nominal value of 365.7^0C occurring at time equals 400 minutes (Figures 5.24a, 5.27a, 5.30a, and 5.33a). The disturbances load of the hot stream are direct send to cooler because cold temperature is controlled by valve bypass so the disturbance can be reject by cooler duty. The heat load disturbance from the hot stream make effect smaller than heat load disturbance from the cold stream because no phase change in FHEE as shown in figures 5.24c, 5.27c, 5.30c, and 5.33c. The dynamic responses of product purity are slightly well and have a little bit effect to product ion flow rate. However if we consider the product column temperature, cold stream outlet temperature of FEHE the dynamic response of CS3 is smoother than CS1, CS2 and CS4.

5.4.1.3 Change in the Flow rates of Main Process stream

Figures 5.25, 5.28, 5.31 and 5.34 show the dynamic responses of the control systems of DME process base case to a change in the flow rates of main process stream. This disturbance is made by decreasing flow rates from 328.5 kgmol/h to 323.5 kgmol/h occurring at time equals 10 minutes, and the flow rates is increased from 323.5 kgmol/h to 333.5 kgmol/h occurring at time equals 100 minutes, then its flow rates is returned to its nominal value of 328.5 kgmol/h occurring at time equals 200 minutes (Figures 5.25a, 5.28a, 5.31a and 5.34a).

The dynamic result can be seen that the drop in flow rates reduces the reaction rate, so the production rate is decreased as shown in figures 5.25g, 5.28g, 5.31g and 5.34g but the DME product purity is opposite responded to increase when flow rates decrease as shown in figures 5.25f, 5.28f, 5.31f and 5.34f. The disturbance makes the large effect of the product column temperature but can be controlled to set point value. The CS3 can control the temperature of product column better than CS1, CS2 and CS3 so deviation of DME product quality from nominal value in CS4 smoother than CS1, CS2 and CS3.

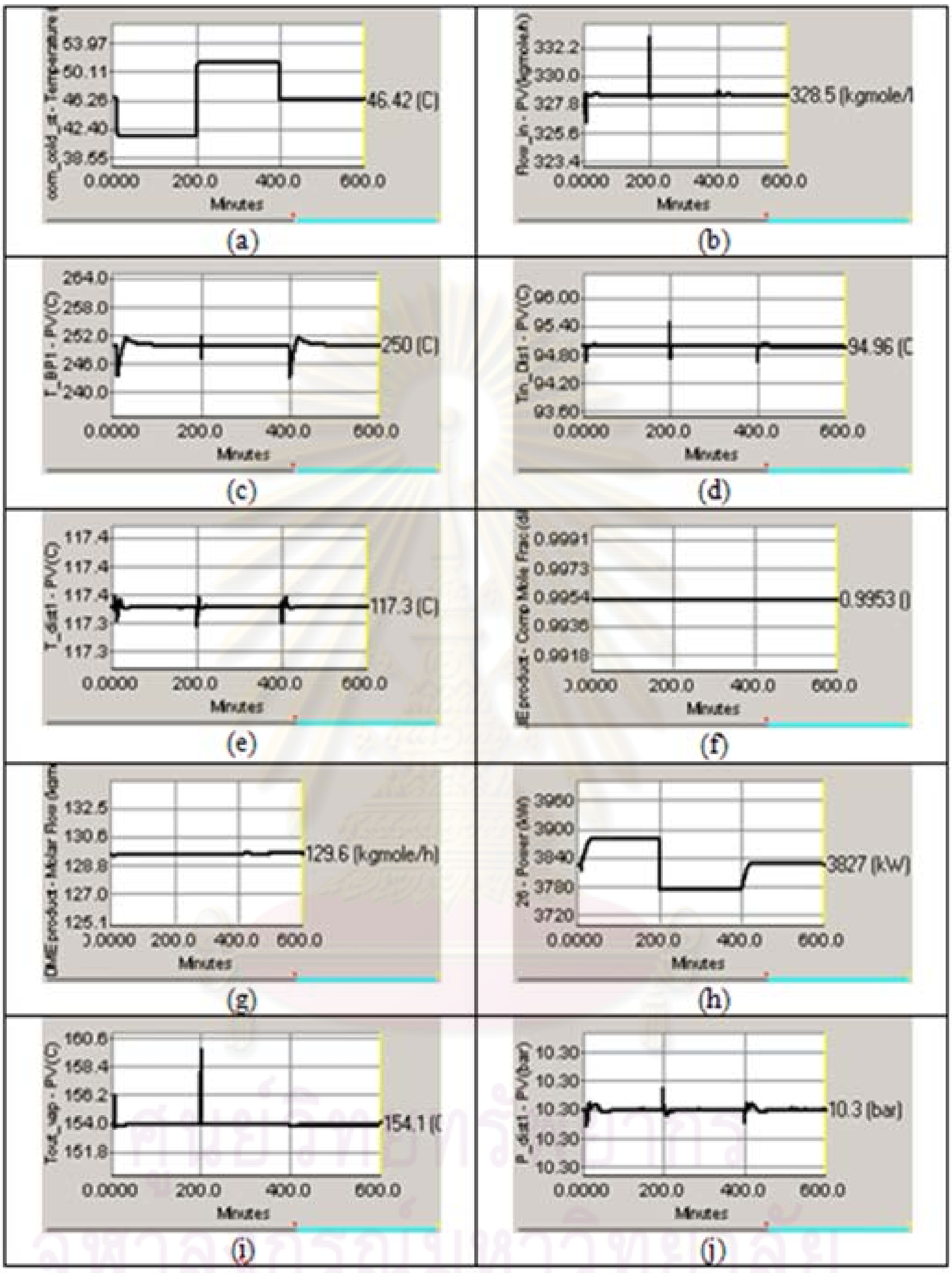

Figure 5.23 Dynamic Responses of the DME Process Base Case to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

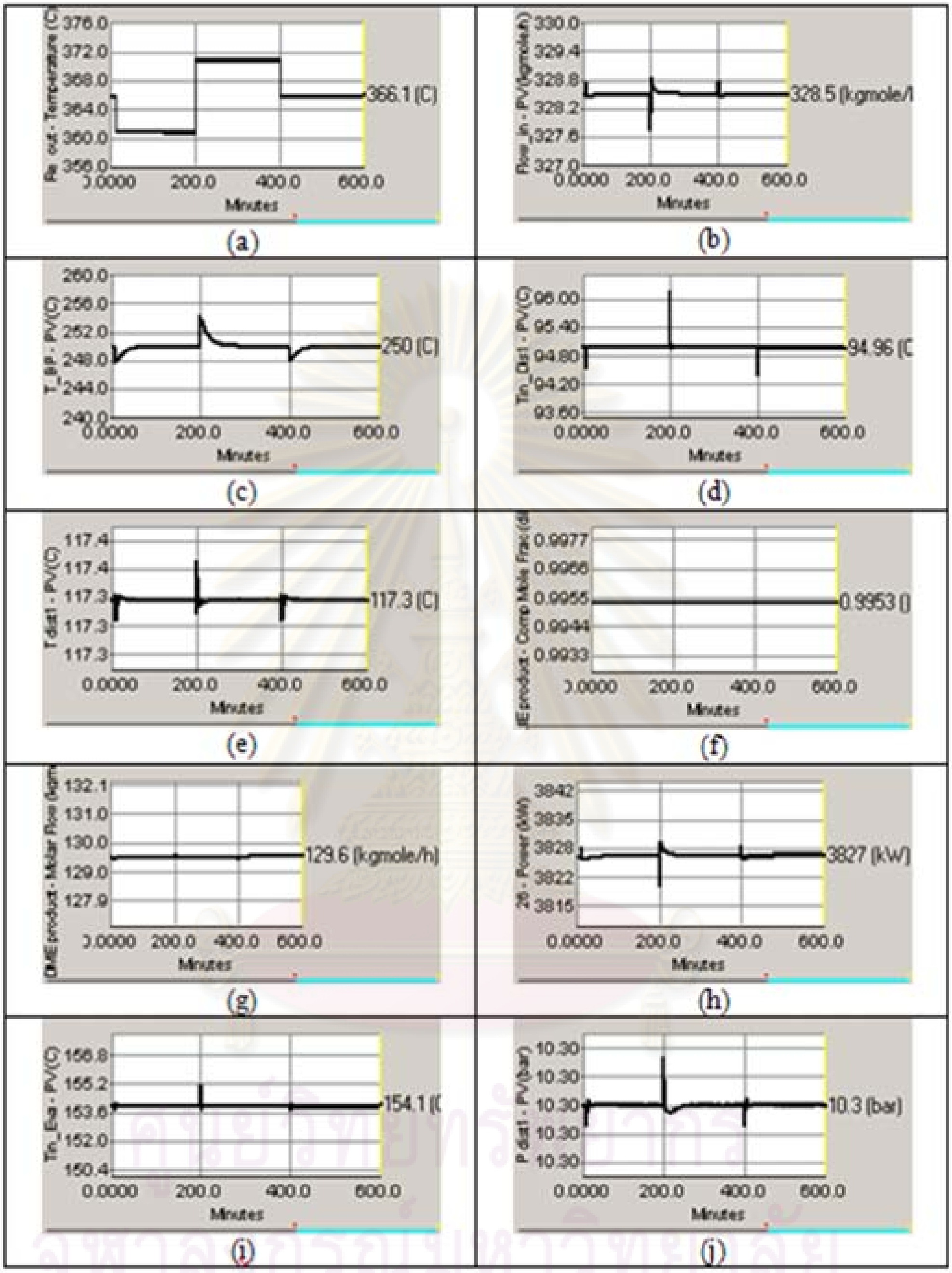

Figure 5.24 Dynamic Responses of the DME Process Base Case to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS1, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

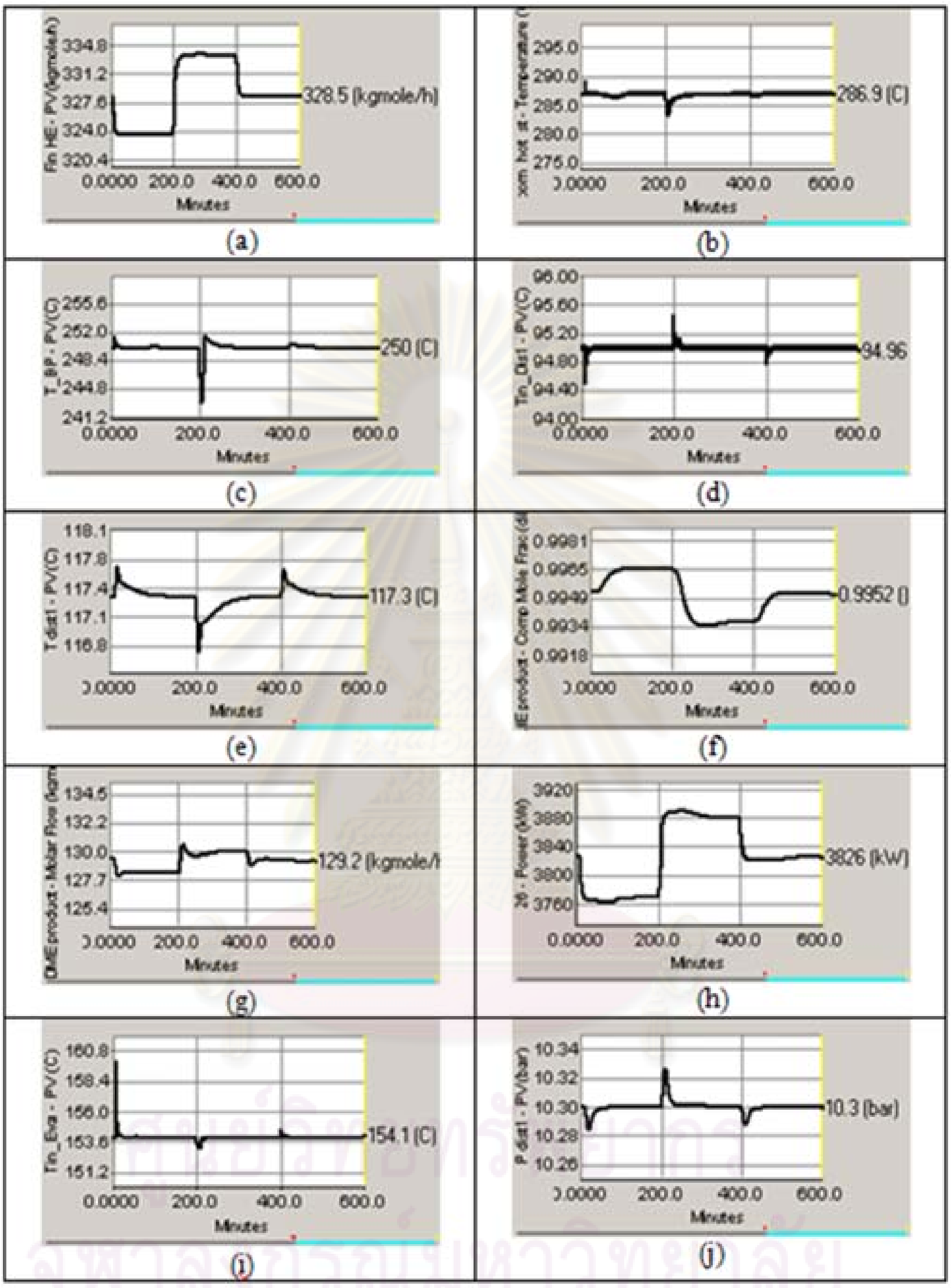

Figure 5.25 Dynamic Responses of the DME Process Base Case to Change the flow rate Disturbance of process stream: CS1, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

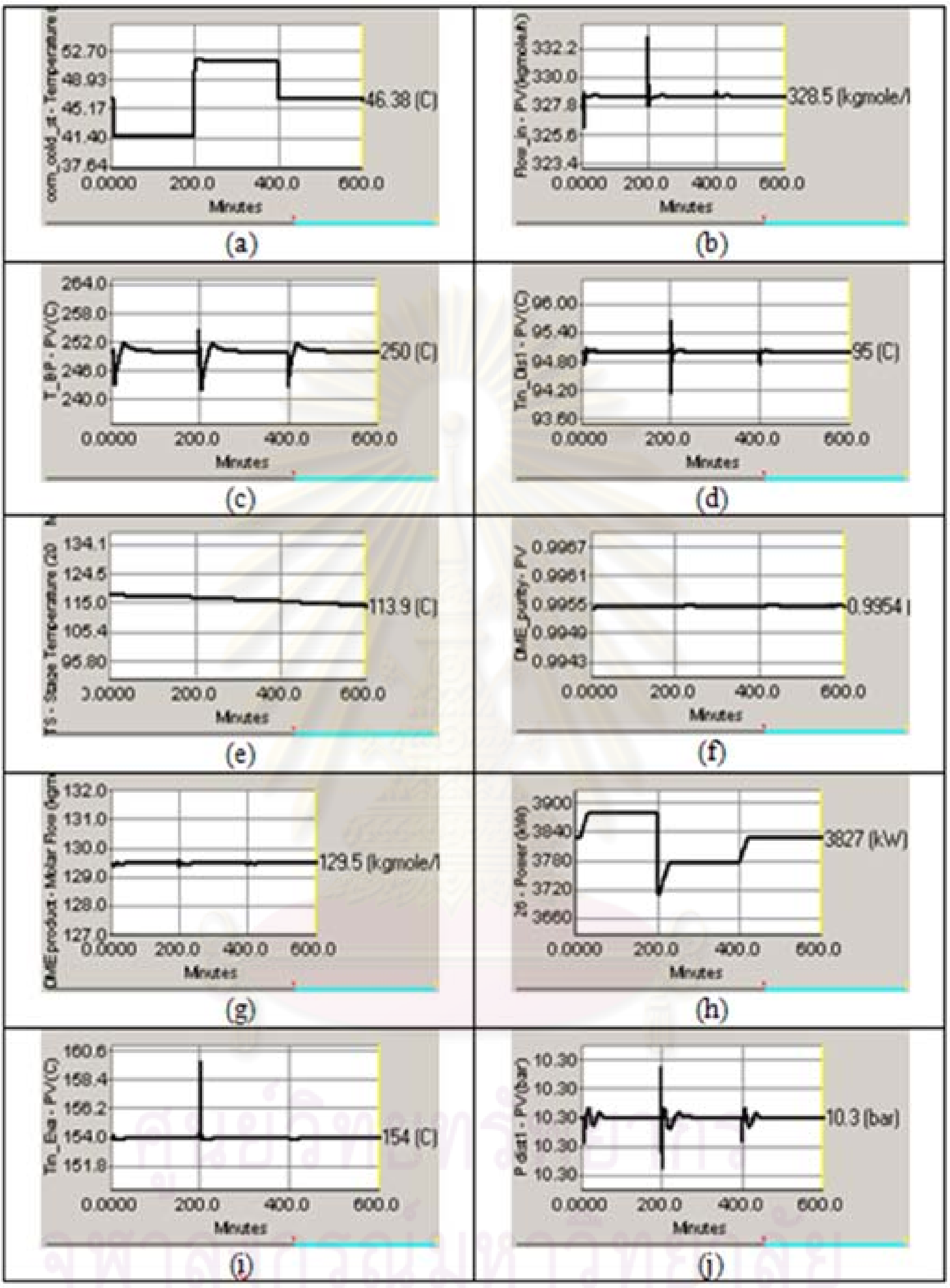

Figure 5.26 Dynamic Responses of the DME Process Base Case to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

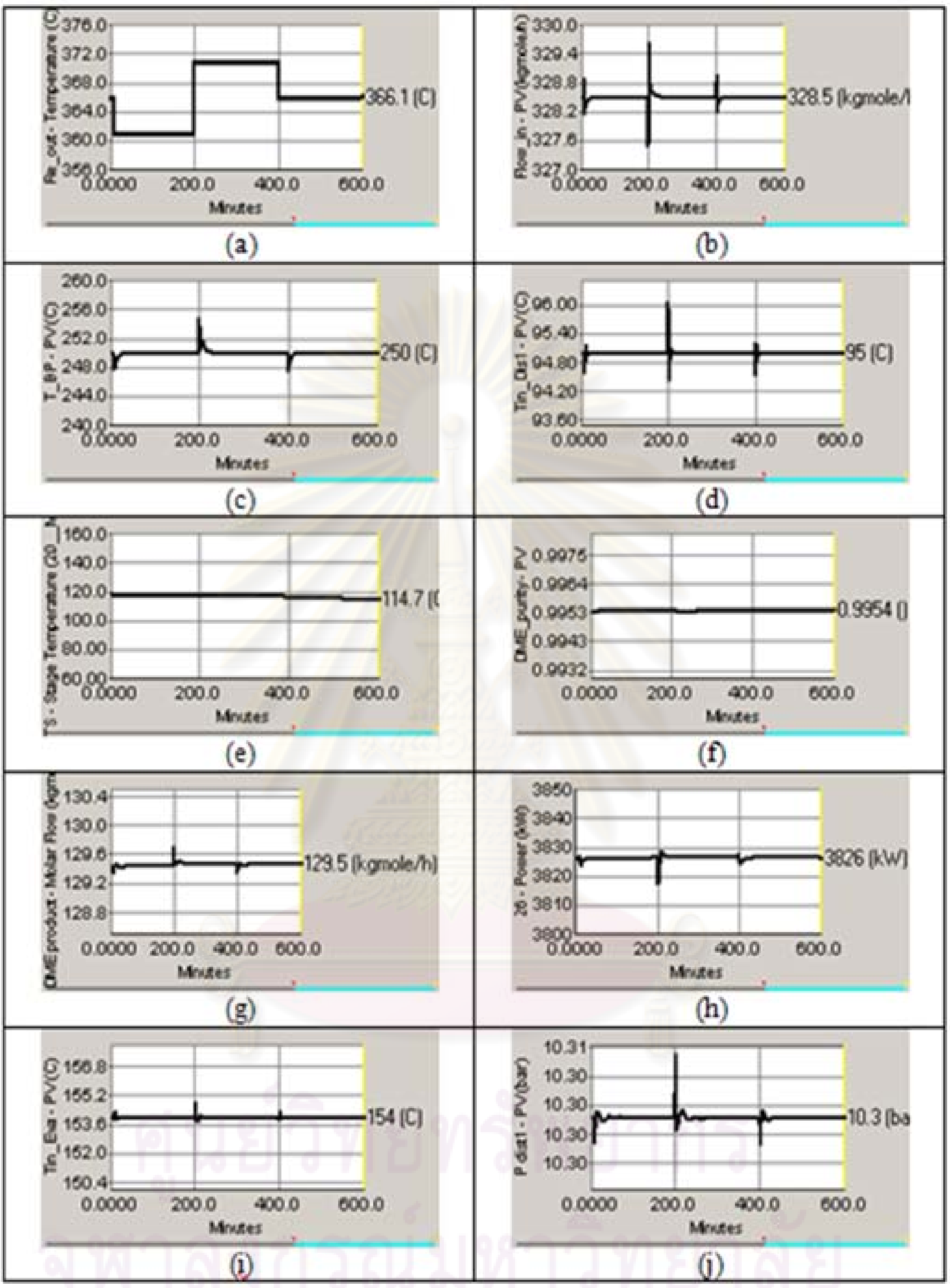

Figure 5.27 Dynamic Responses of the DME Process Base Case to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS2, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

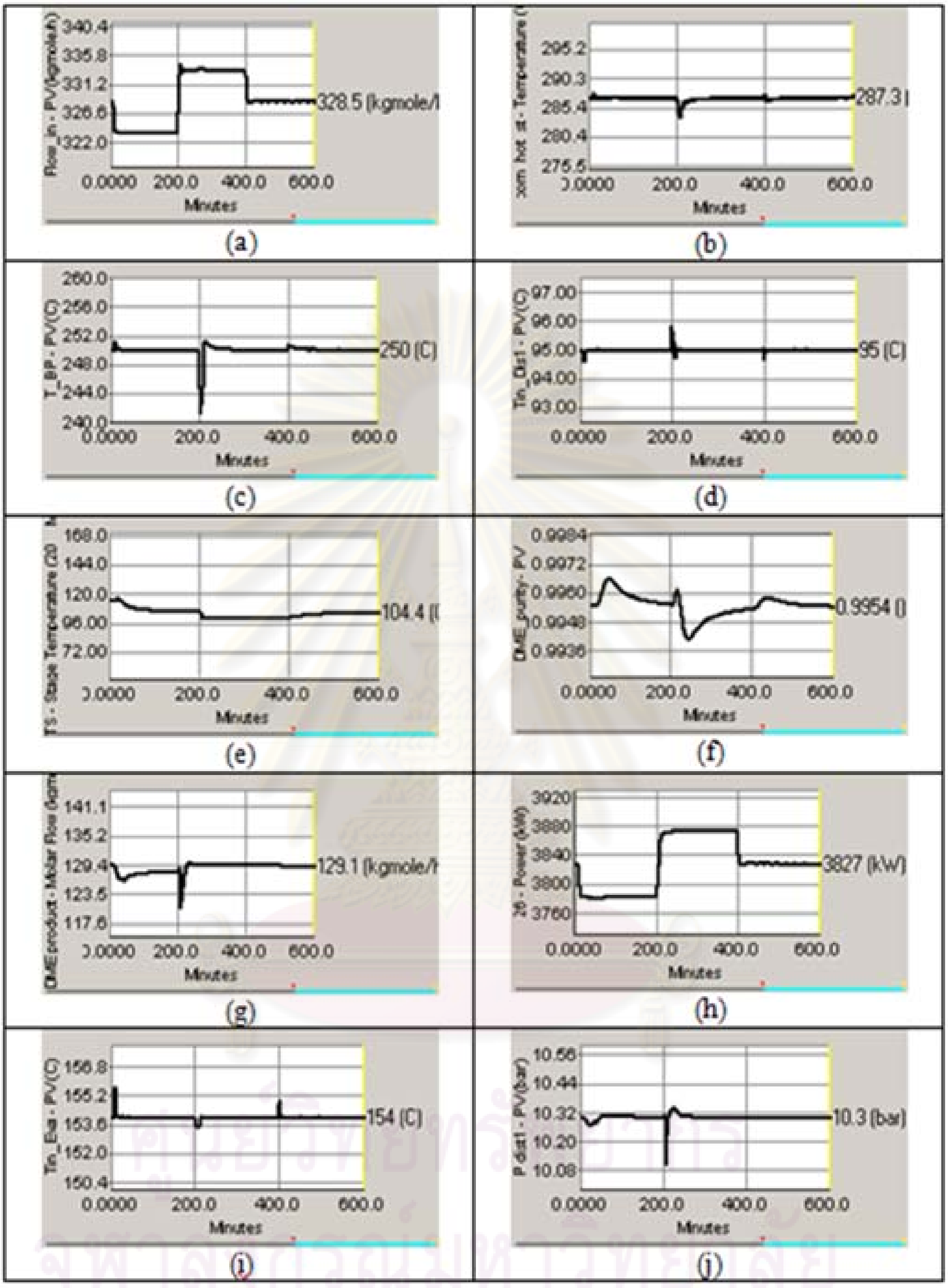

Figure 5.28 Dynamic Responses of the DME Process Base Case to Change the flow rate Disturbance of process stream: CS2, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

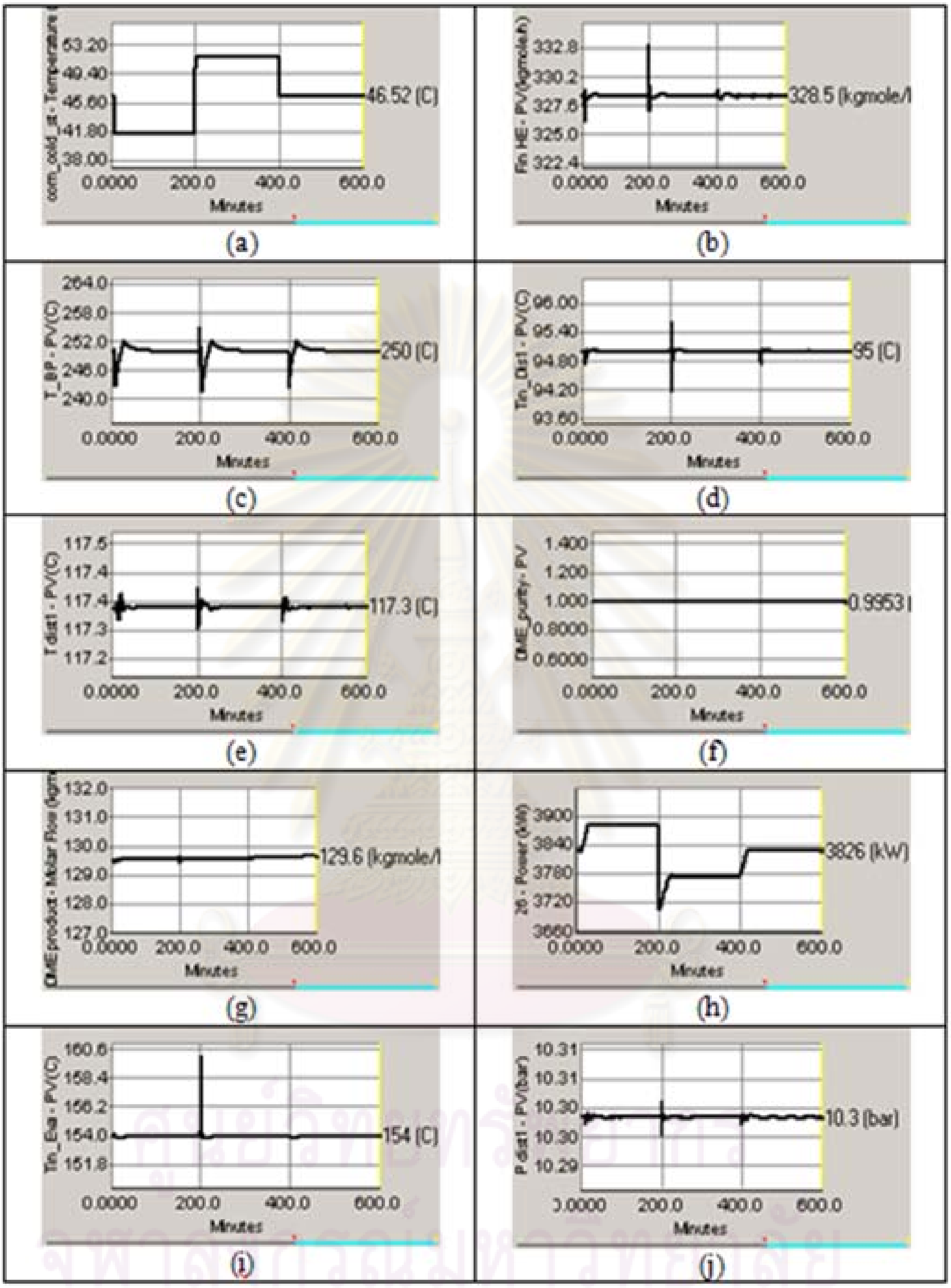

Figure 5.29 Dynamic Responses of the DME Process Base Case to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

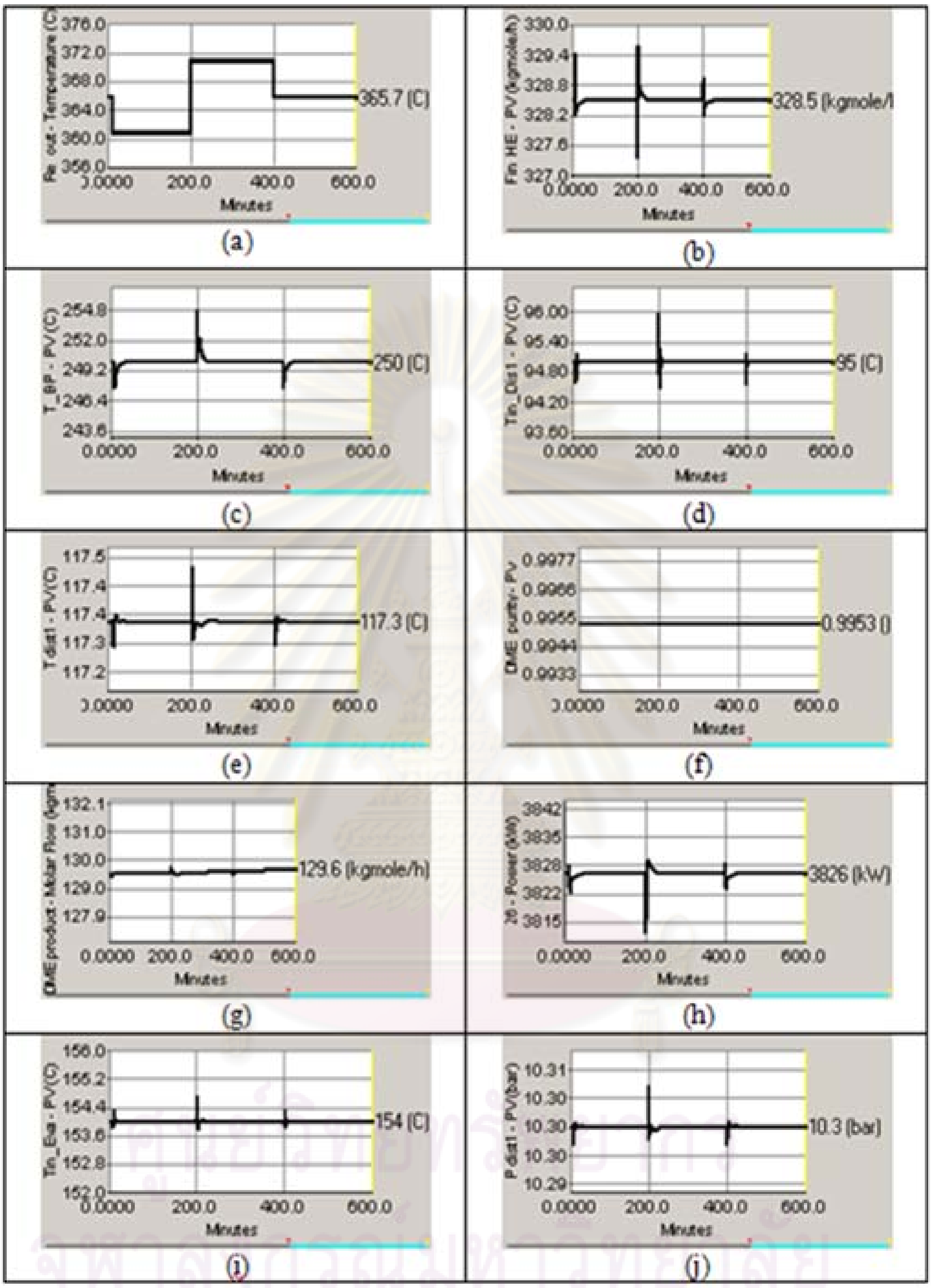

Figure 5.30 Dynamic Responses of the DME Process Base Case to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS3, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

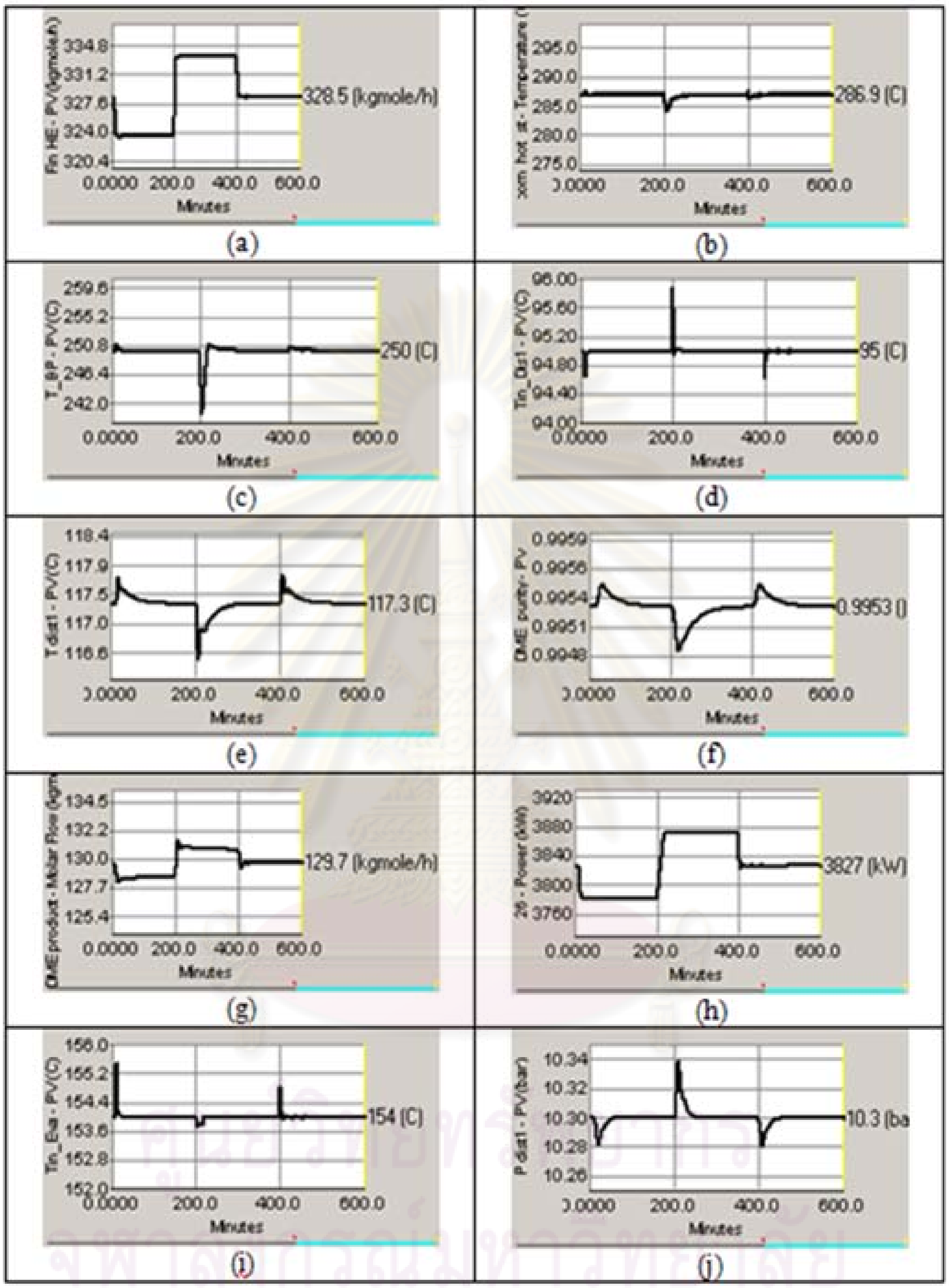

Figure 5.31 Dynamic Responses of the DME Process Base Case to Change the flow rate Disturbance of process stream: CS3, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

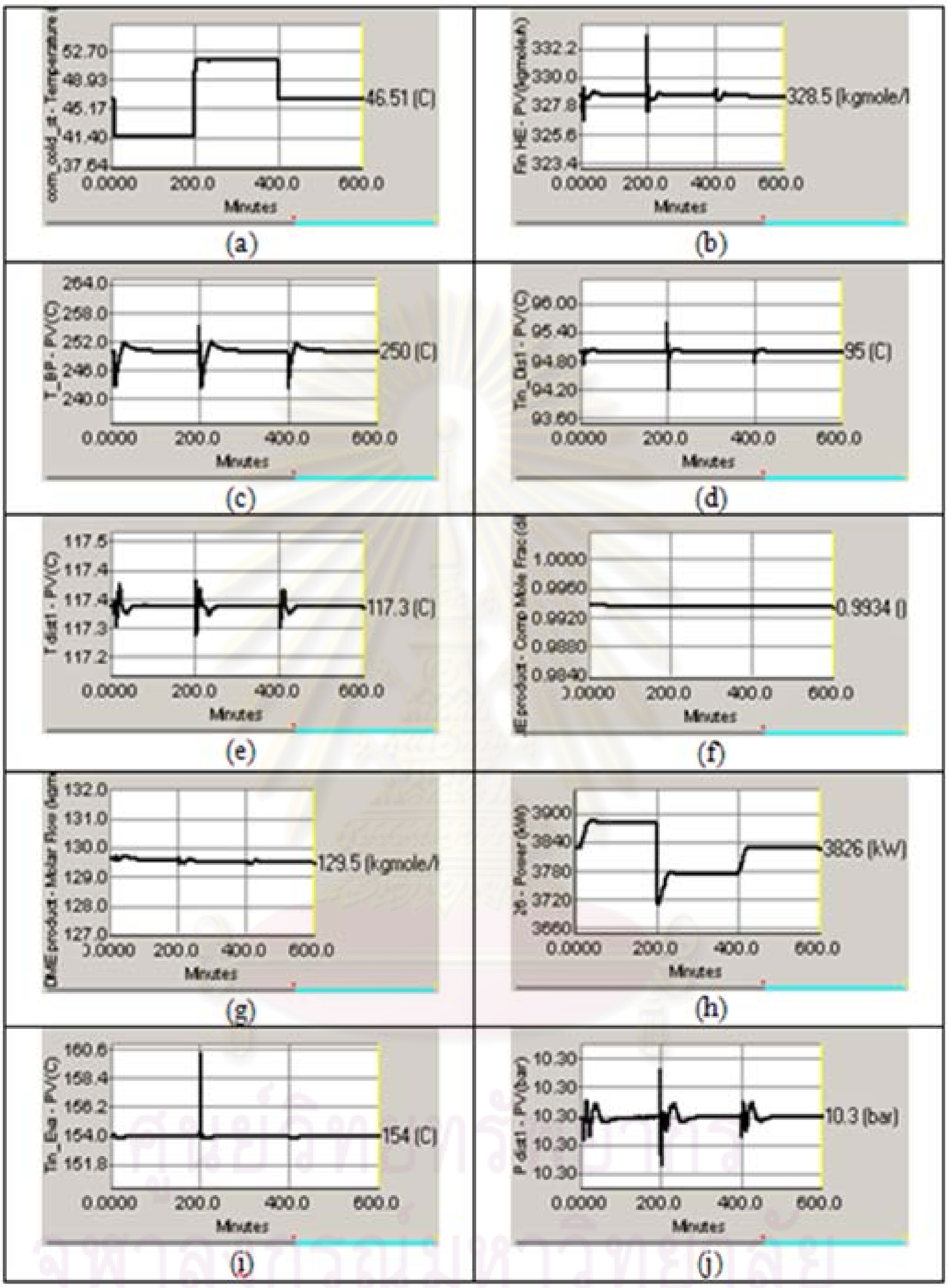

Figure 5.32 Dynamic Responses of the DME Process Base Case to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS4, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

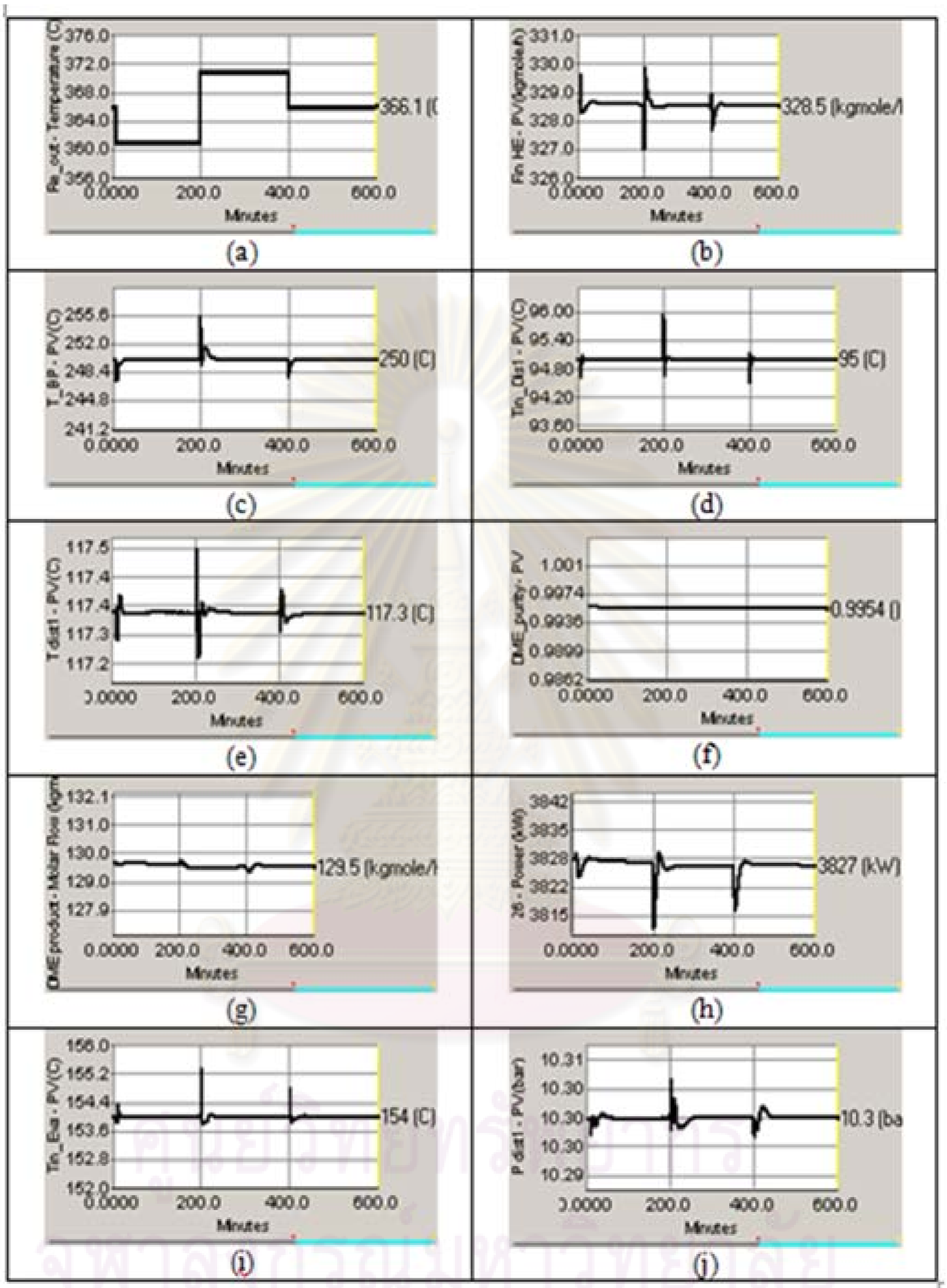

Figure 5.33 Dynamic Responses of the DME Process Base Case to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS4, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

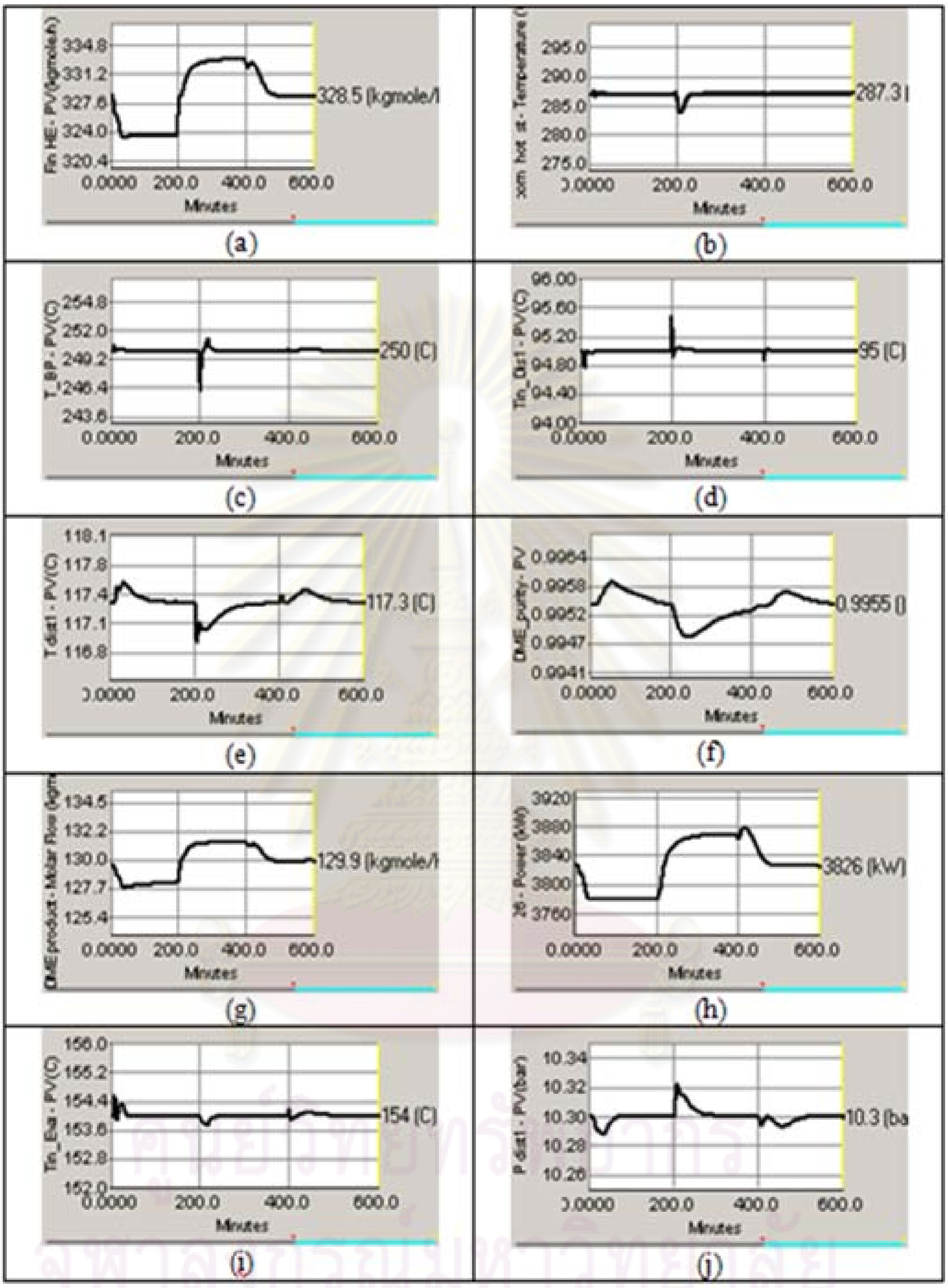

Figure 5.34 Dynamic Responses of the DME Process Base Case to Change the flow rate Disturbance of process stream: CS4, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

5.4.2 Dynamic Simulation Results of the DME Process Alternative 1

Three disturbance loads are used to evaluate the dynamic performance of the base control structure CS1 and design control structures from Luyben's heuristic CS2 and fixture point method CS3 and CS4 for DME process base case.

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

Figures 5.35, 5.38, 5.41 and 5.44 show the dynamic responses of the control systems of DME process for alternative 1 to a change in the heat load disturbance of the cold stream (reactor feed stream). In order to make this disturbance, first the temperature of combine stream between feed stream and recycle is decreased from 46.4^0C to 41.4^0C occurring at time equals 10 minutes, and the temperature is increased from 41.4^0C to 51.4^0C occurring at time equals 200 minutes, then its temperature is returned to its nominal value of 41.4^0C occurring at time equals 400 minutes (Figures 5.35a, 5.38a, 5.41a, and 5.44a).

This alternative has been managed heat load propagation of cold stream difference with base case that all control structures including base control structure CS1 and new control structure CS2, CS3 and CS4 give the same result to shift the heat load disturbance to FEHE outlet of cold stream and control temperature of FEHE outlet of hot stream by adjusting bypass valve and reject heat load disturbance of cold stream same as base case by vaporizer duty adjusting as shown in figures 5.35h, 5.38h, 5.41h and 5.44h. The dynamic responsibility of vaporizer outlet temperature of this alternative is better than Base Case because the cold stream is pre-heat by FEHE before send to vaporizer so the disturbance makes a little liquid at stream outlet of vaporizer at short period. The dynamic response of CS3 and CS4 are smoother than CS1 and CS2 because CS3 and CS4 are controlled both of product column temperature and DME product purity but CS1 control only product column temperature and CS2 control only DME product.

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

Figures 5.36, 5.39, 5.42 and 5.45 show the dynamic responses of the control systems of DME process of alternative 1 to a change in the heat load disturbance of hot stream (Reactor Product stream).

In order to make this disturbance, first the temperature of the reactor outlet is decreased from 365.7^0C to 360.7^0C occurring at time equals 10 minutes, and the temperature is increased from 360.7^0C to 370.7^0C occurring at time equals 200 minutes, then its temperature is returned to its nominal value of 365.7^0C occurring at time equals 400 minutes (Figures 5.36a, 5.39a, 5.42a and 5.45a). The disturbances propagation of the hot stream are difference with base case that shift load to the cold stream, hot stream outlet temperature of FEHE is controlled by bypass valve and cold stream temperature are reject by vaporizer duty. The heat load disturbance from the hot stream make effect smaller than the heat load disturbance from the cold stream because no phase change in FHEE as shown in figures 5.36c, 5.39c, 5.42c and 5.45c. The dynamic responses of product purity are slightly well and have a little bit affect production flow rate. However if we consider the product column temperature, the cold stream outlet temperature of FEHE, the dynamic response of CS3 and CS4 are smoother than CS1 and CS2 because CS3 and CS4 are controlled both of product column temperature and DME product purity but CS1 control only product column temperature and CS2 control only DME product purity.

5.4.2.3 Change in the Flow rates of Main Process stream

Figures 5.37, 5.40, 5.43 and 5.46 show the dynamic responses of the control systems of DME process alternative 1 to a change in the flow rates of main process stream. This disturbance is made by decreasing flow rates from 328.5 kgmole/h to 323.5 kgmol/h occurring at time equals 10 minutes, and the flow rates is increased from 323.5 kgmol/h to 333.5 kgmol/h occurring at time equals 100 minutes, then its flow rates is returned to its nominal value of 328.5 kgmol/h occurring at time equals 200 minutes (Figures 5.37a, 5.40a, 5.43a and 5.46a).

The dynamic result same as Base Case that the drop in flow rates reduces the reaction rate, so the production rate is decreased but DME product purity increased. The CS4 can control the temperature of product column better than CS1, CS2 and CS3 so deviation of DME product quality from nominal value in CS4 smoother than CS1, CS2 and CS3.

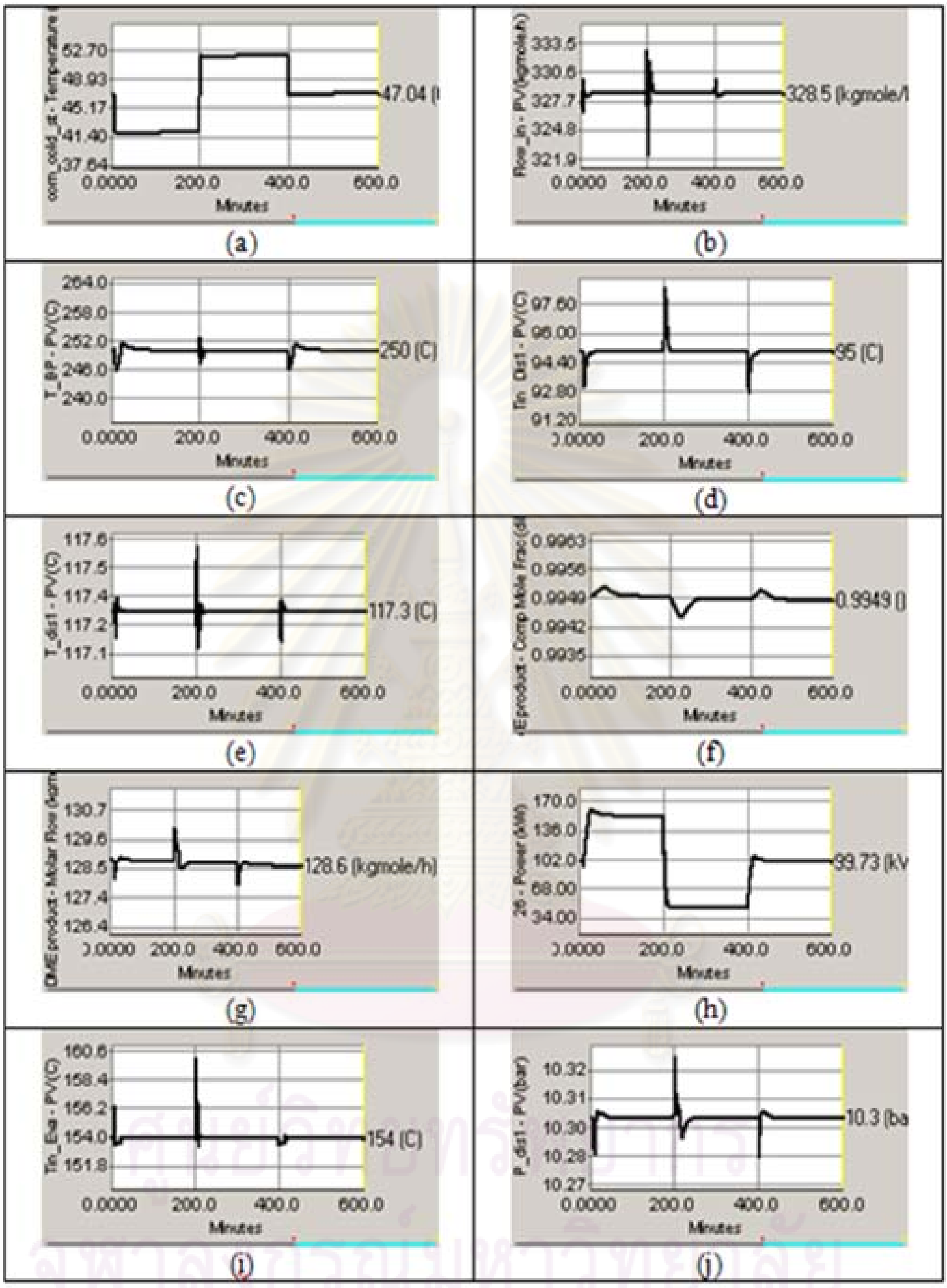

Figure 5.35 Dynamic Responses of the DME Process Alternative 1 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

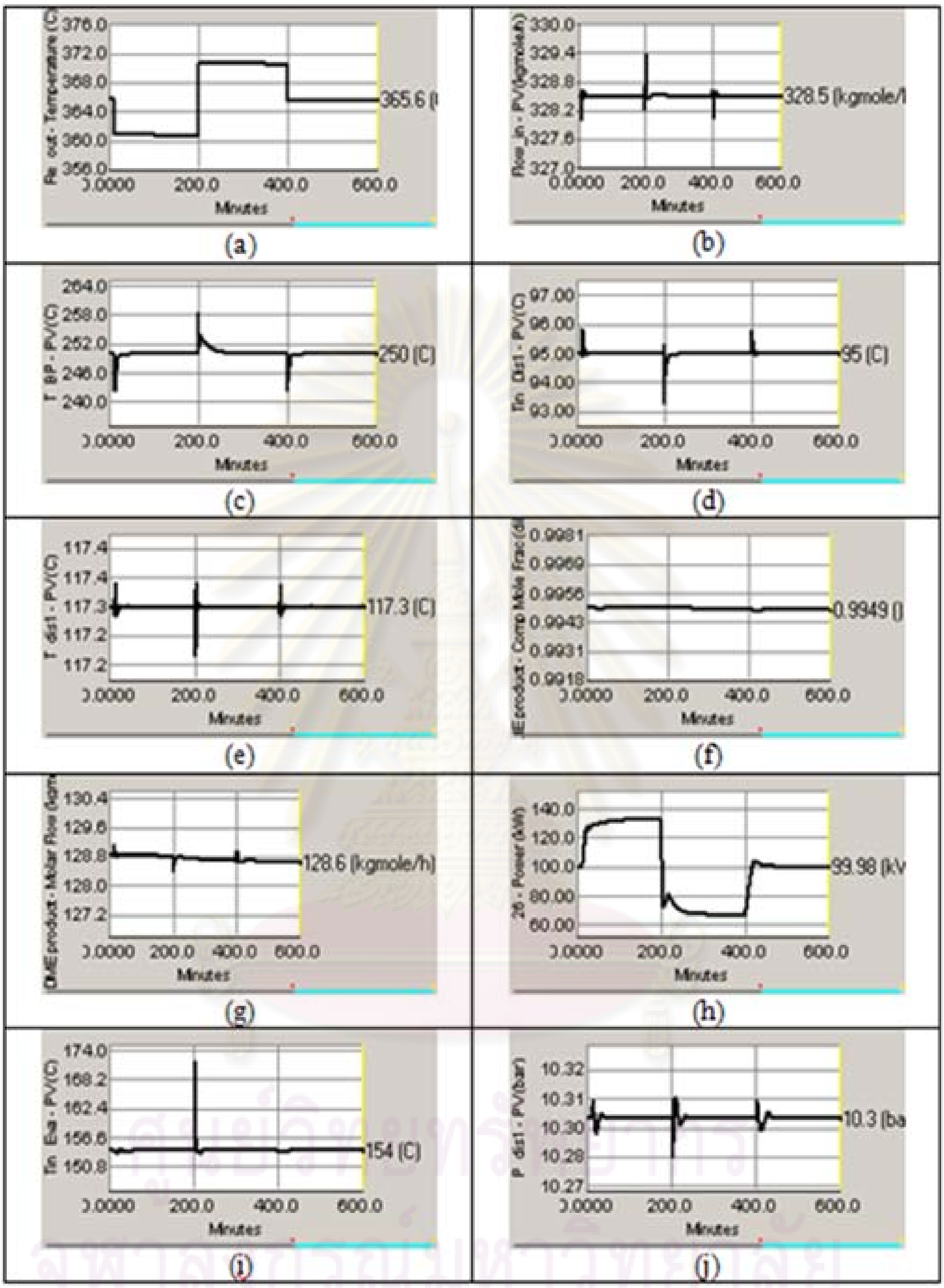

Figure 5.36 Dynamic Responses of the DME Process Alternative 1 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS1, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

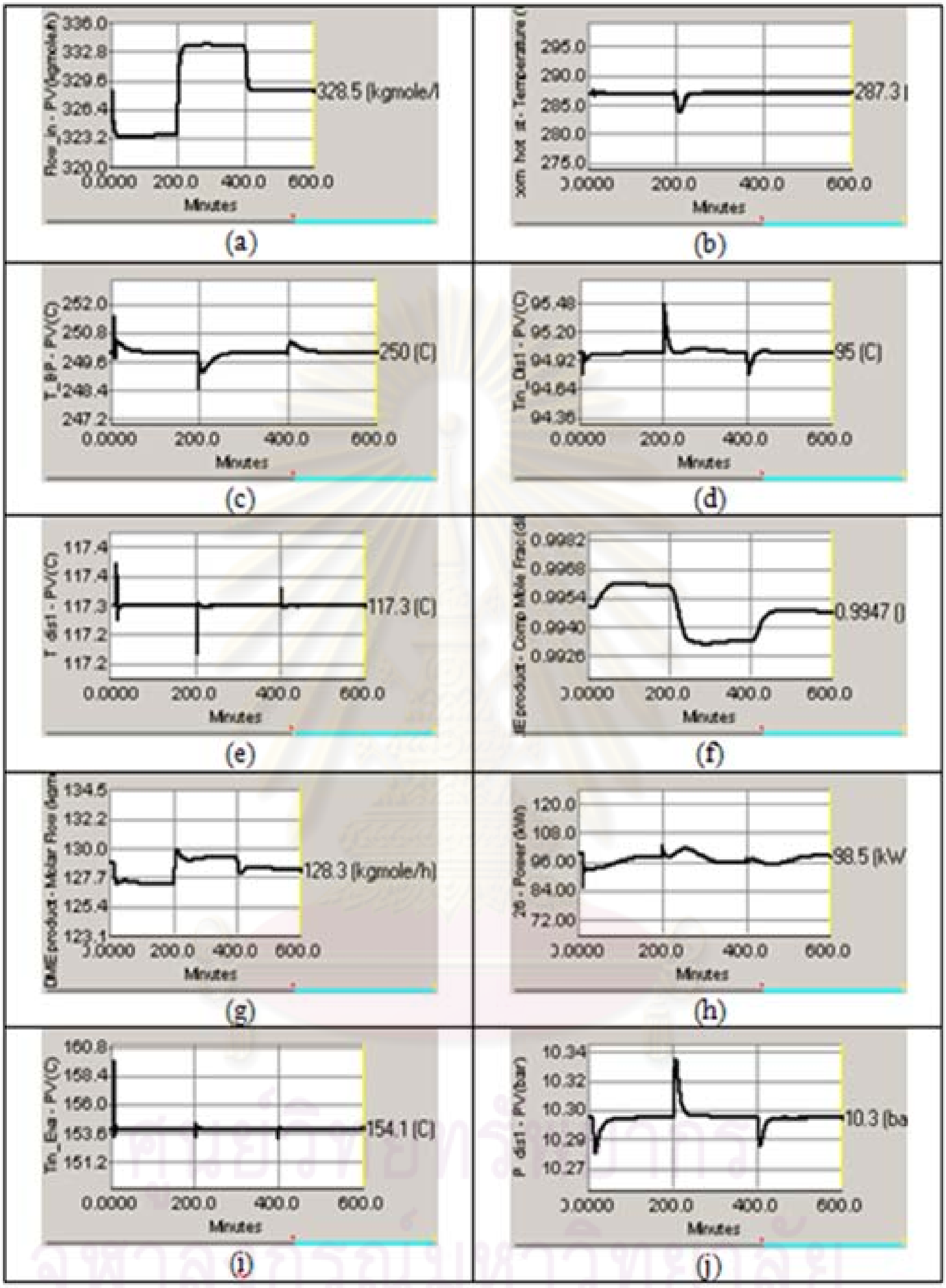

Figure 5.37 Dynamic Responses of the DME Process Alternative 1 to Change the flow rate Disturbance of process stream: CS1, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

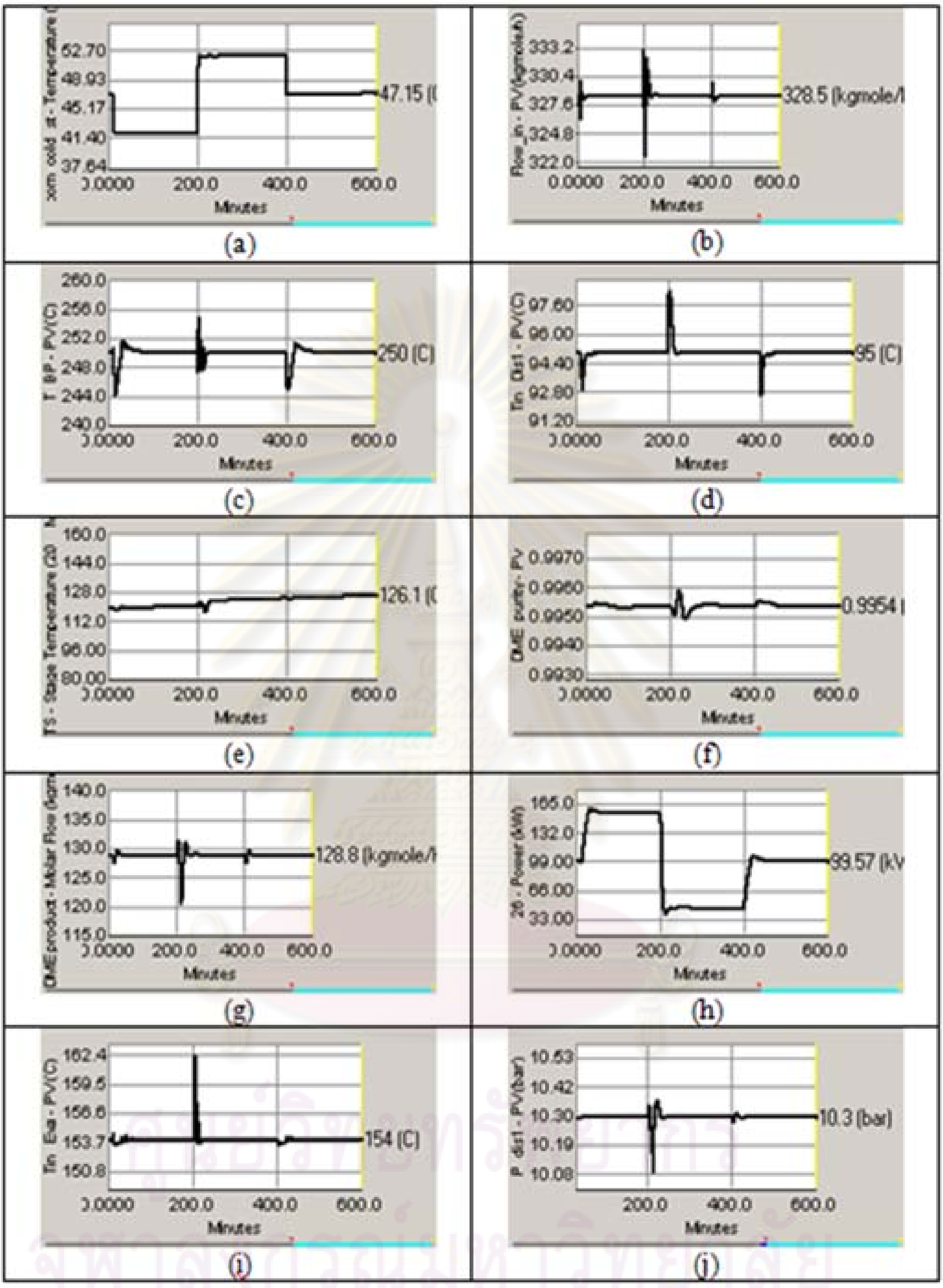

Figure 5.38 Dynamic Responses of the DME Process Alternative 1 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

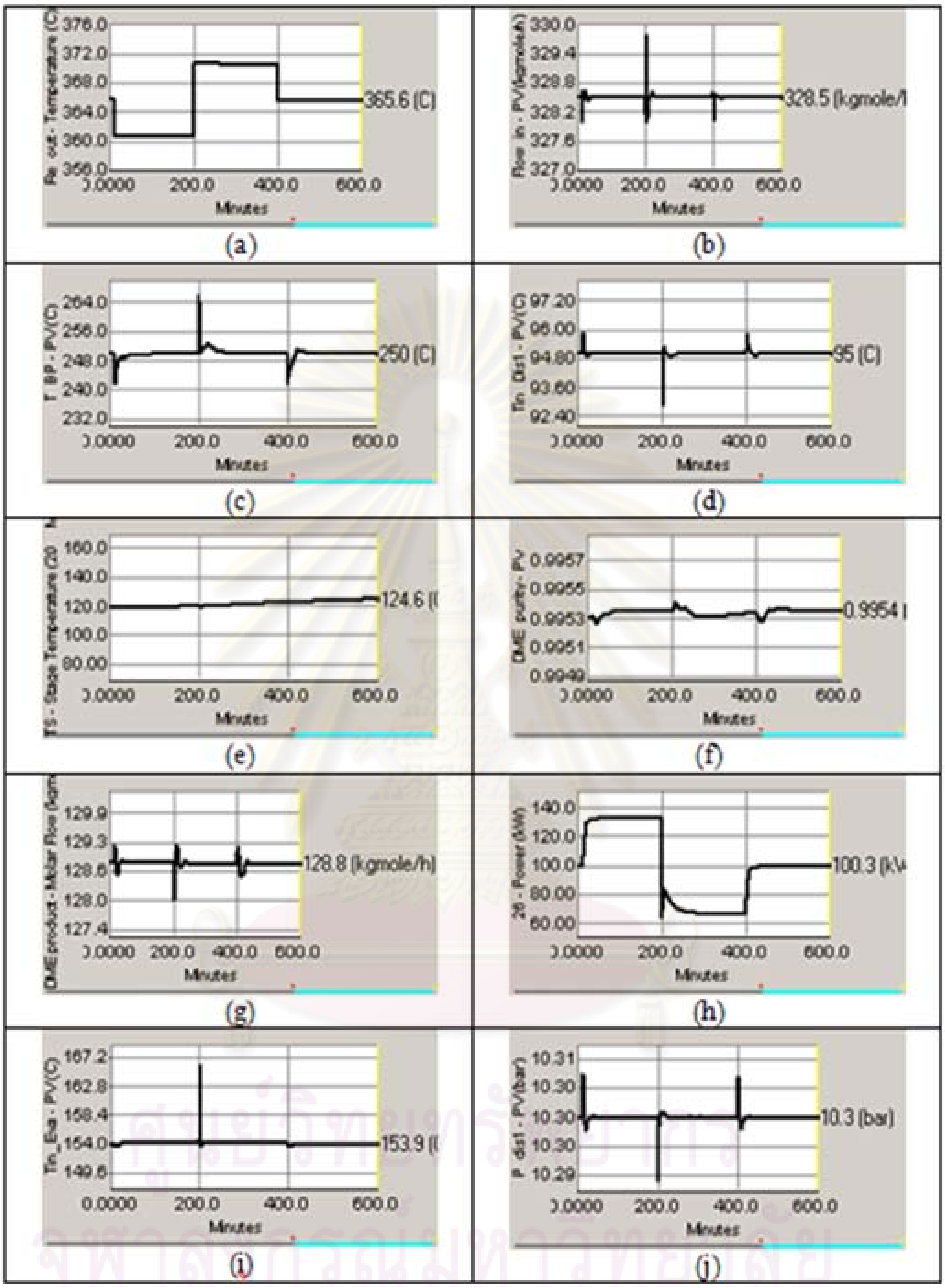

Figure 5.39 Dynamic Responses of the DME Process Alternative 1 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS2, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

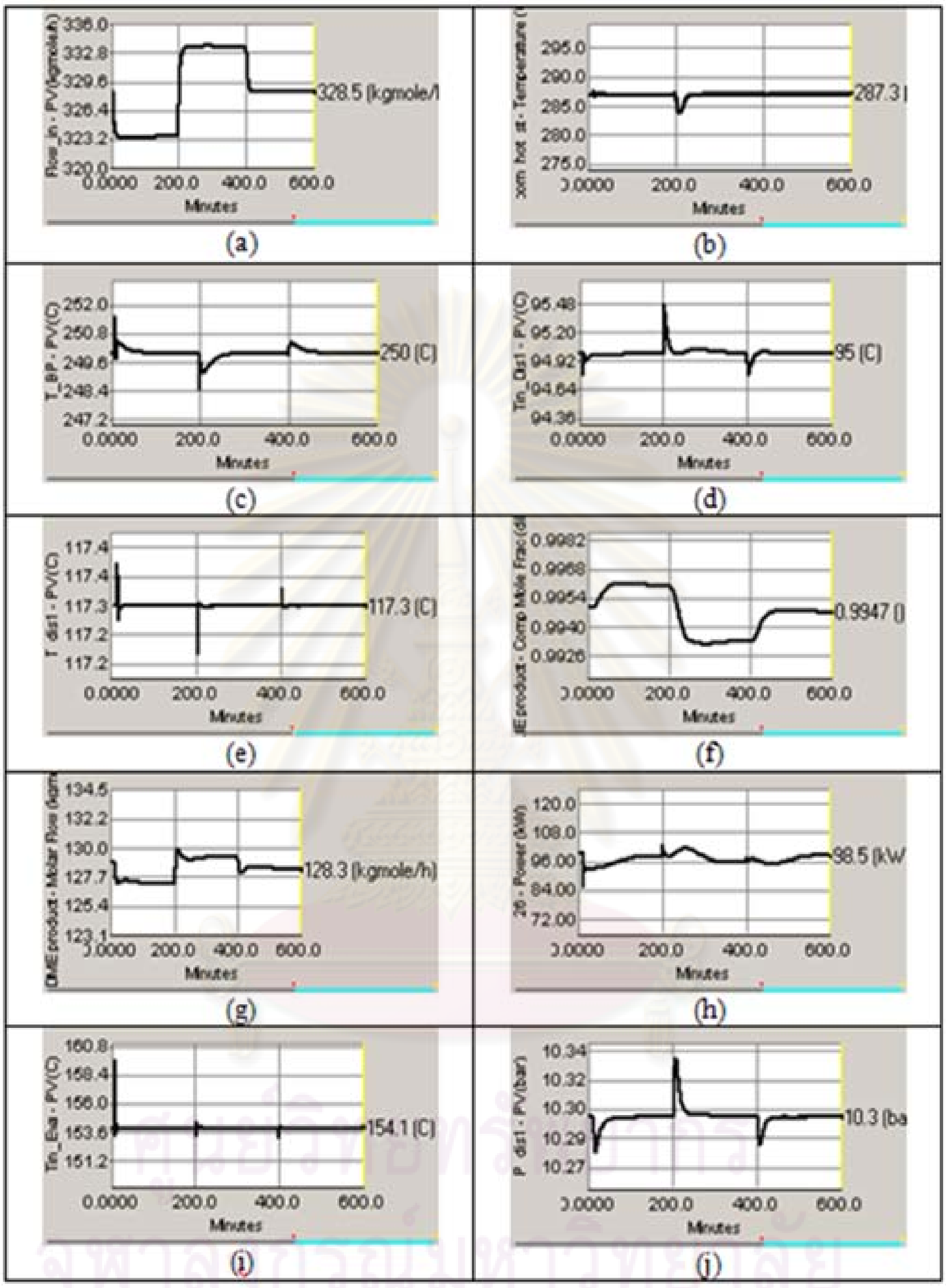

Figure 5.40 Dynamic Responses of the DME Process Alternative 1 to Change the flow rate Disturbance of process stream: CS2, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

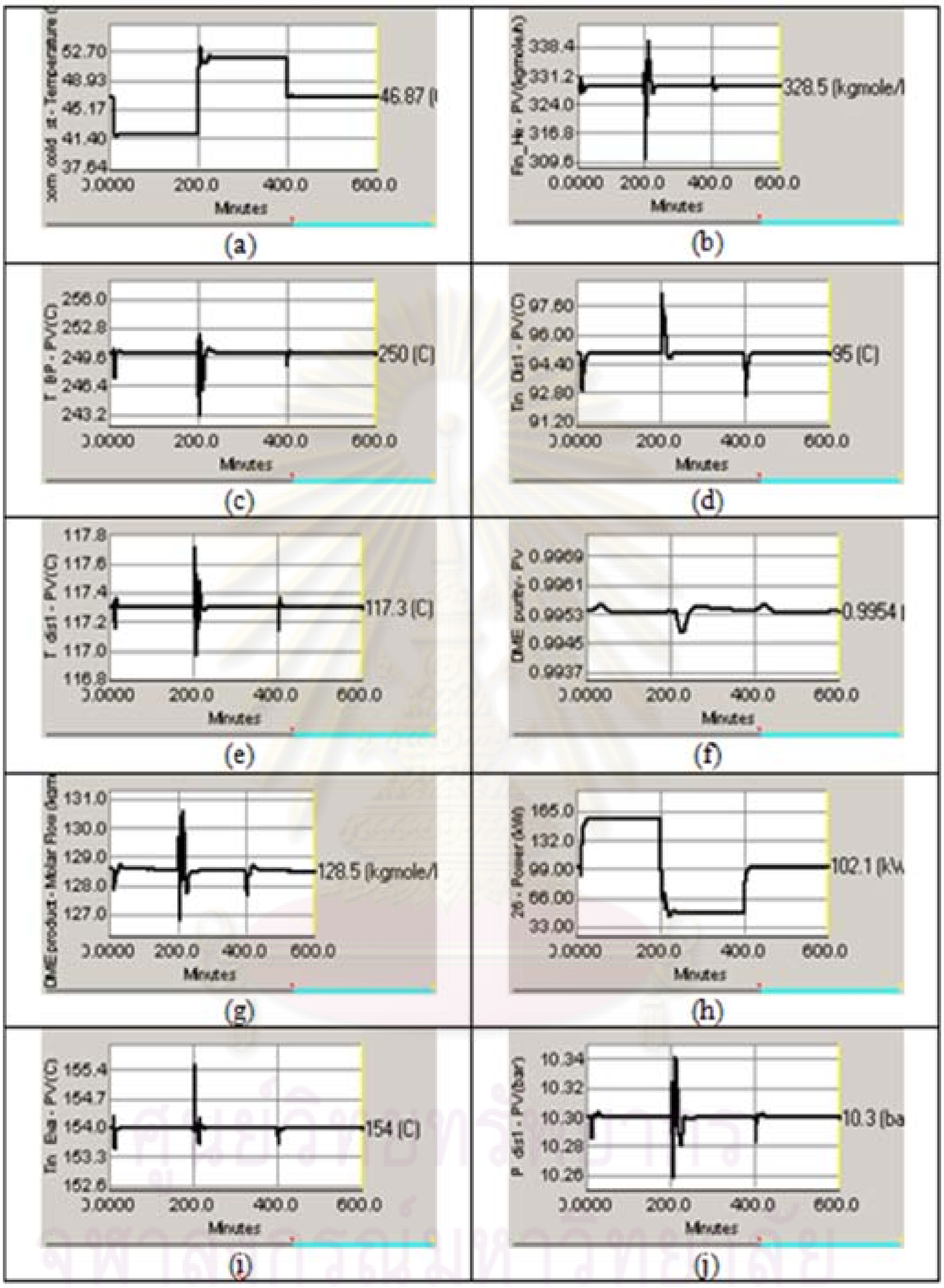

Figure 5.41 Dynamic Responses of the DME Process Alternative 1 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

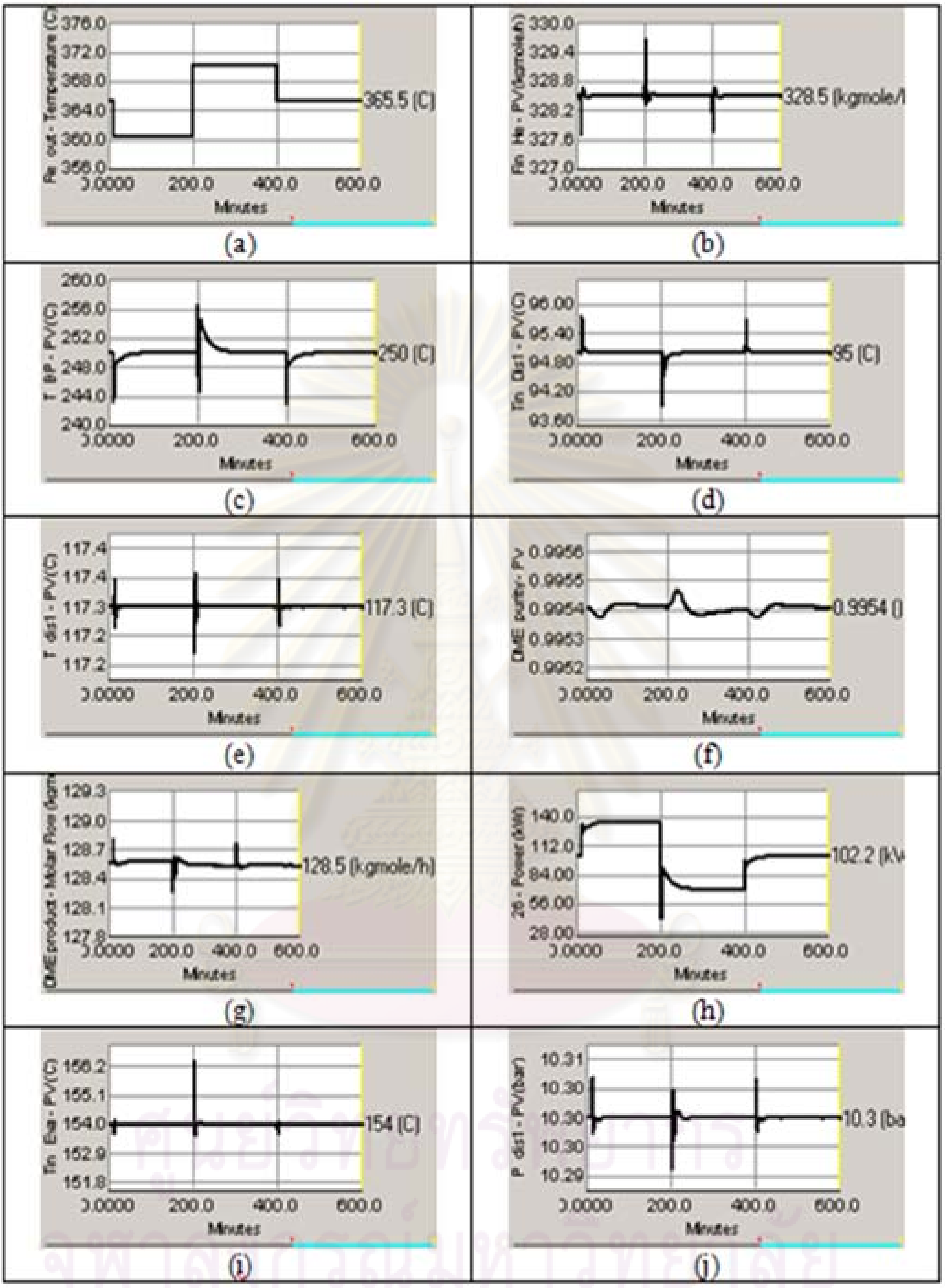

Figure 5.42 Dynamic Responses of the DME Process Alternative 1 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS2, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

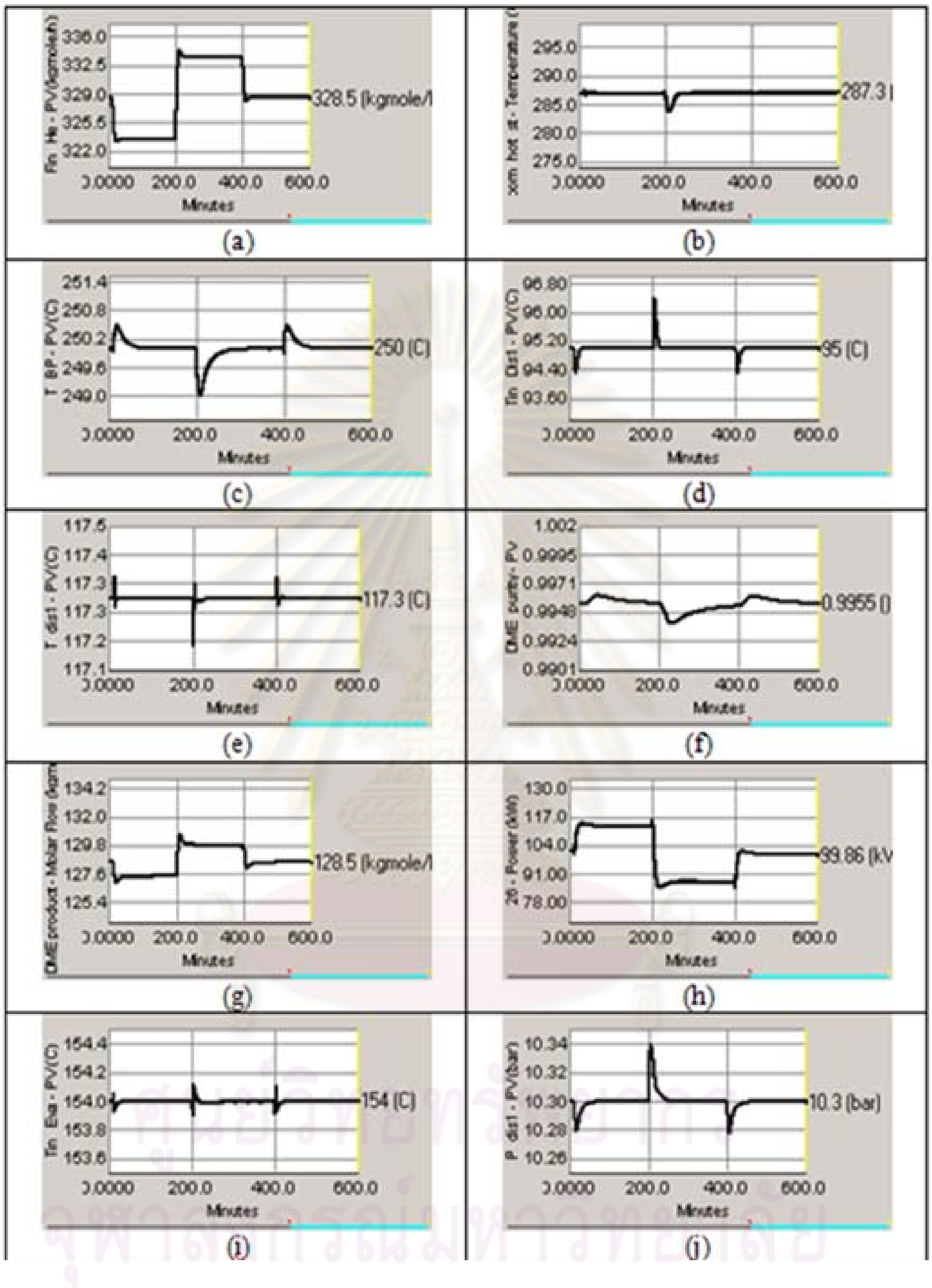

Figure 5.43 Dynamic Responses of the DME Process Alternative 1 to Change the flow rate Disturbance of process stream: CS3, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

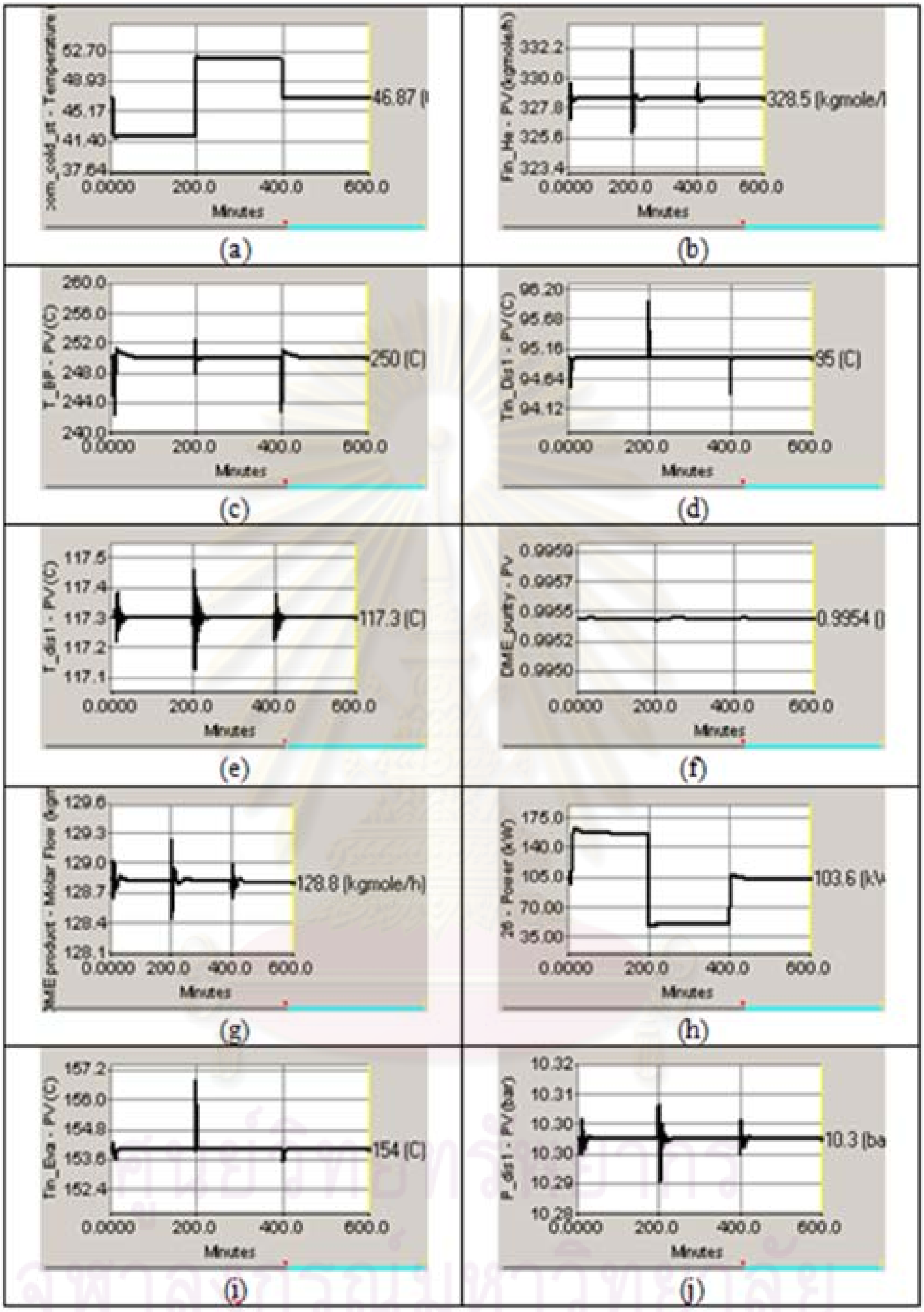

Figure 5.44 Dynamic Responses of the DME Process Alternative 1 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS4, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

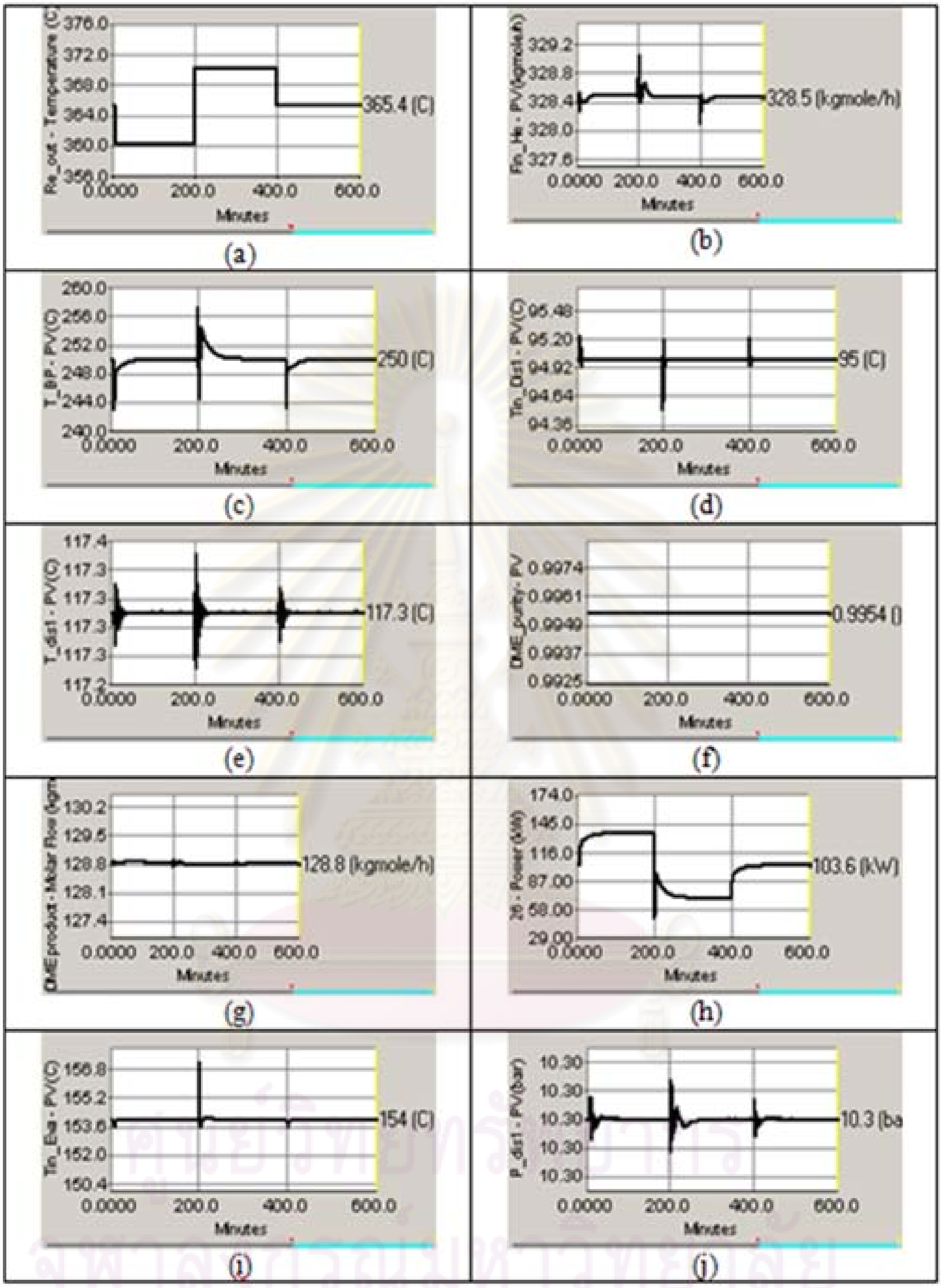

Figure 5.45 Dynamic Responses of the DME Process Alternative 1 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS4, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

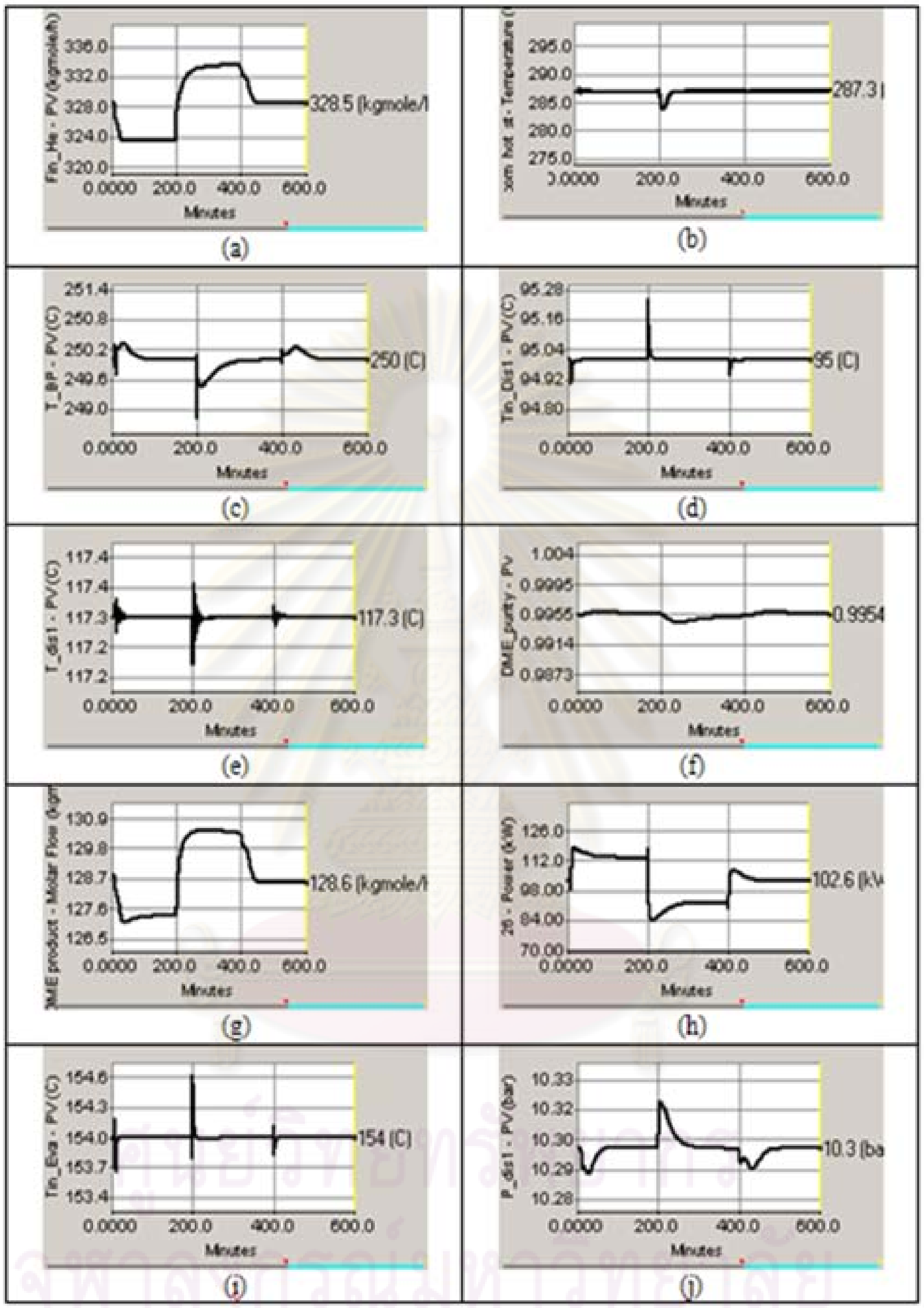

Figure 5.46 Dynamic Responses of the DME Process Alternative 1 to Change the flow rate Disturbance of process stream: CS4, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

5.4.3 Dynamic Simulation Results of the DME Process Alternative 2

Three disturbance loads are used to evaluate the dynamic performance of the base control structure CS1 and design control structures from Luyben's heuristic CS2 and fixture point method CS3 and CS4 for DME process base case.

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

Figures 5.47, 5.50, 5.53 and 5.56 show the dynamic responses of the control systems of DME process for alternative 2 to a change in the heat load disturbance of the cold stream (reactor feed stream). In order to make this disturbance, first the temperature of combine stream between feed stream and recycle is decreased from 46.4^0C to 41.4^0C occurring at time equals 10 minutes, and the temperature is increased from 41.4^0C to 51.4^0C occurring at time equals 200 minutes, then its temperature is returned to its nominal value of 41.4^0C occurring at time equals 400 minutes (Figures 5.47a, 5.50a, 5.53a and 5.56a).

This alternative has been managed heat load propagation of cold stream same as Base Case that all control structures including base control structure CS1 and new control structure CS2, CS3 and CS4 give the same result to shift the heat load vaporizer directly and using duty to reject disturbance as shown in figures 5.47h, 5.50h, 5.53h and 5.56h. The dynamic responsibility of this alternative same as base case because it makes vapor fraction at vaporizer outlet not equal one for short period so reactor inlet temperature is decreased for short period. However CS4 can control temperature loop, production rate and DME product purity better than CS1, CS2 and CS3 because CS4 control pressure of steam before feed to FEHE, it makes temperature control smoothly at vaporizer outlet.

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

Figures 5.48, 5.51, 5.54 and 5.57 show the dynamic responses of the control systems of DME process of alternative 2 to a change in the heat load disturbance of hot stream (Reactor Product stream).

In order to make this disturbance, first the temperature of reactor outlet is decreased from 365.7^0C to 360.7^0C occurring at time equals 10 minutes, and the temperature is increased from 360.7^0C to 370.7^0C occurring at time equals 200 minutes, then its temperature is returned to its nominal value of 365.7^0C occurring at time equals 400 minutes (Figures 5.48a, 5.51a, 5.54a and 5.57a). The disturbances propagation of the hot stream are difference with the base case and alternative 1 that shift disturbance to the hot stream outlet of FEHE and reject disturbance by cooler. This alternative control temperature of product column by bypass valve so the deviation of temperature larger than base case and alternative 1 as shown in figures 5.48e, 5.51e, 5.54e and 5.57e. However when compare between control structures that the dynamic response of CS3 and CS4 are smoother than CS1, CS2 because CS3 and CS4 are controlled both of product column temperature and DME product purity but CS1 control only product column temperature and CS2 control only DME product purity.

5.4.3.3 Change in the Flow rates of Main Process stream

Figures 5.49, 5.52, 5.55 and 5.58 show the dynamic responses of the control systems of DME process alternative 1 to a change in the flow rates of main process stream. This disturbance is made by decreasing flow rates from 328.5 kgmol/h to 323.5 kgmol/h occurring at time equals 10 minutes, and the flow rates is increased from 323.5 kgmol/h to 333.5 kgmol/h occurring at time equals 100 minutes, then its flow rates is returned to its nominal value of 328.5 kgmol/h occurring at time equals 200 minutes (Figures 5.49a, 5.52a, 5.55a and 5.58a).

The dynamic result is the same as the base case and alternative 1 that the drop in flow rates reduces the reaction rate, so the production rate is decreased but DME product purity increased. The CS4 can control the temperature of product column better than CS1, CS3 and CS3 so deviation of DME product quality from nominal value in CS4 smoother than CS1, CS2 and CS3.

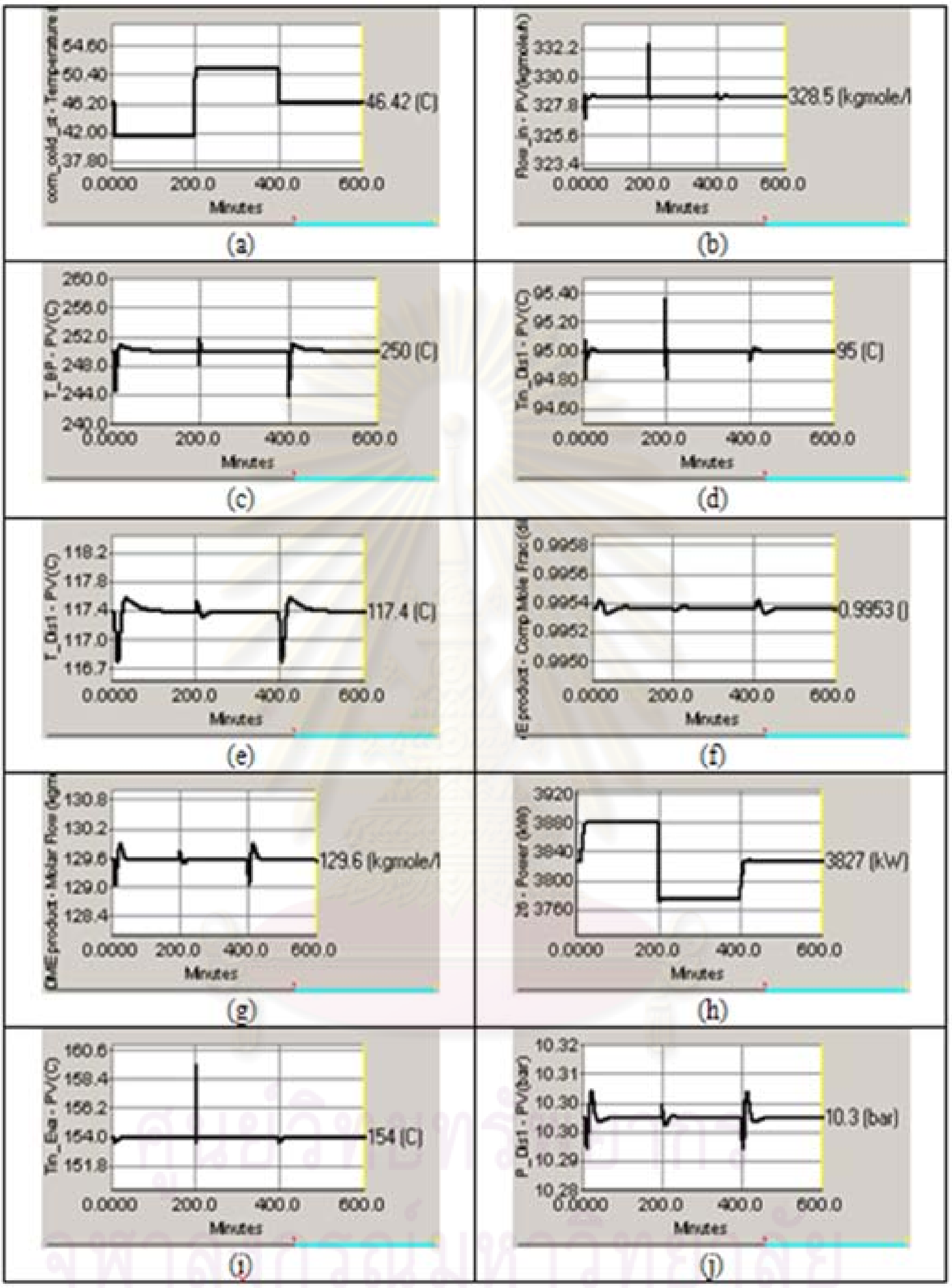

Figure 5.47 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

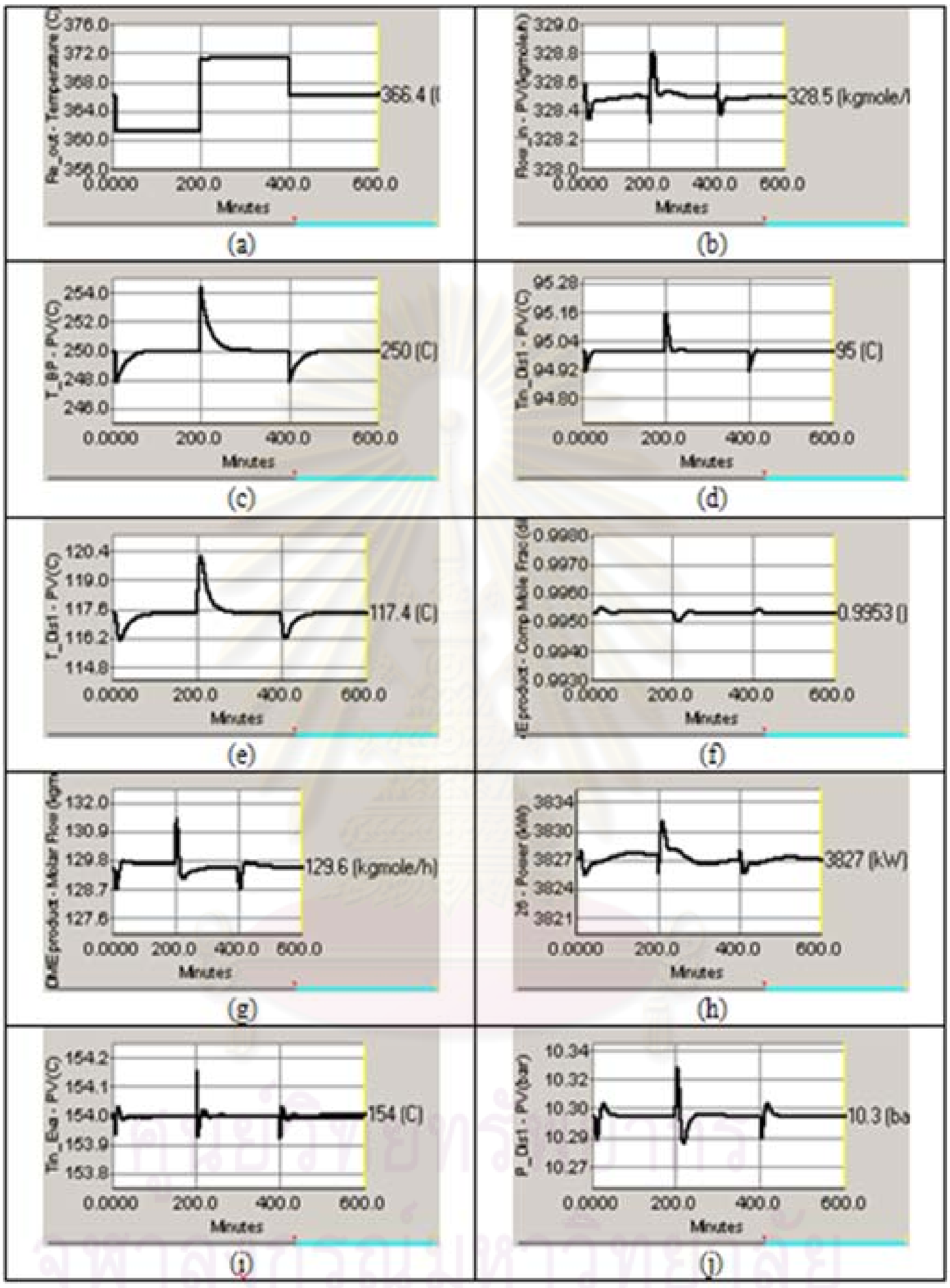

Figure 5.48 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS1, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

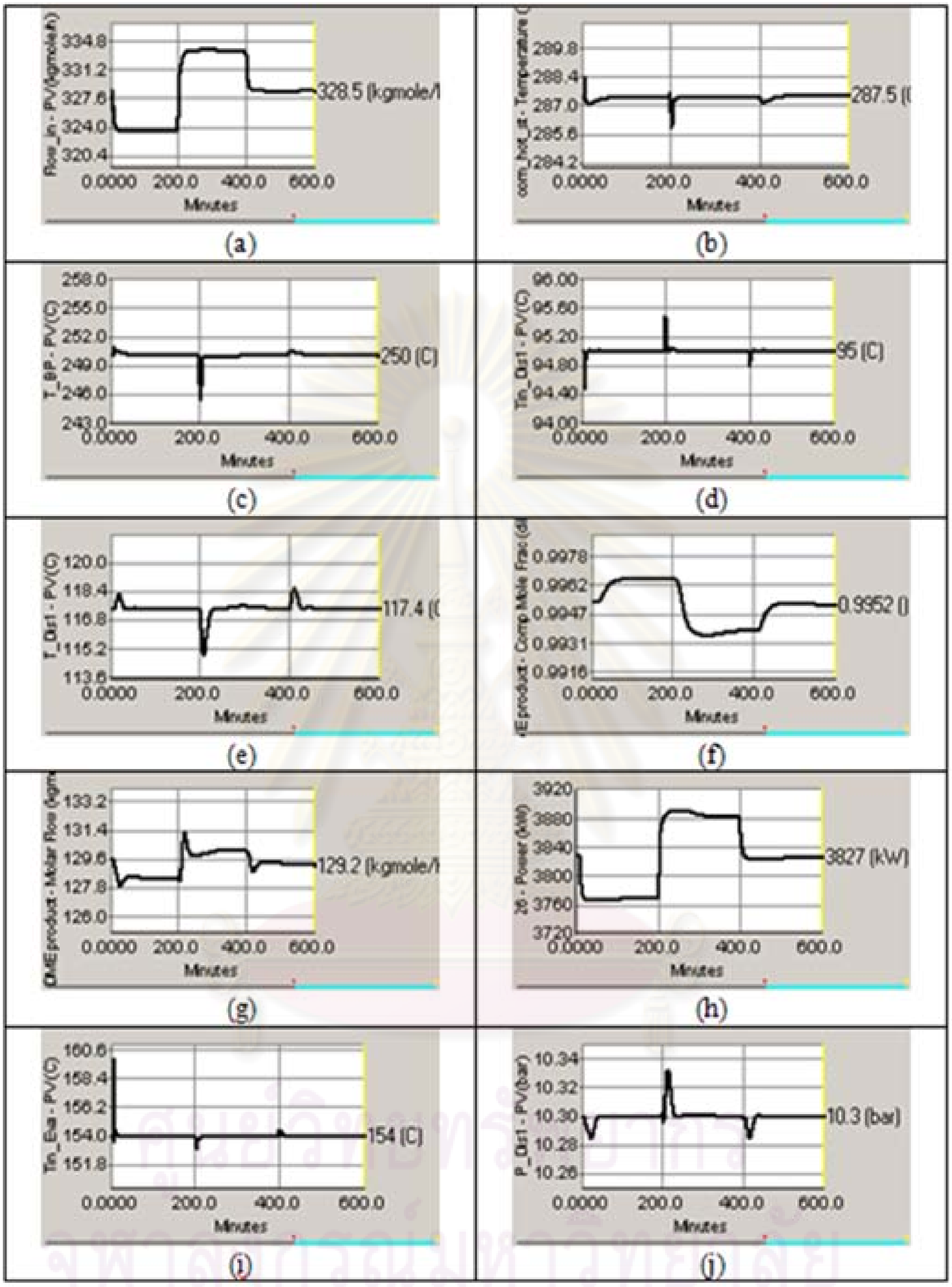

Figure 5.49 Dynamic Responses of the DME Process Alternative 2 to Change the flow rate Disturbance of process stream: CS1, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

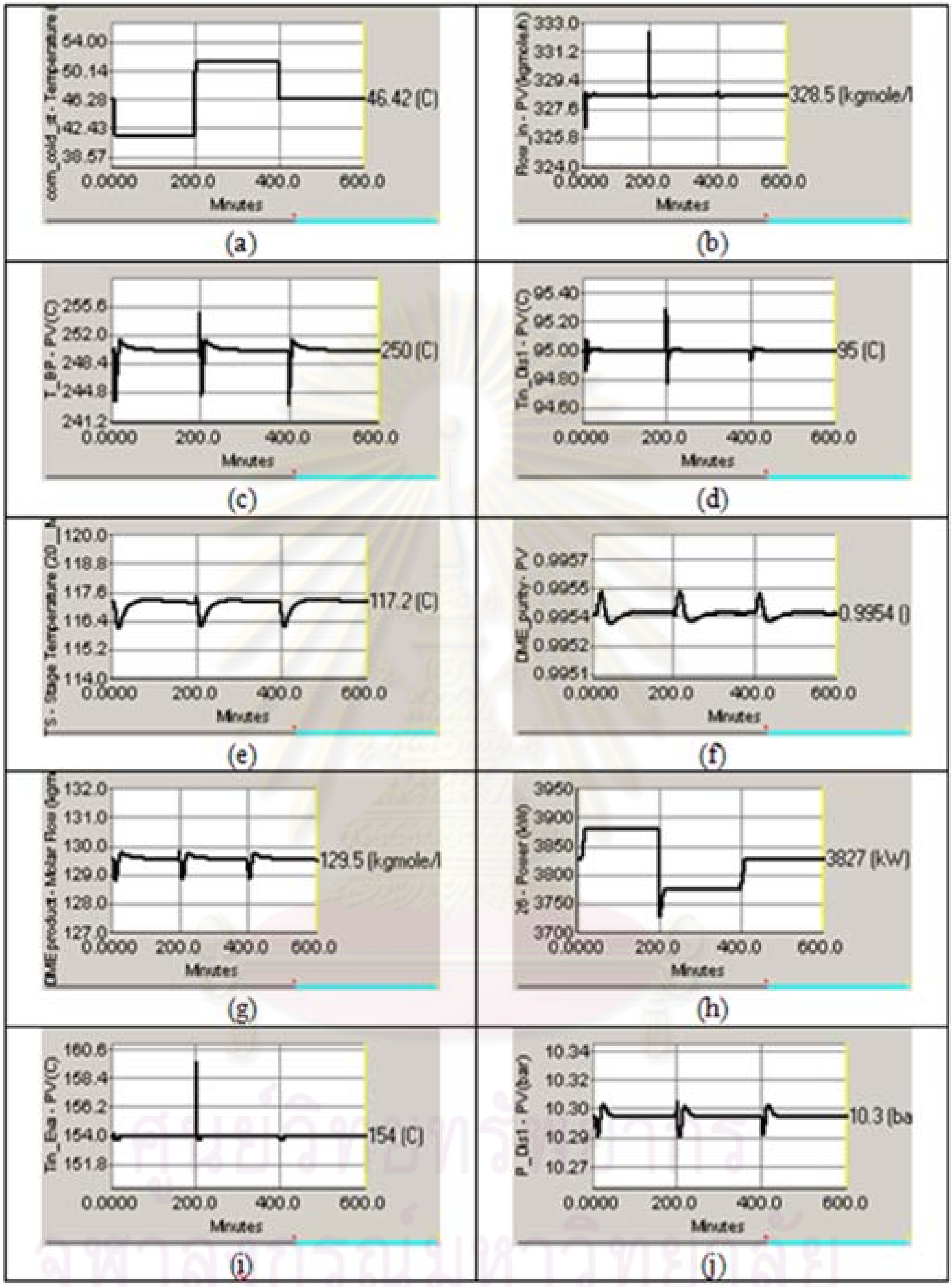

Figure 5.50 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

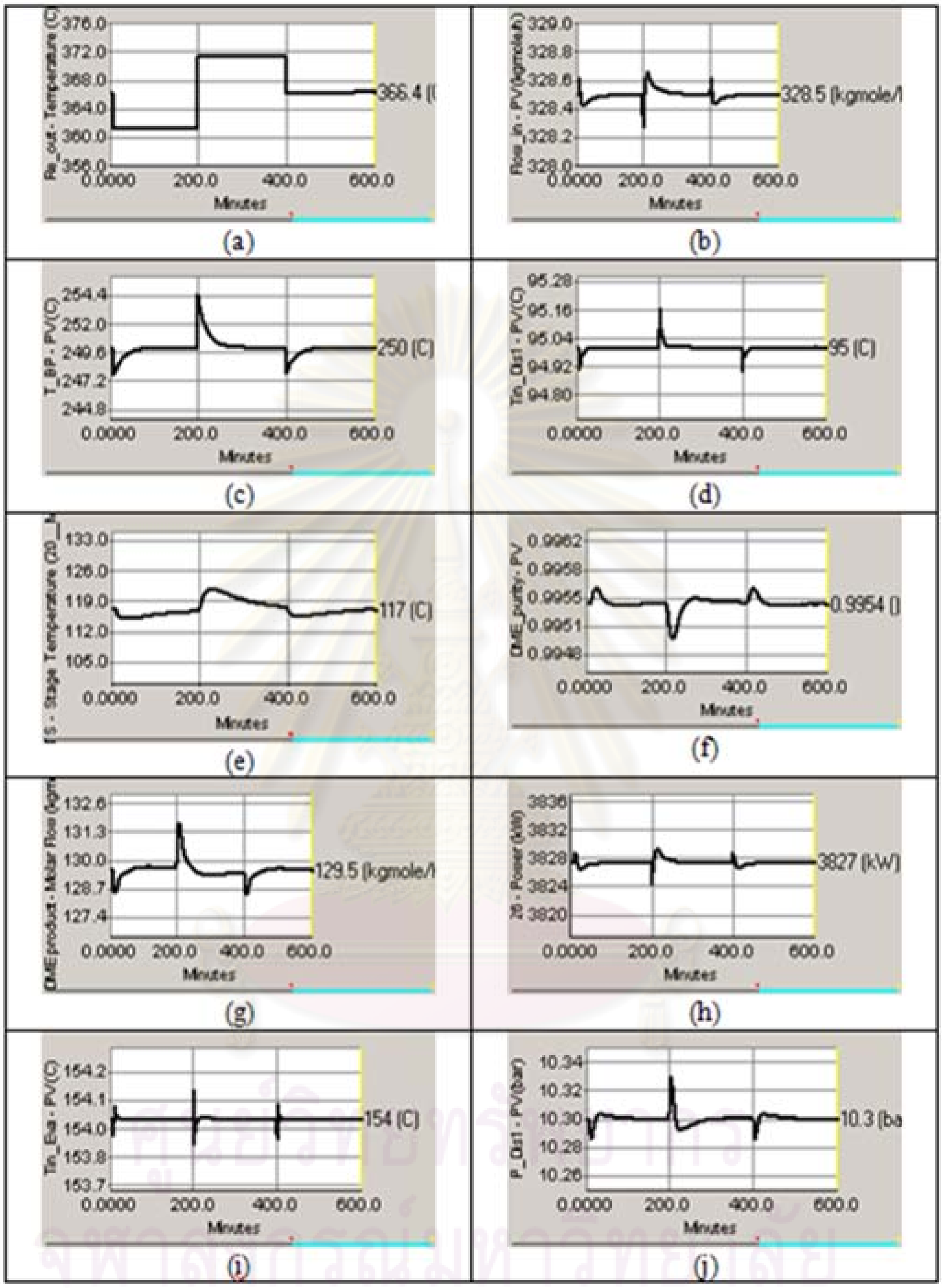

Figure 5.51 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS2, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

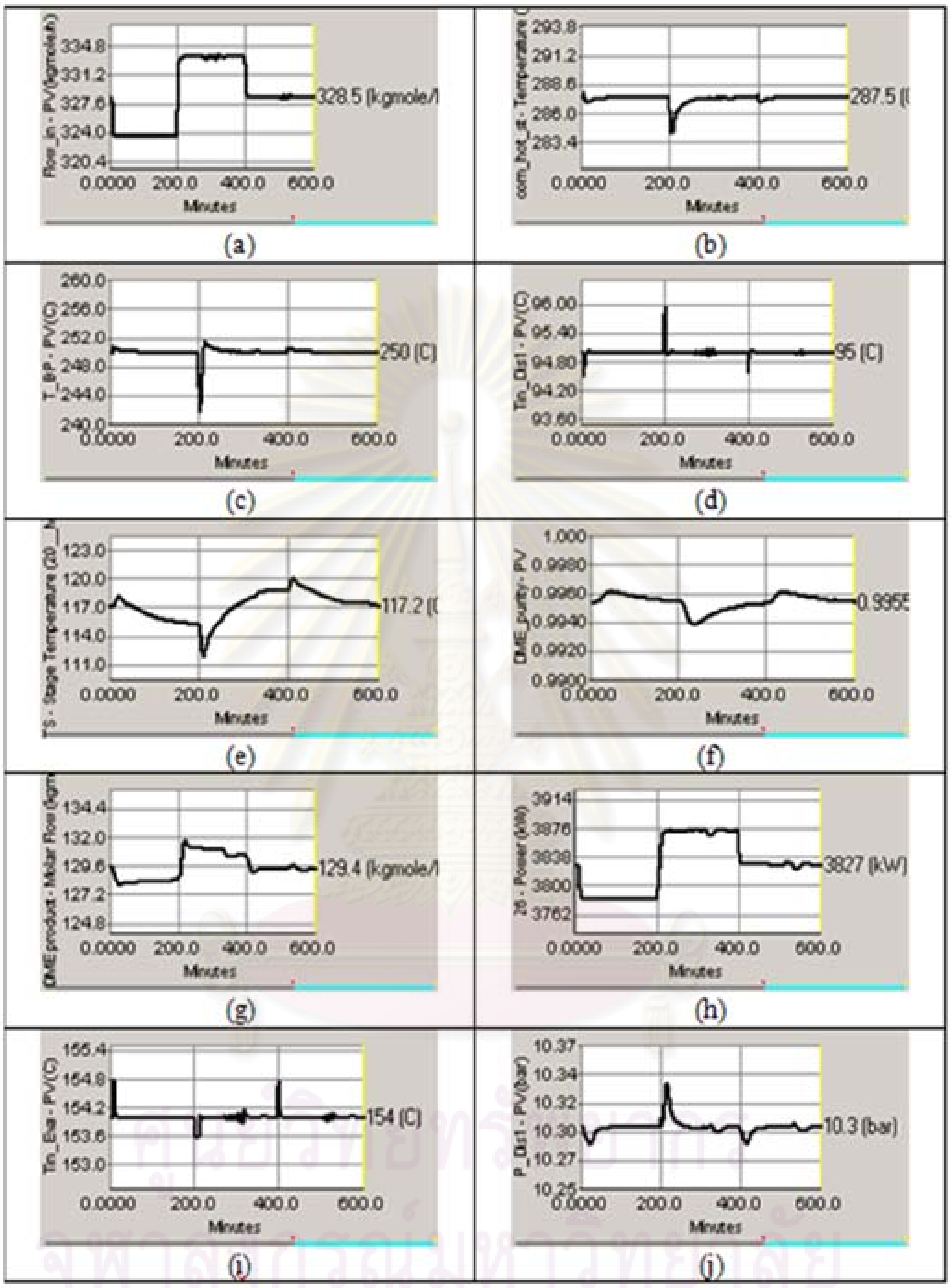

Figure 5.52 Dynamic Responses of the DME Process Alternative 2 to Change the flow rate Disturbance of process stream: CS2, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

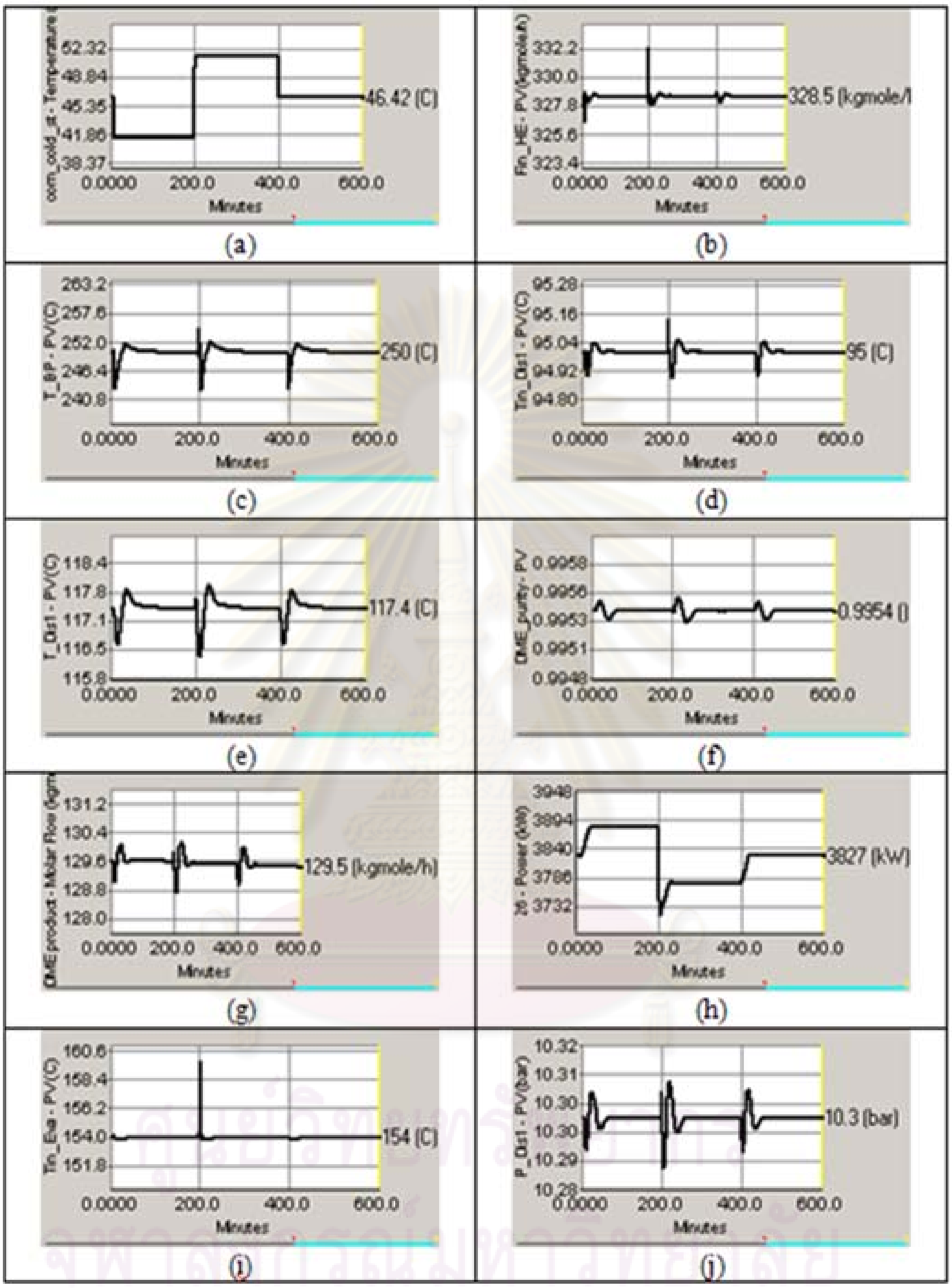

Figure 5.53 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS3, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

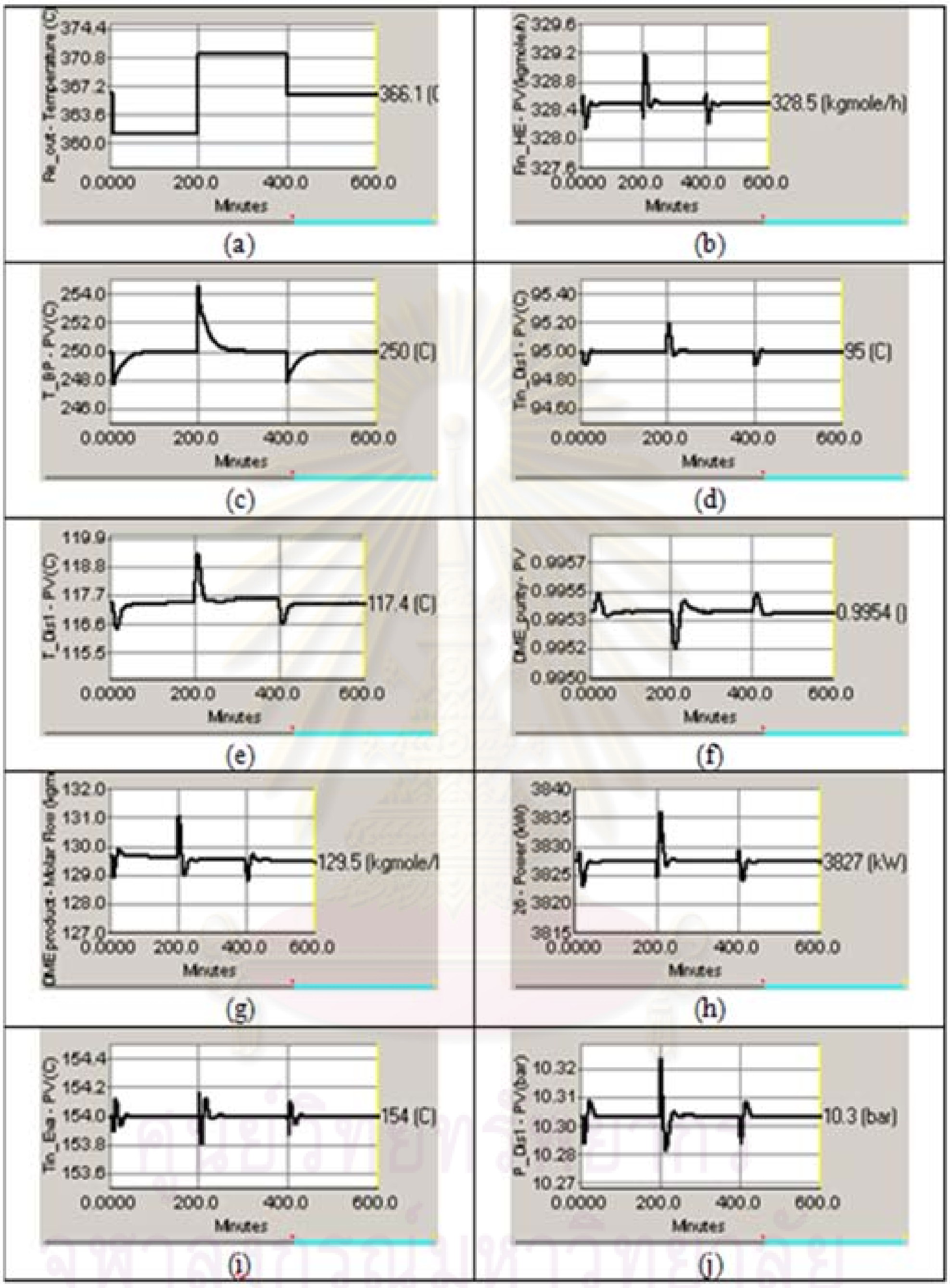

Figure 5.54 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS3, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

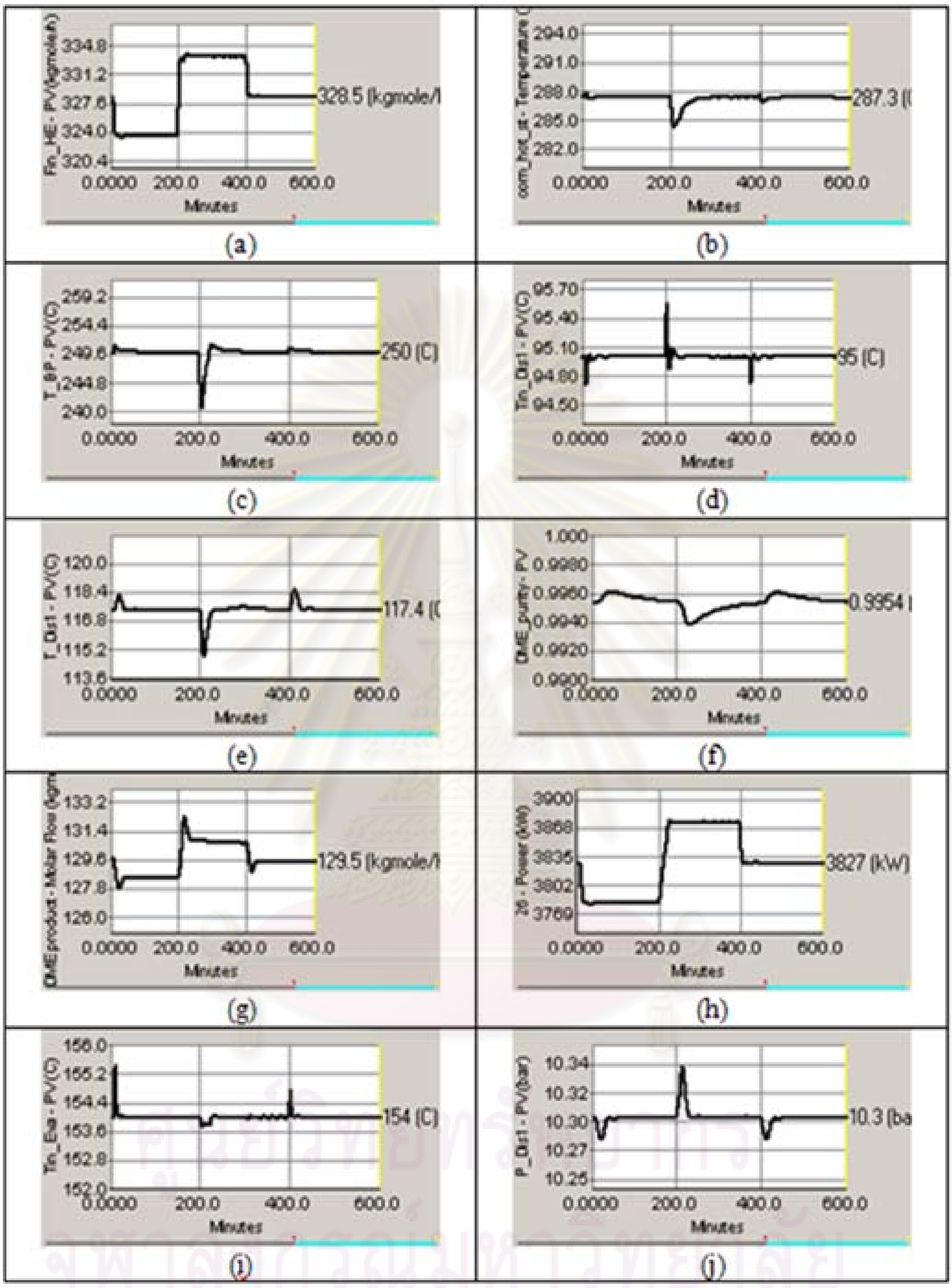

Figure 5.55 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS3, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

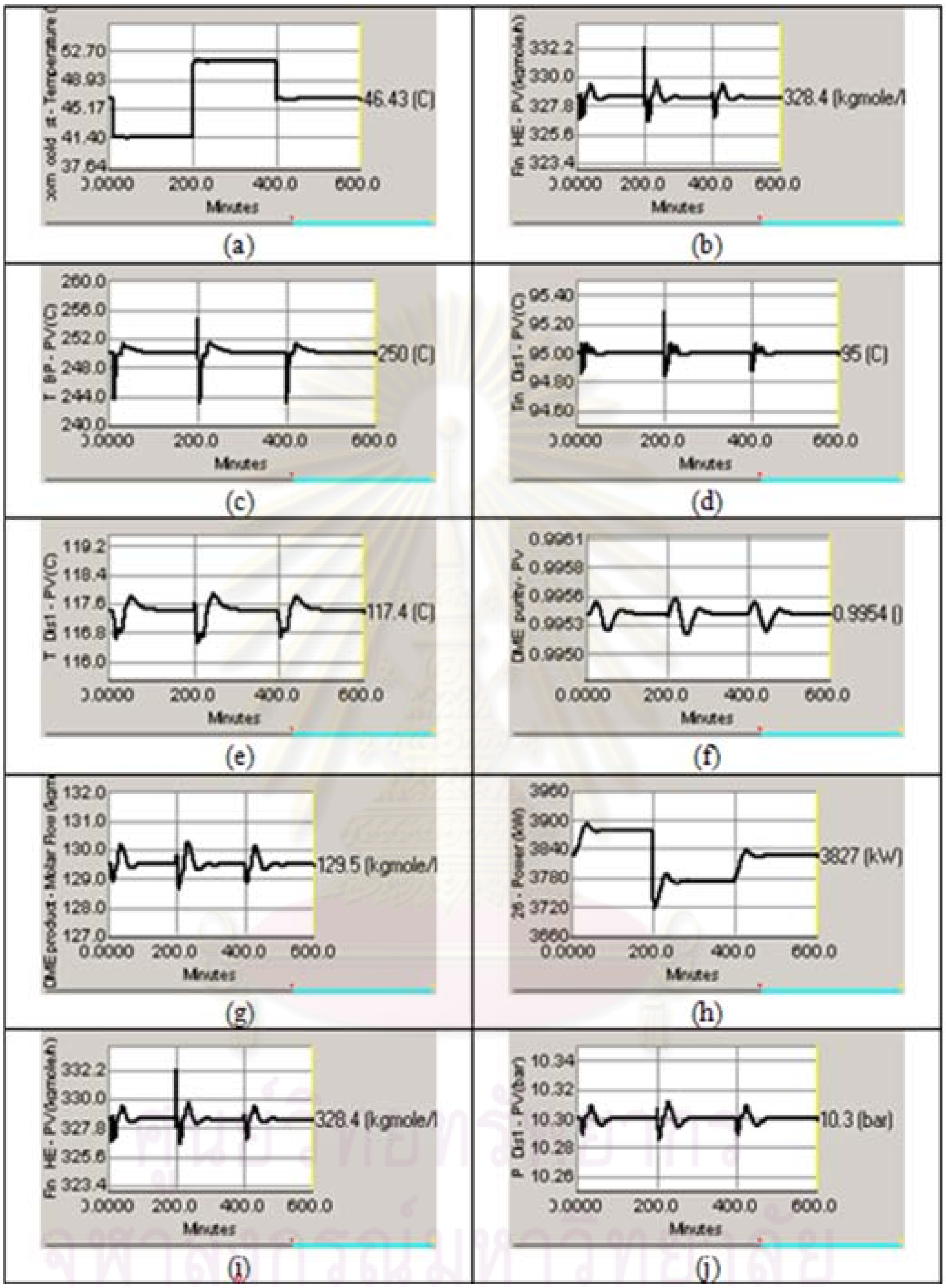

Figure 5.56 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS4, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

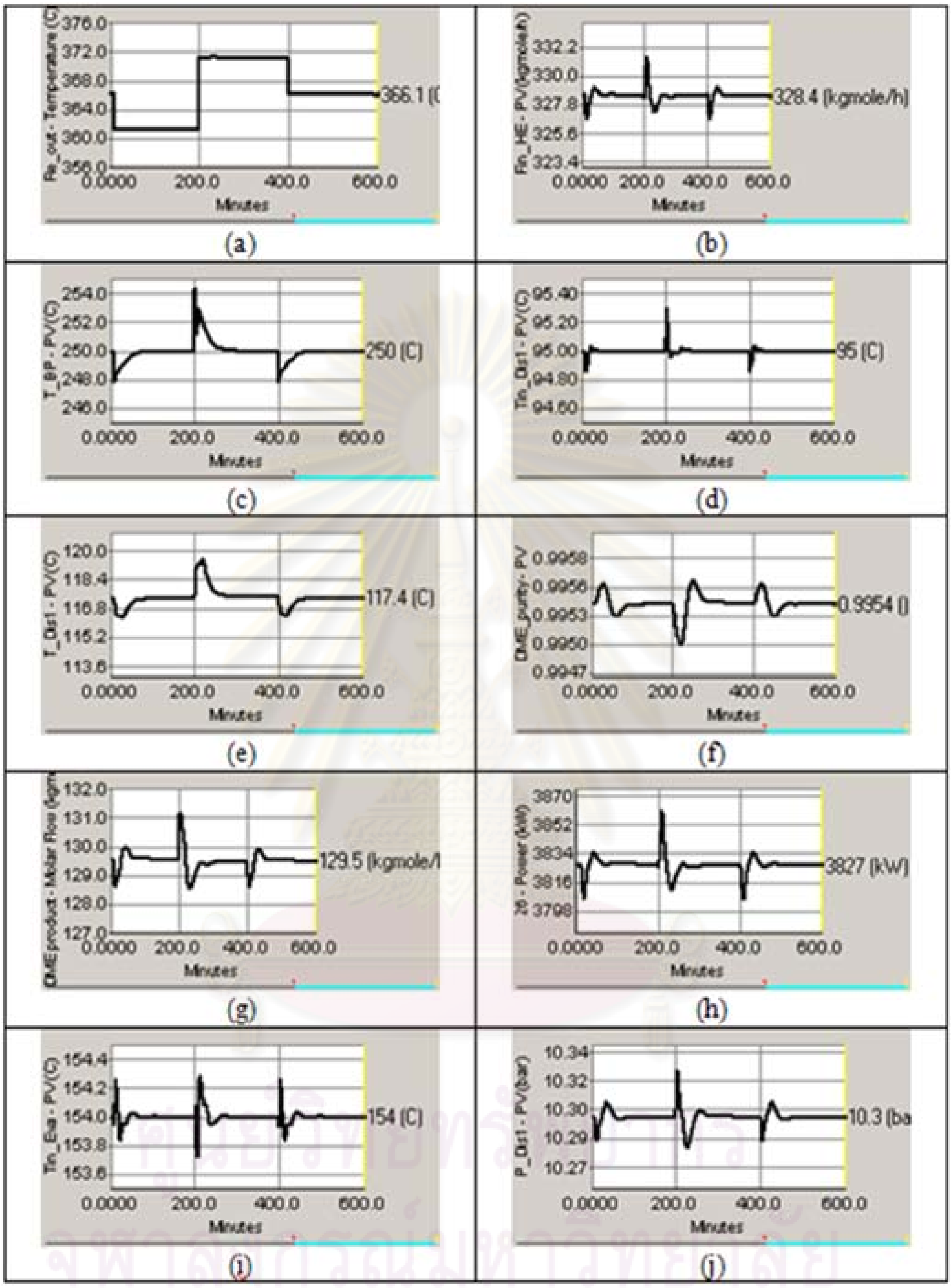

Figure 5.57 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS4, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

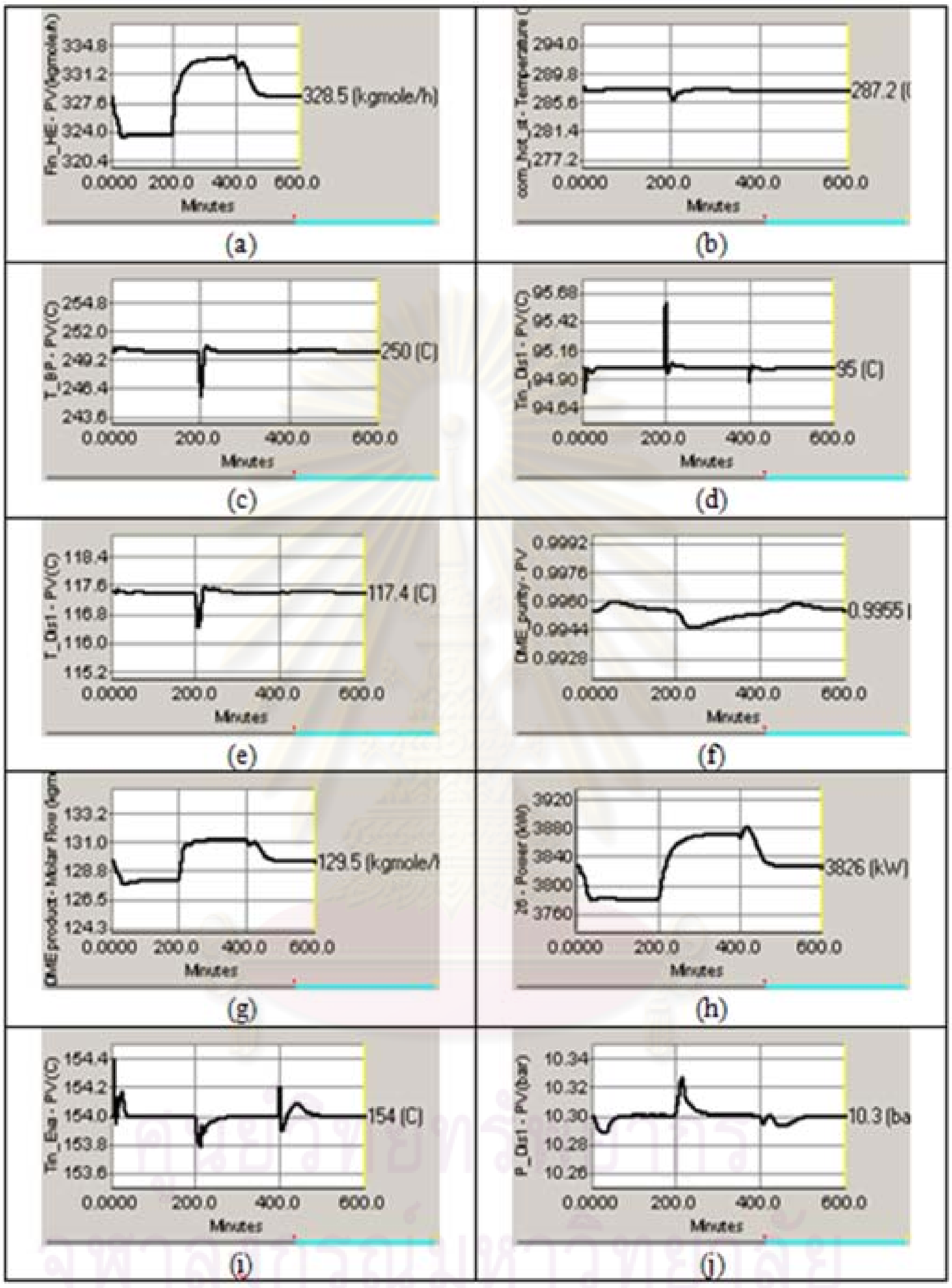

Figure 5.58 Dynamic Responses of the DME Process Alternative 2 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS4, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

5.4.4 Dynamic Simulation Results of the DME Process Alternative 3

Three disturbance loads are used to evaluate the dynamic performance of the base control structure CS1 and design control structures from Luyben's heuristic CS2 and fixture point method CS3 and CS4 for DME process base case.

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

Figures 5.59, 5.62, 5.65 and 5.68 show the dynamic responses of the control systems of DME process for alternative 3 to a change in the heat load disturbance of cold stream (reactor feed stream). In order to make this disturbance, first the temperature of combine stream between feed stream and recycle is decreased from 46.40C to 41.40C occurring at time equals 10 minutes, and the temperature is increased from 41.40C to 51.40C occurring at time equals 200 minutes, then its temperature is returned to its nominal value of 41.40C occurring at time equals 400 minutes (Figures 5.59a, 5.62a, 5.65a and 5.68a).

This alternative has been managed heat load propagation of cold stream same as alternative 1 to shift the heat load to FEHE outlet of cold stream and control temperature of FEHE outlet of hot stream by adjusting bypass valve and reject heat load disturbance of the cold stream is the same as base case by vaporizer duty adjusting figures 5.59h, 5.62h, 5.65h and 5.68h. The dynamic responsibility of this alternative is better than the base case because the cold stream is pre-heat by FEHE before send to vaporizer so the disturbance makes a little liquid at stream outlet of vaporizer at short period. However CS4 can control temperature loop, production rate and DME product purity better than CS1, CS2 and CS3 because CS4 control pressure of steam before feed to FEHE, it makes temperature control smoothly at vaporizer outlet.

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

Figures 5.60, 5.63, 5.66 and 5.69 show the dynamic responses of the control systems of DME process of alternative 3 to a change in the heat load disturbance of hot stream (Reactor Product stream).

In order to make this disturbance, first the temperature of the reactor outlet is decreased from 365.70C to 360.70C occurring at time equals 10 minutes, and the temperature is increased from 360.70C to 370.70C occurring at time equals 200 minutes, then its temperature is returned to its nominal value of 365.720C occurring at time equals 400 minutes (figures 5.60a, 5.63a, 5.66a and 5.69a). The disturbances propagation of the hot stream is the same as alternative 2 that shift disturbance to the hot stream outlet of FEHE and reject disturbance by cooler. This alternative control temperature of product column by bypass valve so the deviation of temperature larger than the base case and alternative 1as shown in figures 5.60e, 5.63e, 5.66e and 5.69e. However when compare between the control structures that the dynamic response of CS3 and CS4 are smoother than CS1, CS2 because CS3 and CS4 control both product column temperature and DME product purity but CS1 control only product column temperature and CS2 control only DME product purity.

5.4.4.3 Change in the Flow rates of Main Process stream

Figures 5.61, 5.64, 5.67 and 5.70 show the dynamic responses of the control systems of DME process alternative 3 to a change in the flow rates of main process stream. This disturbance is made by decreasing flow rates from 328.5 kgmol/h to 323.5 kgmol/h occurring at time equals 10 minutes, and the flow rates is increased from 323.5 kgmol/h to 333.5 kgmol/h occurring at time equals 100 minutes, then its flow rates is returned to its nominal value of 328.5 kgmol/h occurring at time equals 200 minutes (Figures 5.61a, 5.64a, 5.67a and 5.70a).

The dynamic result same as Base Case alternative 1 and alternative 2 that the drop in flow rates reduces the reaction rate, so the production rate is decreased but DME product purity increased. The CS4 can control the temperature of product column better than CS1, CS3 and CS4 so deviation of DME product quality from nominal value in CS4 smoother than CS1, CS2 and CS3.

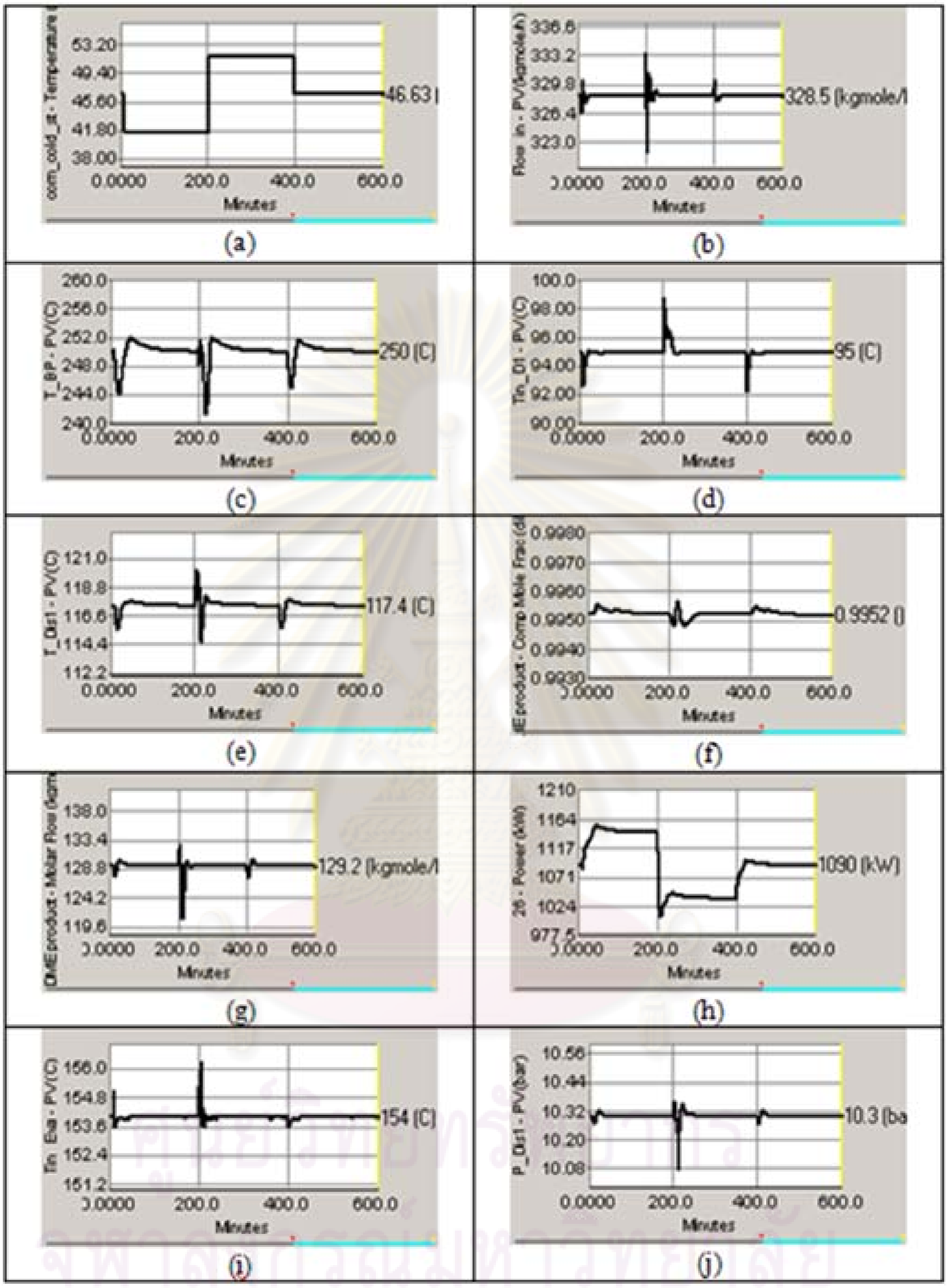

Figure 5.59 Dynamic Responses of the DME Process Alternative 3 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS1, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

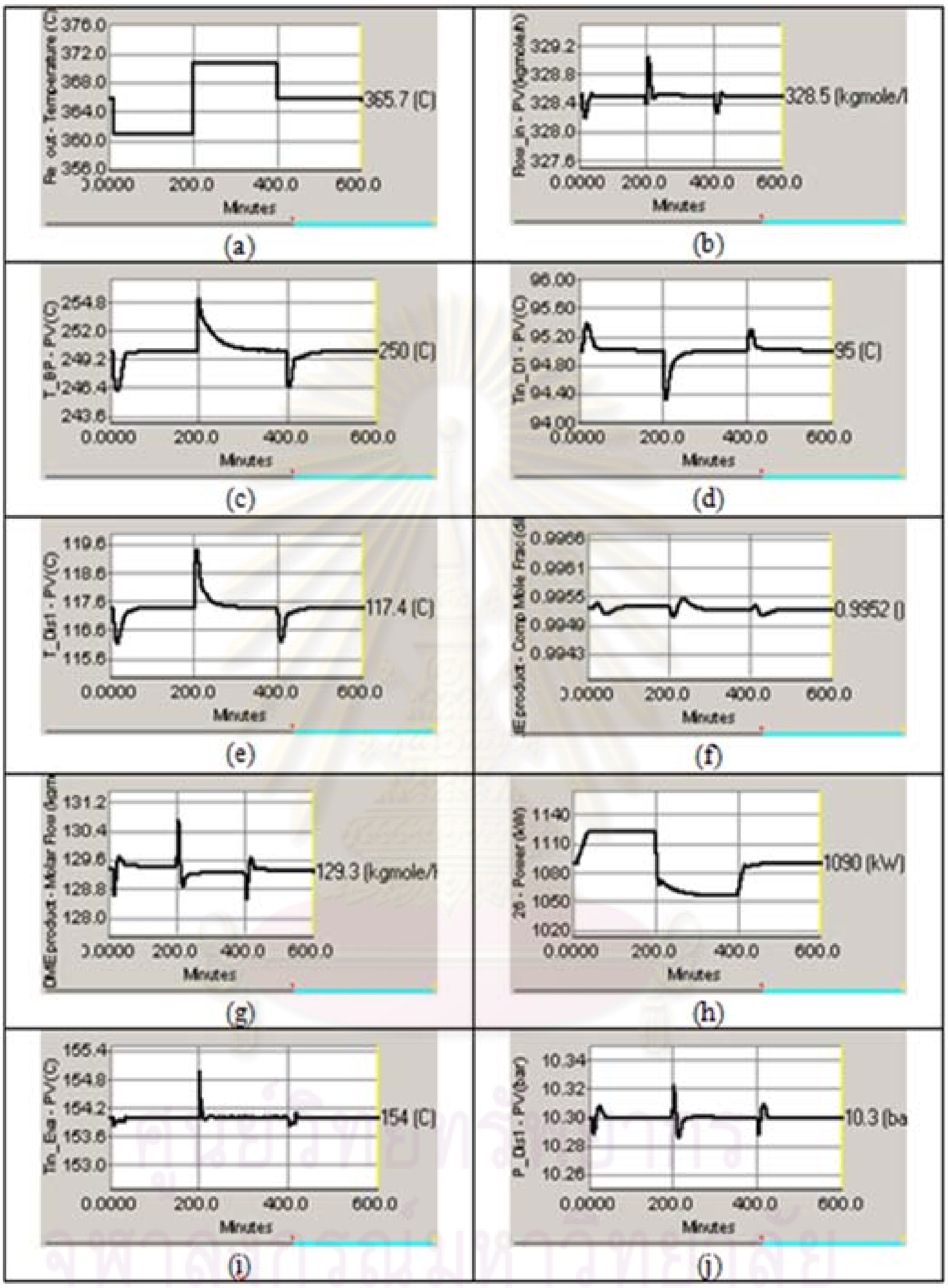

Figure 5.60 Dynamic Responses of the DME Process Alternative 3 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS1, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

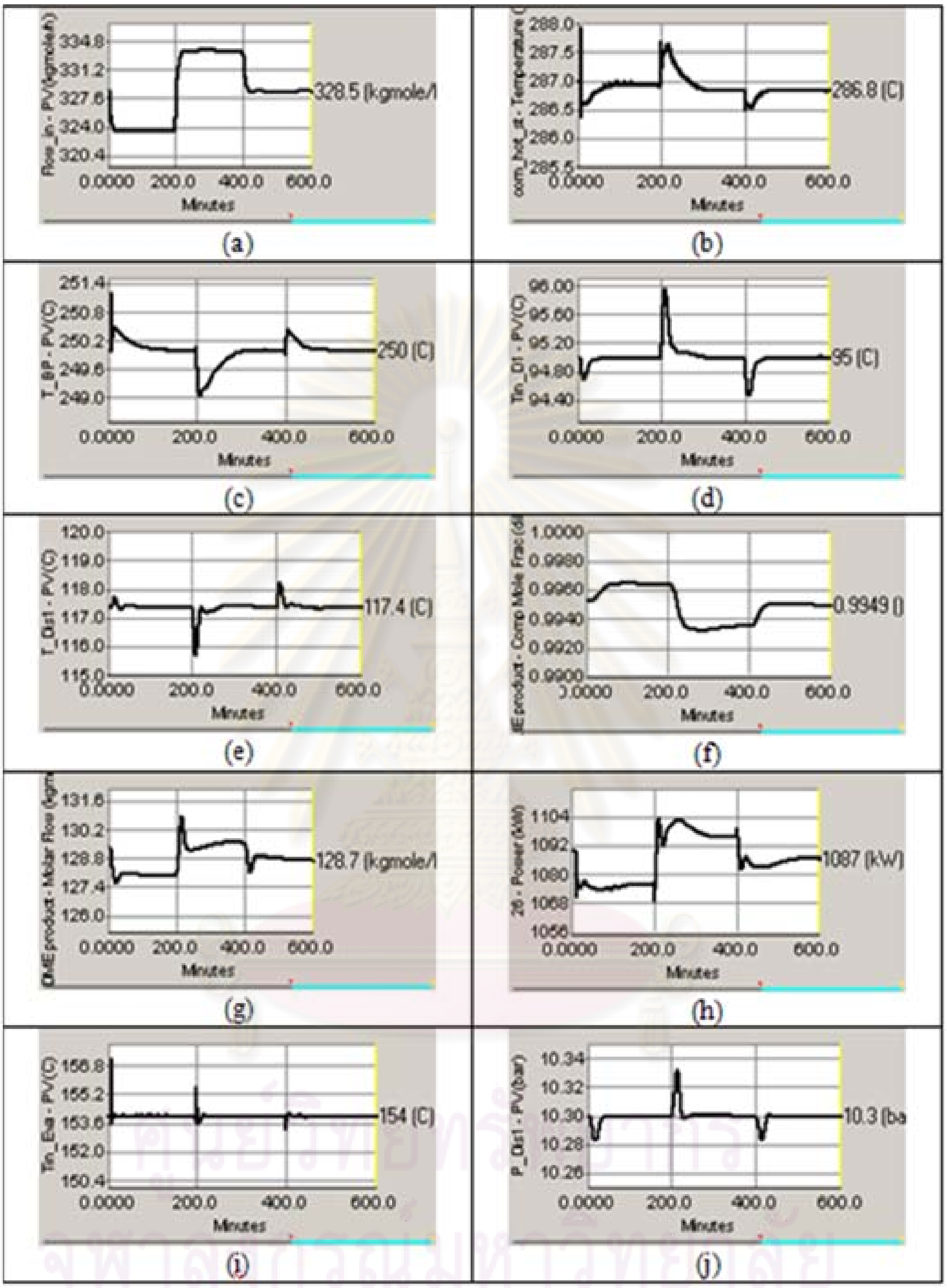

Figure 5.61 Dynamic Responses of the DME Process Alternative 3 to Change the flow rate Disturbance of process stream: CS1, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

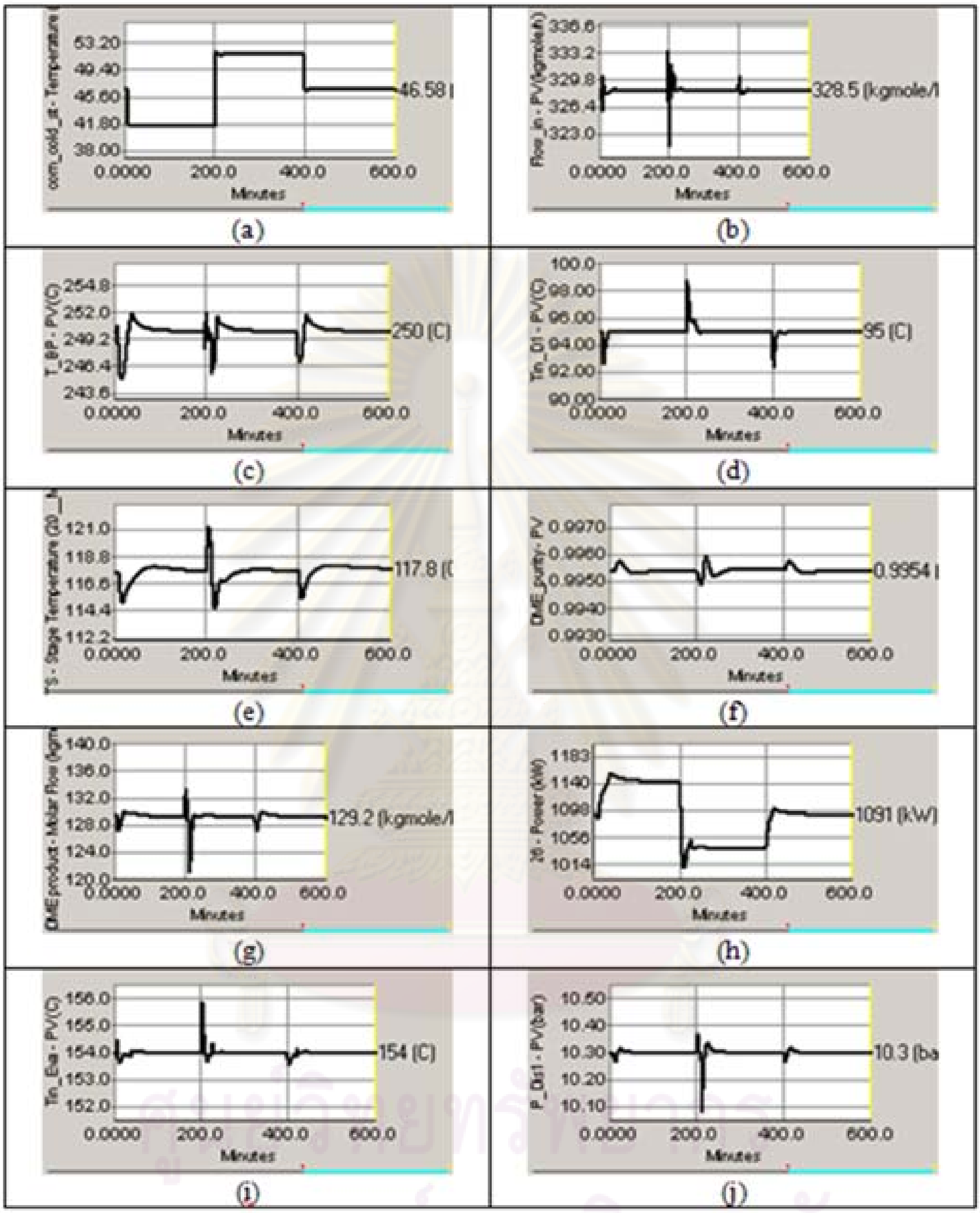

Figure 5.62 Dynamic Responses of the DME Process Alternative 3 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS2, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

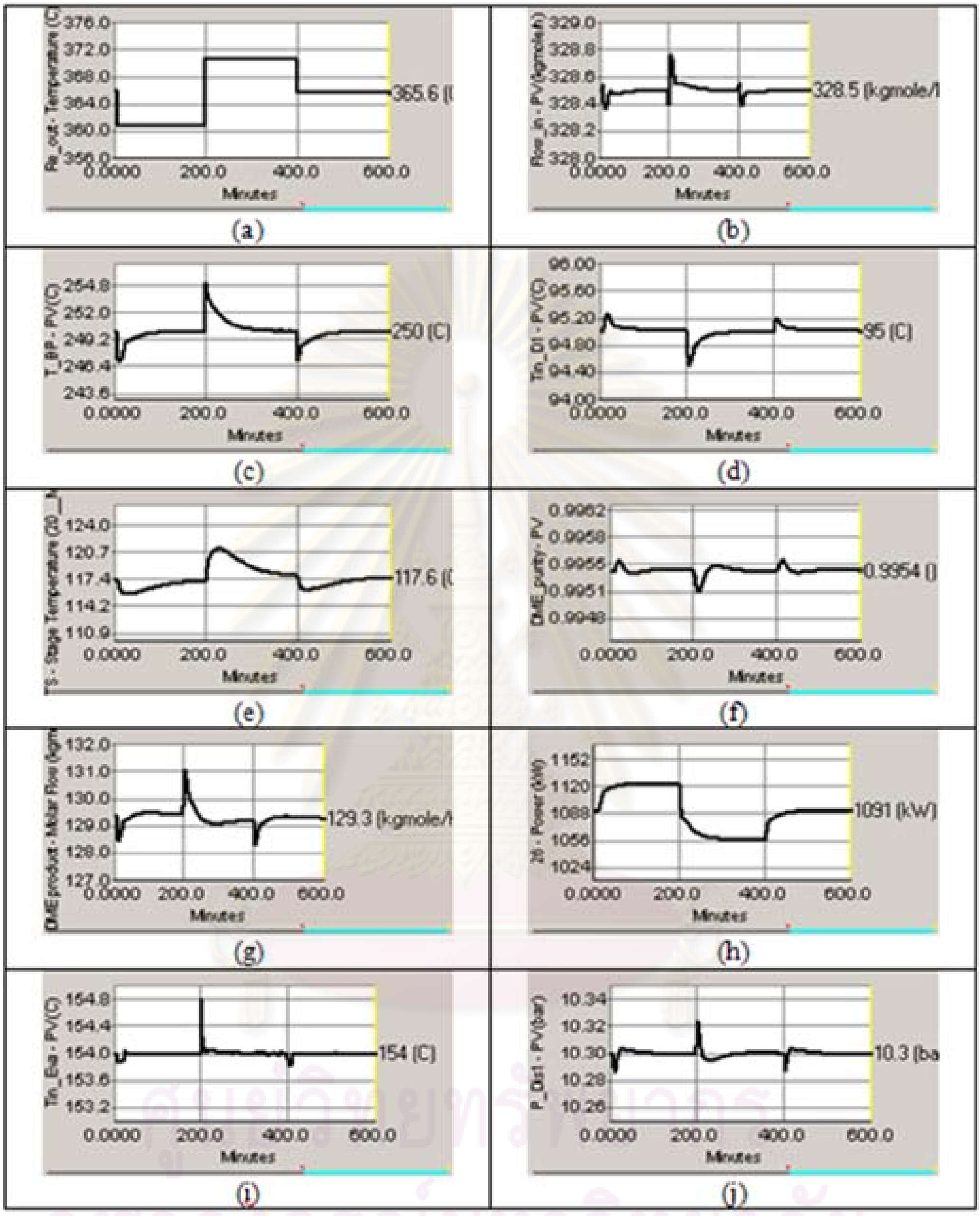

Figure 5.63 Dynamic Responses of the DME Process Alternative 3 to Change the Heat Load Disturbance of Hot Stream (Reactor Product Stream): CS2, where: (a) reactor outlet stream temperature, (b) fresh feed flow rate, (c) reactor inlet temperature , (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

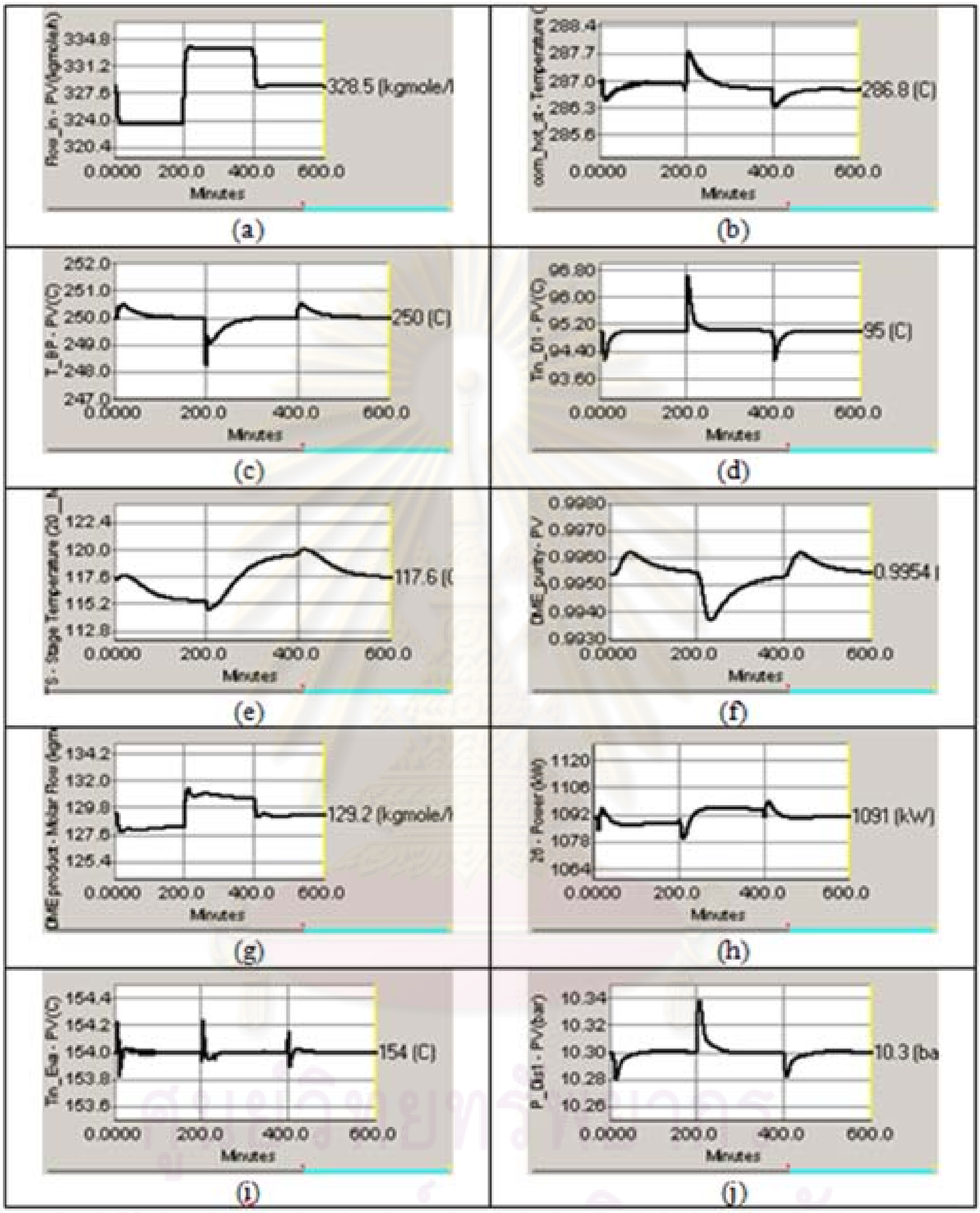

Figure 5.64 Dynamic Responses of the DME Process Alternative 3 to Change the flow rate Disturbance of process stream: CS2, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

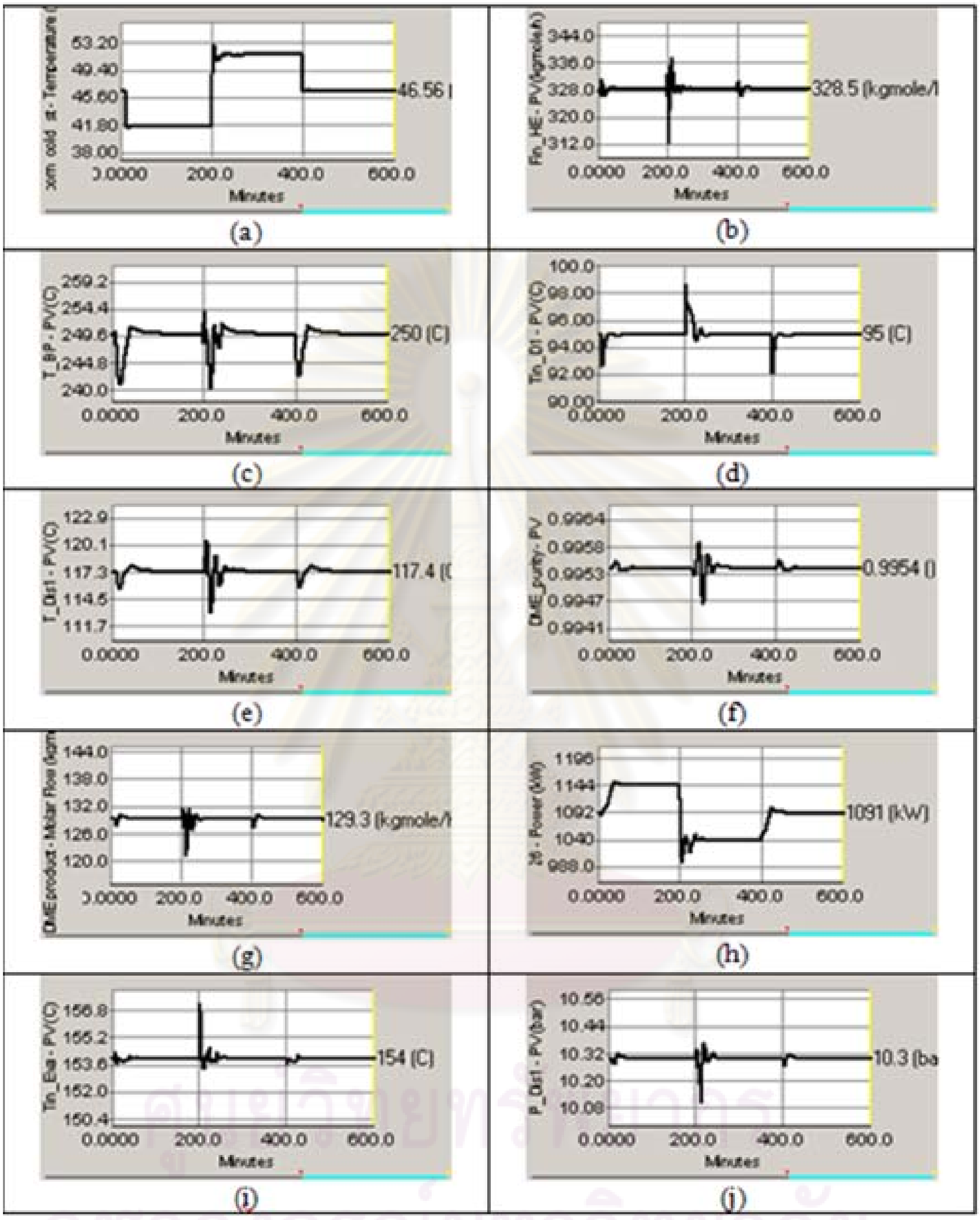

Figure 5.65 Dynamic Responses of the DME Process Alternative 3 to Change the flow rate Disturbance of process stream: CS2, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

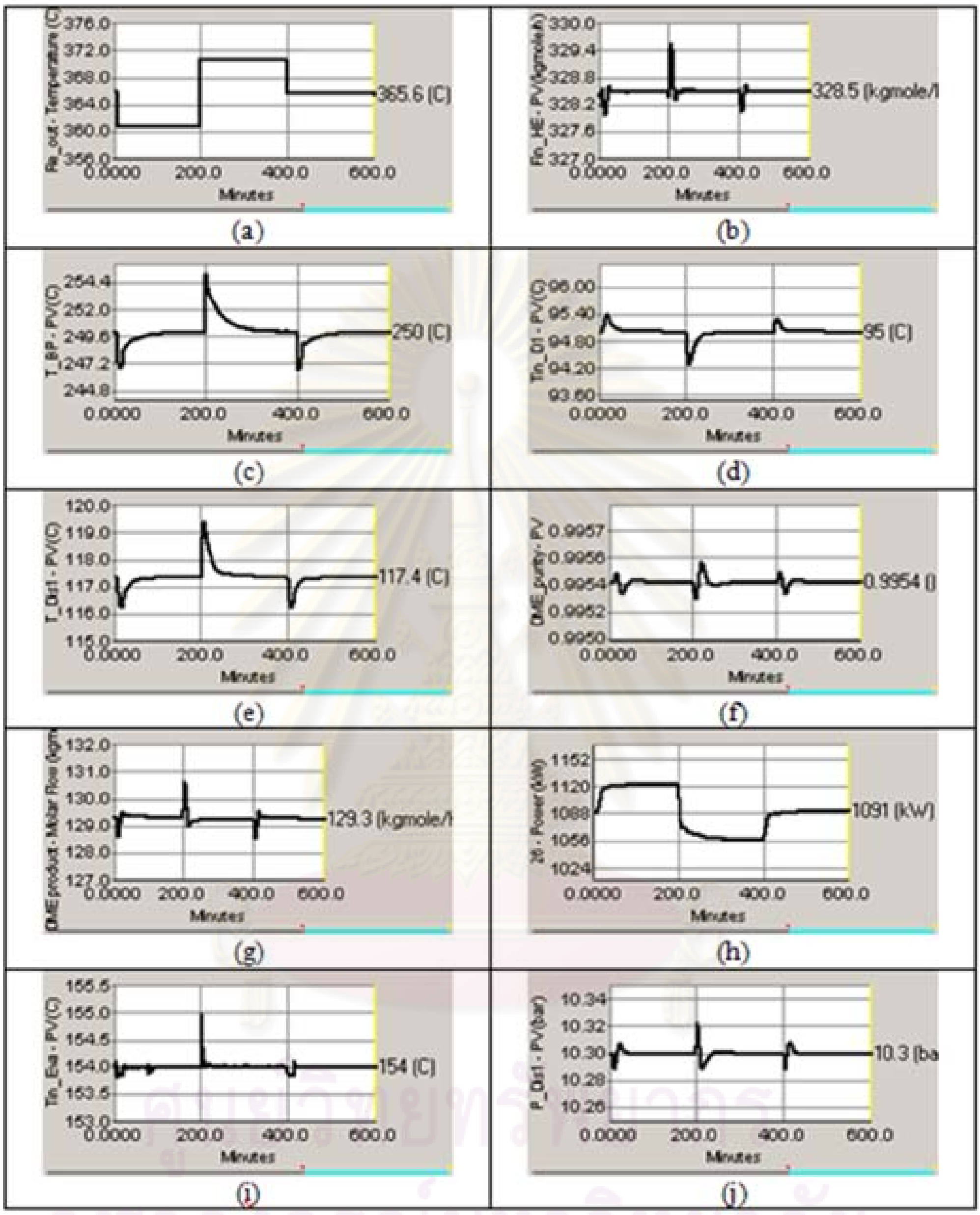

Figure 5.66 Dynamic Responses of the DME Process Alternative 3 to Change the flow rate Disturbance of process stream: CS2, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

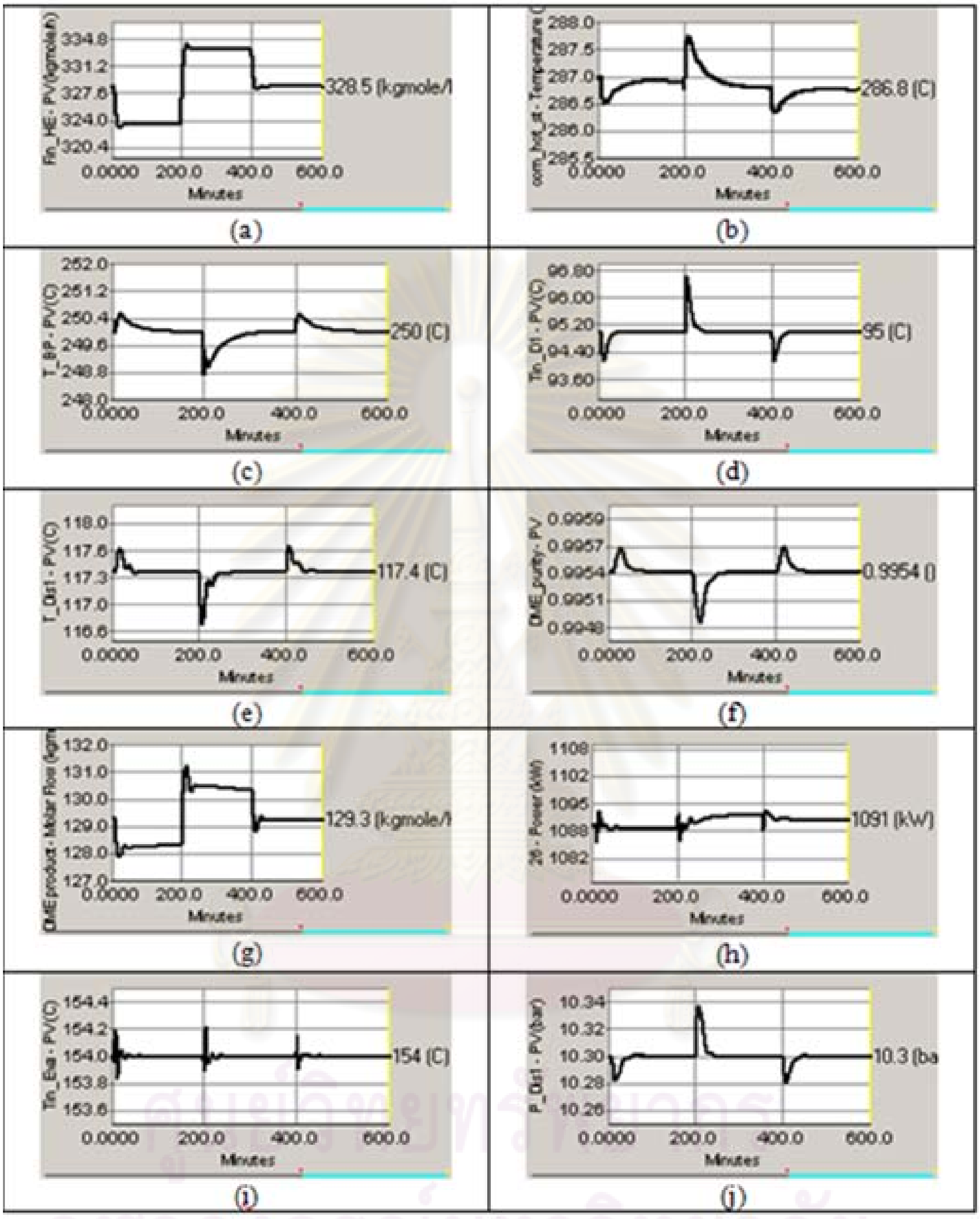

Figure 5.67 Dynamic Responses of the DME Process Alternative 3 to Change the flow rate Disturbance of process stream: CS3, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

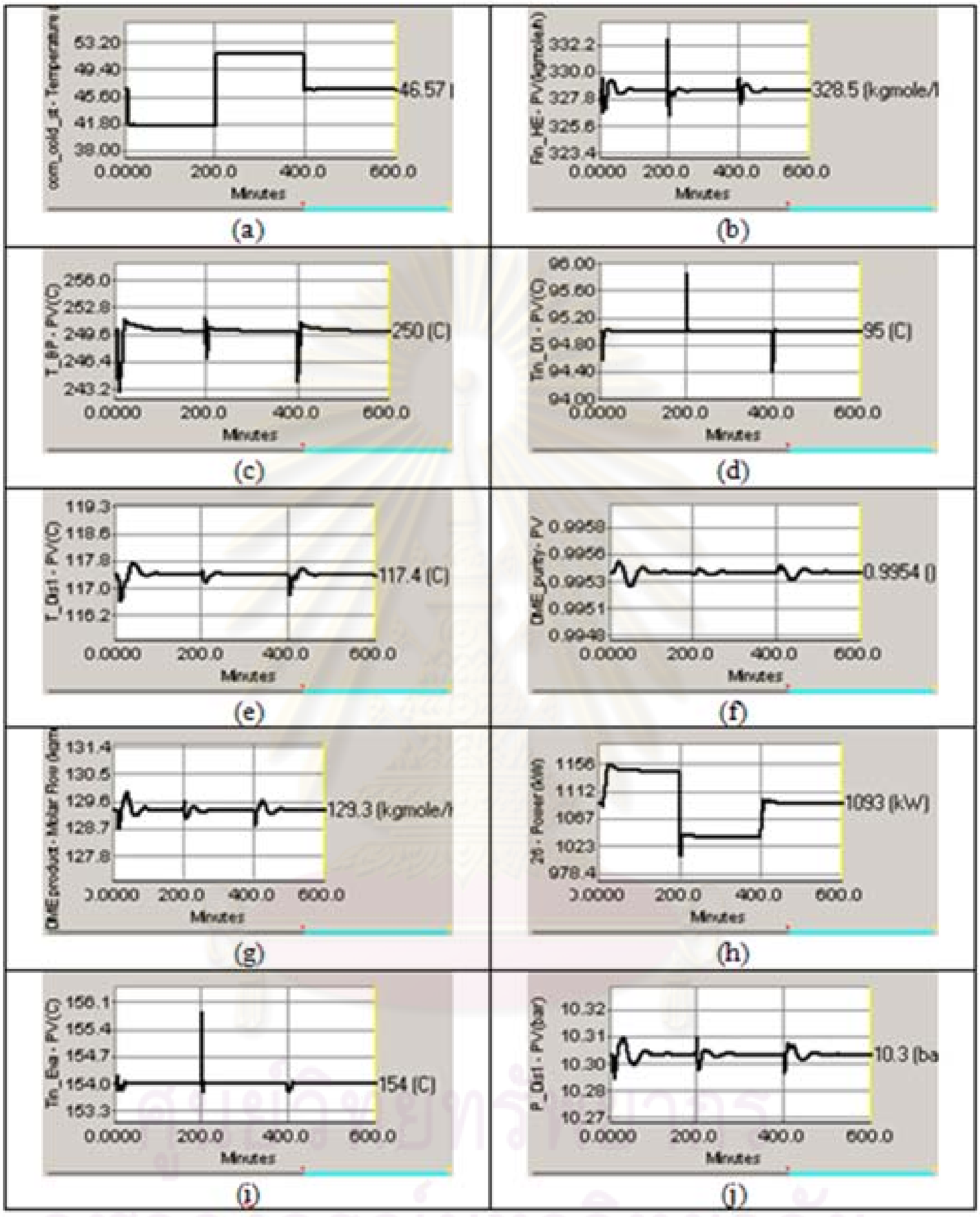

Figure 5.68 Dynamic Responses of the DME Process Alternative 3 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS4, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

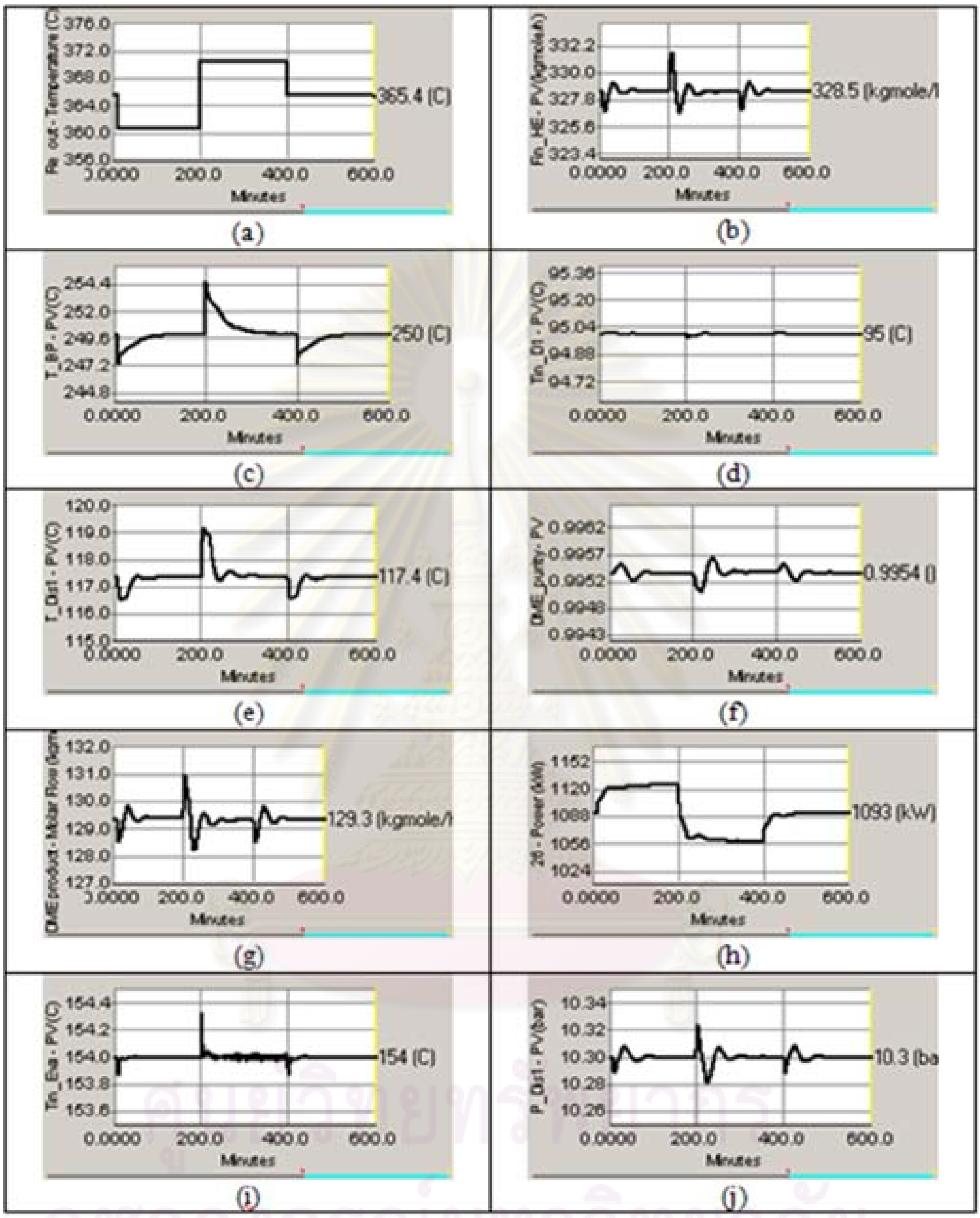

Figure 5.69 Dynamic Responses of the DME Process Alternative 3 to Change the Heat Load Disturbance of Cold Stream (Reactor Feed Stream): CS4, where: (a) combination of fresh feed and recycle stream temperature, (b) Fresh feed flow rate, (c) Reactor inlet temperature , (d) Product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

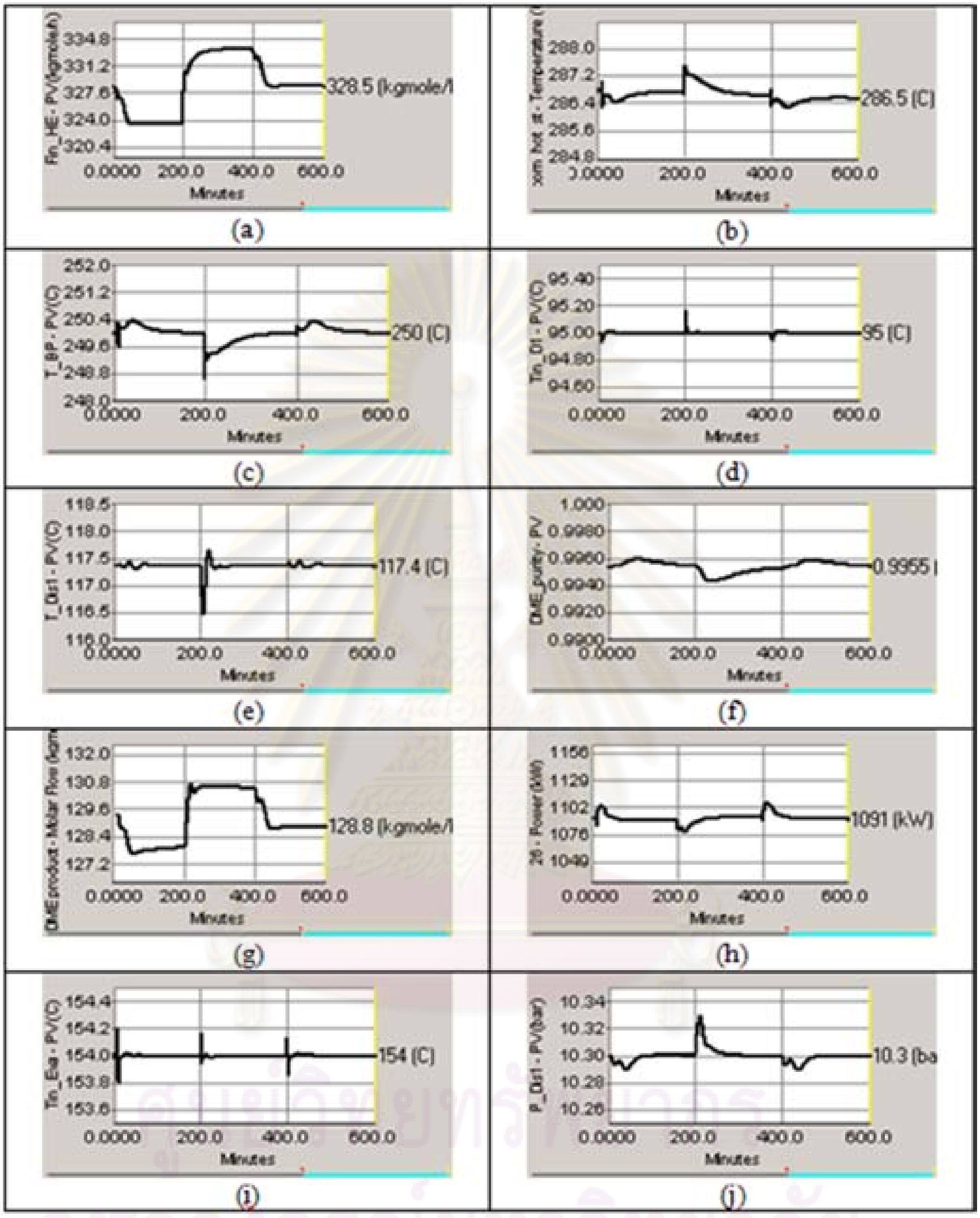

Figure 5.70 Dynamic Responses of the DME Process Alternative 3 to Change the flow rate Disturbance of process stream: CS4, where: (a) process stream flow rate, (b) reactor outlet stream temperature, (c) reactor inlet temperature, (d) product column inlet temperature, (e) product column stage temperature, (f) DME product purity, (g) DME production flow, (h) vaporizer power usage, (i) vaporizer outlet temperature, (j) product column pressure.

5.5 Evaluation of the Dynamic Performance

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

$$\text{IAE} = \int |\epsilon(t)|dt$$

Note that ϵ(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 5.1a to Table 5.4c show the IAE result for the change in the disturbance load including cold steam, hot stream and process flow stream in Base Case, Table 5.2a to Table 5.2c show the IAE result for the change in the disturbance load of alternative 1, Table 5.3a to Table 5.3c show the IAE result for the change in the disturbance load of alternative 2and Table 5.3a to Table 5.3c show the IAE result for the change in the disturbance load of alternative 3

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 process stream flow rates, the value of IAE in process CS4 is smaller than the other control structures so the controllability performance of CS is better than CS1, CS2 and CS3

Table 5.1a The IAE result of the DME process in the base case to a change in the disturbance load of cold stream (reactor feed stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	0.57025	0.00000	1.02570	1.40404
T-BP	0.61350	1.13097	1.15717	1.09836
Tin-dis1	2.09282	0.65296	0.65360	0.60062
Tin-vap	1.51666	0.76526	0.84473	0.87334
DME-purity	0.00000	1.17339	0.09034	1.73627
Sum	4.79324	3.72259	3.77155	5.71263
Average	1.19831	0.93065	0.75431	1.14253

Table 5.1b The IAE result of the DME process in the alternative 1 a change in the disturbance load of hot stream (reactor product stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	0.56396	0.00000	0.77301	1.66303
T-BP	1.55764	0.67438	0.67487	0.69310
Tin-dis1	1.86525	0.67196	0.62055	0.54224
Tin-vap	1.51616	0.42761	0.31708	0.83915
DME-purity	0.00000	1.10783	0.09486	1.39730
Sum	5.50301	2.88179	2.48037	5.13483
Average	1.37575	0.72045	0.49607	1.02697

Table 5.1c The IAE result of the DME process in the alternative 1 a change in the disturbance load of hot stream (reactor product stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	0.93449	0.00000	1.05329	1.01222
T-BP	1.11153	1.15142	1.32694	0.41011
Tin-dis1	2.08004	0.53882	0.44550	0.43564
Tin-vap	1.88248	0.59805	0.48092	0.63855
DME-purity	0.00000	1.65772	0.30292	1.03936
Sum	6.00854	3.94601	3.60957	3.53588
Average	1.50214	0.98650	0.72191	0.70718

Table 5.2a The IAE result of the DME process in the alternative 1 a change in the disturbance load of hot stream (reactor product stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	0.80065	0.00000	1.38815	0.81119
T-BP	1.15192	1.26619	0.38759	0.79429
Tin-dis1	1.26295	1.27871	1.37328	0.78507
Tin-vap	1.28953	1.14890	0.23175	0.62983
DME-purity	0.00000	1.47253	1.47232	0.75516
Sum	4.50505	5.16632	4.85309	3.77553
Average	1.12626	1.29158	0.97062	0.75511

Table 5.2b The IAE result of the DME process in the alternative 1 a change in the disturbance load of hot stream (reactor product stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	0.93940	0.00000	1.10949	0.95111
T-BP	0.84117	1.12426	0.91187	0.92270
Ti-dis1	1.19573	1.23266	1.11116	0.16044
Tin-vap	1.06467	1.19439	0.22539	0.31554
DME-purity	0.00000	0.80512	0.73784	0.15703
Sum	4.04098	4.35644	4.09576	2.50682
Average	1.01024	1.08911	0.81915	0.50136

Table 5.2c The IAE result of the DME process in the alternative 1 to a change in the flow rate of process stream

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	0.86098	0.00000	0.89572	1.24330
T-BP	0.69134	0.95802	0.68567	0.66497
Tin-dis1	0.73613	1.50900	1.36329	0.09158
Tin-vap	1.11370	1.22353	0.11550	0.14726
DME-purity	0.00000	0.89710	1.19041	0.91249
Sum	3.40215	4.58765	4.25060	3.05960
Average	0.85054	1.14691	0.85012	0.61192

Table 5.3a The IAE result of the DME process in the alternative 2 to a change in the disturbance load of cold stream (reactor feed stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	0.49792	0.00000	1.04615	0.45593
T-BP	0.49062	0.90348	1.48682	0.91907
Tin-dis1	0.46740	0.57814	1.35212	1.02338
Tin-vap	0.70940	0.94664	1.05469	0.68927
DME-purity	0.00000	0.68361	0.73454	0.58185
Sum	2.16534	3.11188	5.67432	3.66951
Average	0.54133	0.77797	1.13486	0.73390

Table 5.3b The IAE result of the DME process in the alternative 2 a change in the disturbance load of hot stream (reactor product stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	1.14510	0.00000	0.67019	0.84704
T-BP	0.99568	1.00091	1.00005	0.90337
Tin-dis1	0.74522	0.71437	1.14884	0.79158
Tin-vap	0.54419	0.19148	0.97309	0.69124
DME-purity	0.00000	1.20013	0.48237	1.01749
Sum	3.43019	3.10689	4.27453	4.25071
Average	0.85755	0.77672	0.85491	0.85014

Table 5.3c The IAE result of the DME process in the alternative 3 to a change in the flow rate of process stream

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	1.58811	0.00000	1.02013	0.39176
T-BP	0.53139	1.19520	1.63567	0.63774
Tin-dis1	0.78161 1	.18850	1.28547	0.74441
Tin-vap	1.11517	1.01399	1.01457	0.85627
DME-purity	0.00000	1.12570	1.06158	0.81272
Sum	4.01628	4.52339	6.01742	3.44290
Average	1.00407	1.13085	1.20348	0.68858

Table 5.4a The IAE result of the DME process in the alternative 3 to a change in the disturbance load of cold stream (reactor feed stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	1.24585	0.00000	1.44532	1.30883
T-BP	1.35307	0.76412	1.36364	1.51917
Tin-dis1	1.33737	1.23105	1.38750	0.04408
Tin-vap	1.12644	1.12961	1.42148	1.32247
DME-purity	0.00000	1.79030	0.82536	0.38434
Sum	5.06273	4.91508	6.44329	4.57890
Average	1.26568	1.22877	1.28866	0.91578

Table 5.4b The IAE result of the DME process in the alternative 3 a change in the disturbance load of hot stream (reactor product stream)

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	0.97547	0.00000	0.99348	1.03105
T-BP	0.99976	1.00107	0.99929	0.99988
Tin-dis1	1.30489	1.27452	1.37060	0.04998
Tin-vap	1.25877	1.09152	1.19407	0.45565
DME-purity	0.00000	0.99970	0.44078 1	.55952
Sum	4.53890	4.36681	4.99822	4.09608
Average	1.13472	1.09170	0.99964	0.81922

Table 5.4c The IAE result of the DME process in the alternative 1 to a change in the flow rate of process stream

Controller	Integral Absolute Error (IAE)			
	CS1	CS2	CS3	CS4
T-dis1	1.59929	0.00000	0.83620	0.56450
T-BP	0.90471	1.02804	1.03094	1.03631
Tin-dis1	1.07625	1.52970	1.35704	0.03701
Tin-vap	1.81856	1.05140	0.69710	0.43294
DME-purity	0.00000	1.59704	0.16152	1.24144
Sum	5.39881	5.20619	4.08280	3.31220
Average	1.34970	1.30155	0.81656	0.66244

CHAPTER VI

CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusion

In this thesis, we considered two main objectives that to design heat integrated process and design control structure of the DME process that given by Turton (1998). First objective to design heat integrated process of DME process for three alternatives by using disturbance load propagation method (Wongsri,1990) to minimize energy usage when compare with Base Case. From steady state simulation result by HYSYS we can save the energy usage 58.78% for alternative 1, 15.55% for alternative 2 and 58.77% for alternative 3 compare with Base Case

Second objective, we considered the plantwide control structure of DME process to matching with heat integration alternatives. The plantwide control structures are carry out in to four control structure, control structure 1(CS1) is design by Turton (1998) for base control structure, control structure 2 (CS2) is designed follow nine step approach of Luyben and co-workers, control structure 3 (CS3) and control structure 4 (CS4) are designed follow the fixture point method to make a good controllability and maintaining good control performance.

The plantwide control structures are designed using the disturbance load propagation method and HEN design follows resilient HEN synthesis method (Wongsri, 1990) and heat pathway heuristics,(HPH) (Wongsri and Hermawan, 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 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 result of control structures CS3 and CS4 are nearly and CS1 give the result near CS2 because CS3 have a control structure similarly CS4 and CS1 have a control structure similarly CS2, however CS4 give the best performance of control structure to minimize the IAE score. But IAE score of CS2 to high because CS2 exhibited very slow dynamics and is more sensitive to the disturbances.

The heat-integrated plants of DME 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.

6.2 Recommendations

1. Study and design the highly complex heat-integrated plants of DME process point of view.

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

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

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APPENDICES

APPENDIX A

PROCESS STREAM DATA FOR SIMULATION AND EQUIPMENT DATA SPECIFICATION OF THE DME PLANT

Table A.1 Process Stream Data for Base Case of DME Process

Stream name	**Methanol feed (1-1)**	**Pump out (1-2)**	**Vap_feed (3)**	**Vap_out (4)**	**HEN_feed (4-1)**
Temperature [^0C]	25.00	25.87	46.33	154.00	154.00
Pressure [bar]	1.00	16.00	15.50	15.20	15.10
MolarFlow[kgmole/hr]	262.20	262.20	328.51	328.51	328.51
Methanol, mole fraction	0.9905	0.9905	0.9841	0.9841	0.9841
Water, mole frection	0.0095	0.0095	0.0114	0.0114	0.0114
DME mold fraction	0.0000	0.0000	0.0045	0.0045	0.0045

Stream name	**Reac_feed (5)**	**Reac_out (6)**	**Cooler_inlet (7)**	**Dis1_inlet (8)**	**DME_product (10)**
Temperature [^0C]	250.00	365.96	287.09	95.00	46.35
Pressure [bar]	14.70	13.90	13.80	13.40	10.30
MolarFlow[kgmole/hr]	328.51	328.51	328.51	328.51	129.7
Methanol, mole fraction	0.9841	0.1982	0.1982	0.1982	0.0046
Water, mole frection	0.0114	0.4044	0.4043	0.4043	0.0000
DME mold fraction	0.0045	0.3974	0.3974	0.3974	0.9954

Stream name	**Bott_dis1 (11)**	**Dis2_feed (12)**	**Dis2_distill (16)**	**Bott_dis2 (14)**
Temperature [^0C]	151.04	138.36	121.69	166.11
Pressure [bar]	10.40	7.38	7.30	7.45
MolarFlow[kgmole/hr]	198.81	198.81	66.29	132.51
Methanol, mole fraction	0.3246	0.3246	0.9590	0.0071
Water, mole frection	0.6681	0.6681	0.0191	0.9929
DME mold fraction	0.0073	0.0073	0.0219	0.0000

Table A.2 Equipment data and Specifications of heat-integrated plant of DME process

Equipments	Specifications	Heat integrated process of DME process			
		BC	Alternative 1	Alternative 2	Alternative 3
Reactor	Diameter (m)	0.750	0.750	0.750	0.750
	Length (m)	13.475	13.475	13.475	13.475
	Number of tube	1	1	1	1
Vaporizer	Tube volume (m^3)	0.1	0.1	0.1	0.1
Cooler	Tube volume (m^3)	0.1	-	0.1	-
FEHE1	Shell volume (m^3)	2.272	2.272	2.272	2.272
	ube volume (m^3)	0.193	0.193	0.193	0.193
	UA (kJ/C-h)	1.69 x 10^4	1.69 x 10^4	1.69 x 10^4	1.69 x 10^4
FEHE2	Shell volume (m3)		2.272		2.272
	Tube volume (m^3)	-	0.193	-	0.193
	UA (kJ/C-h)		1.94 x 10^5		7.54x 10^5
Reboiler Column 1 (CC1)	Shell volume (m^3)			2.272	2.272
	Tube volume (m^3)	-	-	0.193	0.193
	UA (kJ/C-h)			7.53 x 10^4	7.54 x 10^4
Tank Bottom C1(TB1)	Vessel volume (m^3)	-	-	1.911	1.911

Table A.3 Column Specifications of DME process base case

Parameters columns	Product Column	Recycle Column
Model	Refluxed absorber	Refluxed absorber
Tray	Sieve Tray	Sieve Tray
Number of tray	22	26
Feed tray	11	14
Pressure (bar)	10.30	7.30
Diameter of Vessel (m)	1.500	5.348
Tray space (m)	0.5499	0.550
Weir length (m)	1.200	4.278
Weir height (m)	0.05	0.05
Specification 1	Distillate rate 129.7 kgmol/h	Distillate rate 66.3 kgmol/h
Specification 2	DME purity 0.9954	Water purity 0.959

APPENDIX B

PARAMETER TUNING OF CONTROL STRUCTURE

B.1 Turning Flow, Level, Pressure, Temperature Control.

Flow Controllers

The dynamics of flow measurement are fast. The time constants for moving control valves are small. Therefore, the controller can be tuned with a small integral or reset time constant τ_I. A value of $\tau_I = 0.3$ minutes works in most flow controllers. The value of controller gain should be kept modest because flow measurement signals are sometime noisy due to the turbulent flow through the orifice plate. A value of controller gain of Kc = 0.5 is often used.

Level Controllers

Most level controllers should use proportional-only action with a gain of 1 to 2. This provides the maximum amount of flow smoothing. Proportional control means there will be steady-state offset (the level will not be returned to its setpoint value). However, maintaining a liquid level at a certain value is often not necessary when the liquid capacity is simply being used as surge volume. So the recommended tuning of a level controller is Kc = 2.

Pressure Controllers

Setting the integral time equal to about 2 to 4 times the process time constant and using a reasonable controller gain usually gives satisfactory pressure control. Of course the gain used depends on the span of the pressure transmitter. Some simple step tests can be used to find the value of controller gain that yields satisfactory pressure control. Typical pressure controller tuning constants for columns and tanks are Kc = 2 and $\tau_I = 10$ minutes.

Temperature Controllers

Temperature dynamic responses are generally slow, so PID control is used. Typically, the controller gain, Kc, should be set between 2 and 10, the integral time, τ_I, should set between 2 and 10 minutes, and the derivative time τ_d, should be set between 0 and 5 minutes.

Table B.1 Parameter tuning of DME process

Controller	controlled variable	manipulated variable		Control action	Kc	Ti (min)	Td (min)
Reaction section							
Flow_ in	Process flow rate		Fresh feed valve (VLV-106)	Reverse	0.5	0.3	-
Fin_ FEHE1	Heat exchange feed flow rate		Main stream valve (VLV-107)	Reverse	0.5	0.3	-
Fin_ dis1	Product column feed flow rate	CS1, CS2, CS3	Main stream valve (VLV-101)	Reverse	0.5	0.3	-
Tout_ vap	Vaporizer outlet temperature		Vaporizer duty (Q26)	Reverse	0.050	0.220	0.049
T_ BP	Reactor inlet temperature		Bypass valve (VLV-100)	Direct	0.100	0.170	0.100
Tin_ dis1	Product column inlet temperature		Cooler duty (Q27)	Direct	0.100	0.100	0.050
Twaste	Waste water temperature		Cooler duty (Q28)	Direct	0.100	0.100	-
Separation section (Product Column)							
P_ dis1	Product column pressure		Condenser duty (Qcon1)	Direct	2	10	-
T_ dis1	Product column stage 20 temperature	CS1, CS2, CS3	Product column reboiler duty (Qreboil1)	Reverse	2	25	0.111
Lcon_ dis1	Product column condenser level		Distillate flow rate valve (VLV-102)	Direct	2	-	-
Lreboil_ dis1	Product column reboiler level	CS1	Bottom flow rate valve (VLV-103)	Direct	2	-	-
		CS2	reboiler duty (Qreboil1)	Direct	2	-	-
		CS3	Cascade control with Fin_dis2	Direct	2	-	-
		CS4	Main stream valve (VLV-101)	Reverse	2	-	-
Separation section (Recycle Column)							
P_dis2	Recycle column pressure		Recycle column condenser duty (Qcon2)	Direct	2	10	-
T_ dis2	Recycle column stage 24 temperature	CS1, CS3	Reboiler duty (Qreboil2)	Reverse	2	15	0.138
		CS4	Bottom flow rate valve (VLV-105)	Direct	4.06	26.70	5.94
Lcon_ dis2	Recycle column condenser level	CS1	Distillation flow rate valve (VLV-104)	Direct	2	-	-
		CS2, CS3, CS4	Reflux valve	Direct	2	-	-
Lreboil_ dis2	Recycle column reboiler level	CS1, CS2, CS3	Bottom flow rate valve(VLV-105)	Direct	2	-	-
		CS4	Reboiler duty (Qreboil2)	Direct	2	-	-

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APPENDIX C

FIXTURE POINT THEOREM DATA

Table C.1 List of Manipulate Variable for DME Process

Manipulate Variable	Description
VLV-100	Manipulate bypass stream of FEHE1
VLV-101	Manipulate flow inlet the product column
VLV-102	Manipulate distillate rate of product column
VLV-103	Manipulate bottom flow rate of product column
VLV-104	Manipulate distillate rate of recycle column (recycle flow)
VLV-105	Manipulate bottom flow rate of recycle column
VLV-106	Manipulate fresh feed flow
VLV-107	Manipulate flow inlet FEHE1
Flow top1	Manipulate reflux flow of product column
Flow top2	Manipulate reflux flow of recycle column
Qevap	Manipulate hot utility of vaporizer
Qcooler	Manipulate cold utility of cooler
Qcon1	Manipulate cold utility of product column condenser
Qcon2	Manipulate cold utility of recycle column condenser
Qreboil1	Manipulate hot utility of product column reboiler
Qreboil2	Manipulate hot utility of recycle column reboiler

Table C.2 IAE Result of Temperature Deviation at Process Stream

Steam	VLV -100	VLV -101	VLV -102	VLV -103	VLV -104	VLV -105	VLV -106	VLV -107	Flow top1	Flow top2	Qevap	Qcooler	Qcon1	Qcon2	Qreboil1	Qreboil2	SUM IAE
feed (1)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
1-2	0.0015	0.0049	0.0020	0.0026	0.0087	0.0008	0.0072	0.0019	0.0018	0.0030	0.0023	0.0045	0.0017	0.0027	0.0024	0.0031	0.0513
2	0.0014	0.0049	0.0020	0.0026	0.0087	0.0007	0.0073	0.0019	0.0018	0.0030	0.0023	0.0045	0.0016	0.0027	0.0024	0.0031	0.0510
com_cold_st	0.4019	0.2427	0.5337	0.5784	1.1728	0.4730	0.3832	0.3730	0.4331	0.9100	0.6462	0.4040	0.3895	0.7041	0.8311	0.9371	9.4137
4	1.2319	3.2685	1.0927	0.9942	1.2173	0.6352	2.7139	2.0313	1.1345	0.2603	1.4747	2.5174	1.3405	1.1989	0.1827	0.2944	21.5883
4-1	1.2377	3.3813	1.2234	0.9974	1.6025	0.6315	3.2499	2.1393	1.1957	0.3550	1.5022	2.5583	1.5382	1.2050	0.2136	0.4272	23.4582
5	0.9376	2.4860	1.1987	0.4502	0.8946	0.2560	2.4630	1.1000	0.6437	0.6495	1.9609	2.1894	0.8188	0.5578	0.5482	0.5275	17.6819
Re_out	2.4551	0.3141	0.2172	1.3978	1.0369	3.3287	1.2431	2.5568	1.6760	0.0892	0.0344	0.0606	2.6647	2.0652	0.0264	0.1169	19.2832
com_hot_st	3.2518	1.3448	0.6178	1.5558	1.3106	3.4258	1.6683	2.8274	1.8347	0.3928	0.9252	1.0460	2.8576	2.2630	0.2698	0.4017	25.9930
8	1.2176	1.5786	0.8871	0.4601	1.2072	1.4586	1.2800	1.2375	0.7353	0.5287	0.4690	0.7506	1.3141	0.8323	0.4795	0.6102	15.0465
9	0.5486	0.6448	0.3422	0.2305	0.5222	0.6986	0.5568	0.5785	0.3686	0.2599	0.2015	0.3033	0.7006	0.4080	0.2119	0.2436	6.8197
10	0.0953	0.2498	0.6928	0.0968	0.2756	0.0508	0.2249	0.1182	0.7652	0.1686	0.0999	0.1739	0.3907	0.0977	0.0969	0.0910	3.6880
11	0.9136	0.8296	1.4044	0.5695	0.8736	0.6633	0.6043	0.5549	1.6096	0.3895	0.4527	1.0730	0.5565	0.3755	0.5323	0.5416	11.9440
11-1	0.9136	0.8296	1.4044	0.5869	0.8733	0.6637	0.6042	0.5550	1.6088	0.3875	0.4526	1.0817	0.5569	0.3755	0.5330	0.5410	11.9679
12	0.9324	0.6551	1.4734	1.5883	1.7600	0.8992	0.8109	0.7179	1.4437	2.3570	1.6537	1.1912	0.8640	1.7921	2.2188	2.2887	22.6463
16	1.4152	0.9052	2.0915	1.3824	1.4929	1.5821	0.9747	1.2475	1.5282	2.7014	2.0292	1.5424	1.3890	1.7366	2.8034	2.6064	27.4281
13	1.4138	0.9035	2.0899	1.3727	1.4792	1.5811	0.9733	1.2465	1.5264	2.6869	2.0189	1.5427	1.3884	1.7244	2.7899	2.5917	27.3295
13-1	1.4139	0.9035	2.0899	1.3727	1.4792	1.5810	0.9733	1.2465	1.5264	2.6869	2.0189	1.5427	1.3884	1.7244	2.7900	2.5918	27.3297
13-2	1.4139	0.9035	2.0899	1.3727	1.4792	1.5810	0.9733	1.2465	1.5264	2.6869	2.0189	1.5427	1.3884	1.7244	2.7900	2.5918	27.3297
14	0.2030	0.5496	0.5469	4.9886	1.3055	0.4888	0.2882	0.2194	0.4401	2.4840	2.0364	0.4709	0.4503	1.2096	2.6775	2.5913	20.9500

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Table C.3 IAE Result of Pressure Deviation at Process Stream

Steam	VLV -100	VLV -101	VLV -102	VLV -103	VLV -104	VLV -105	VLV -106	VLV -107	Flow top1	Flow top2	Qevap	Qcooler	Qcon1	Qcon2	Qreboil1	Qreboil2	SUM IAE
feed (1)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
1-2	0.7700	1.0358	0.4264	0.3800	0.7884	0.6187	1.1428	0.7555	0.6160	0.4089	0.3665	0.9251	0.5742	0.5225	0.2603	0.5499	10.1412
2	0.8243	1.1093	0.4569	0.4071	0.8447	0.6623	1.3343	0.8100	0.6601	0.4388	0.3935	0.9898	0.6160	0.5595	0.2797	0.5895	10.9759
com_cold_st	0.8243	1.1093	0.4569	0.4071	0.8447	0.6623	1.3343	0.8100	0.6601	0.4388	0.3935	0.9898	0.6160	0.5595	0.2797	0.5895	10.9759
4	0.8437	1.1443	0.4677	0.4008	0.8317	0.6638	1.3082	0.8401	0.6742	0.4347	0.3942	1.0216	0.6342	0.5525	0.2767	0.5825	11.0708
4-1	0.8495	1.1534	0.4711	0.4004	0.8337	0.6698	1.3159	1.5245	0.6781	0.4350	0.3918	1.0301	0.6402	0.5521	0.2767	0.5826	11.8050
5	0.8721	1.1861	0.4827	0.3987	0.8420	0.6904	1.3492	1.5620	0.6936	0.4361	0.3821	1.0614	0.6608	0.5506	0.2767	0.5830	12.0274
Re_out	0.9535	1.3072	0.5348	0.4194	0.8751	0.7554	1.3920	1.6153	0.7522	0.4564	0.4047	1.1638	0.7240	0.5801	0.2890	0.6082	12.8311
com_hot_st	1.3416	1.3345	0.5459	0.4218	0.8786	0.7673	1.3935	1.6170	0.7651	0.4594	0.4090	1.1870	0.7397	0.5843	0.2908	0.6116	13.3472
8	1.3735	1.4111	0.5769	0.4273	0.8895	0.8085	1.4064	1.6318	0.8014	0.4677	0.4178	1.2536	0.7828	0.5942	0.2955	0.6206	13.7585
9	0.7517	0.7094	2.2727	.1944	0.3413	0.8320	0.4926	0.7089	1.9013	0.3044	0.2066	0.5202	2.2553	0.3250	0.1306	0.1885	12.1348
10	0.7499	0.7075	2.2730	0.1938	0.3403	0.8316	0.4913	0.7073	1.9139	0.3037	0.2060	0.5191	2.2581	0.3243	0.1303	0.1878	12.1379
11	0.7522	0.7078	2.2772	0.1945	0.3409	0.8378	0.4927	0.7102	1.8976	2.6715	3.8114	2.0219	2.2583	0.3255	5.1376	0.1875	24.6245
11-1	0.7119	0.6766	2.1263	5.1970	0.3283	0.7500	0.4645	0.6647	1.8333	2.5567	3.7071	1.8793	2.0970	0.3011	5.1056	0.0532	28.4526
12	1.9057	1.0549	1.6685	2.8718	2.0280	2.6119	0.7641	1.2752	1.3272	2.7498	2.2900	0.8640	1.0823	3.6935	1.9124	3.7981	31.8974
16	1.9048	1.0545	1.6678	2.8706	2.0269	2.6106	0.7637	1.2746	1.3268	2.7486	2.2890	0.8637	1.0818	3.6914	1.9116	3.7964	31.8828
13	1.0142	1.0245	0.7119	1.1183	3.2469	1.2871	1.1210	0.5956	0.8499	1.0615	0.8594	0.8653	0.6638	1.4686	0.6803	1.4928	18.0613
13-1	0.8243	1.1093	0.4569	0.4071	0.8447	0.6623	1.3343	0.8100	0.6601	0.4388	0.3935	0.9898	0.6160	0.5595	0.2797	0.5895	10.9759
13-2	0.8243	1.1093	0.4569	0.4071	0.8447	0.6623	1.3343	0.8100	0.6601	0.4388	0.3935	0.9898	0.6160	0.5595	0.2797	0.5895	10.9759
14	1.9084	1.0551	1.6695	2.8825	2.0295	2.6159	0.7647	1.2772	1.3289	2.7504	2.2905	0.8647	1.0835	3.6965	1.9072	3.7992	31.9238

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Table C.4 IAE Result of Flowrate Deviation at Process Stream

Steam	VLV -100	VLV -101	VLV -102	VLV -103	VLV -104	VLV -105	VLV -106	VLV -107	Flow top1	Flow top2	Qevap	Qcooler	Qcon1	Qcon2	Qreboil1	Qreboil2	SUM IAE
feed (1)	0.9119	0.9625	0.6678	0.1038	1.3633	0.0778	1.6219	0.8176	0.5889	1.1079	1.1556	1.1148	0.6573	1.3250	1.1775	1.1265	14.7801
1-2	0.9119	0.9625	0.6678	0.1038	1.3633	0.0778	1.6219	0.8176	0.5889	1.1079	1.1556	1.1148	0.6573	1.3250	1.1775	1.1265	14.7801
2	0.9119	0.9625	0.6678	0.1038	1.3633	0.0778	1.6219	0.8176	0.5889	1.1079	1.1556	1.1148	0.6573	1.3250	1.1775	1.1265	14.7801
com_cold_st	0.9149	1.2447	1.0214	0.0300	0.6145	0.0757	1.2846	1.3469	0.4498	0.3713	0.5392	1.2563	0.9101	0.2896	0.4297	0.4260	11.2048
4	0.8944	1.2441	1.0222	0.0312	0.6146	0.0703	1.2799	1.3316	0.4476	0.3700	0.5388	1.2566	0.8960	0.2829	0.4291	0.4246	11.1341
4-1	0.8944	1.2441	1.0222	0.0312	0.6146	0.0703	1.2799	1.3316	0.4476	0.3700	0.5388	1.2566	0.8960	0.2829	0.4291	0.4246	11.1341
5	0.8815	1.2437	1.0228	0.0320	0.6148	0.0679	1.2770	1.3210	0.4465	0.3692	0.5385	1.2567	0.8875	0.2814	0.4288	0.4237	11.0930
Re_out	0.7496	2.3057	1.0330	0.0440	0.6198	0.0898	1.2360	1.2062	0.4646	0.3609	0.5351	1.3100	0.9023	0.4074	0.4239	0.4118	12.1000
com_hot_st	0.9980	2.3413	1.0374	0.0530	0.6283	0.1409	1.2244	1.2636	0.5366	0.3556	0.5332	1.3202	1.0737	0.5938	0.4215	0.4058	12.9272
8	1.0359	1.8961	1.0190	0.0264	0.6189	0.1237	1.3000	1.4277	0.4874	0.3806	0.5419	1.2944	1.0447	0.4327	0.4318	0.4318	12.4931
9	1.0359	1.8961	1.0190	0.0264	0.6189	0.1237	1.3000	1.4277	0.4874	0.3806	0.5419	1.2944	1.0447	0.4327	0.4318	0.4318	12.4931
10	1.1870	0.9236	3.2969	0.0728	0.8148	0.1270	0.9610	1.0300	3.0193	1.1742	0.9194	0.8862	3.4105	1.0782	0.8339	0.5512	20.2860
11	1.6499	0.4335	1.2852	6.0557	0.7149	0.2381	0.6060	0.9750	3.0876	2.0441	1.5398	1.3340	1.4070	1.6576	1.4971	1.9326	26.4580
11-1	1.6499	0.4335	1.2852	6.0557	0.7149	0.2381	0.6060	0.9750	3.0876	2.0441	1.5398	1.3340	1.4070	1.6576	1.4973	1.9326	26.4581
12	1.6499	0.4335	1.2852	6.0557	0.7149	0.2381	0.6060	0.9750	3.0876	2.0441	1.5398	1.3340	1.4070	1.6576	1.4971	1.9326	26.4580
16	0.8241	0.3455	0.6266	0.1303	1.9501	0.1355	0.5131	0.6552	0.4340	1.4435	1.3980	0.2636	0.6015	1.5918	1.5971	1.5518	14.0616
13	0.8241	0.3455	0.6266	0.1303	1.9501	0.1355	0.5131	0.6552	0.4340	1.4435	1.3980	0.2636	0.6015	1.5918	1.5971	1.5518	14.0616
13-1	0.8241	0.3455	0.6266	0.1303	1.9501	0.1355	0.5131	0.6552	0.4340	1.4435	1.3980	0.2636	0.6015	1.5918	1.5971	1.5518	14.0616
13-2	0.8241	0.3455	0.6266	0.1303	1.9501	0.1355	0.5131	0.6552	0.4340	1.4435	1.3980	0.2636	0.6015	1.5918	1.5971	1.5518	14.0616
14	0.4267	0.0907	0.1407	0.6533	0.2058	17.6212	0.1209	0.3152	0.4476	0.6376	1.0952	0.4674	0.3355	0.6036	1.3280	0.6846	25.1739

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Table C.5 IAE Result of Purity and Level Deviation at Process Stream

DME	VLV -100	VLV -101	VLV -102	VLV -103	VLV -104	VLV -105	VLV -106	VLV -107	Flow top1	Flow top2	Qevap	Qcooler	Qcon1	Qcon2	Qreboil1	Qreboil2	SUM IAE
feed (1)	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
5	0.0045	0.0368	0.0946	0.0079	0.0084	0.0031	0.0040	0.0032	0.0071	0.0545	0.1719	0.2173	0.0042	0.0029	0.2070	0.0400	0.8675
Re_out	5.6955	5.7440	5.4008	5.6813	5.6988	5.6968	5.6889	5.6968	5.6516	5.5749	4.8015	4.5879	5.6957	5.6953	4.5332	5.7859	87.6288
10	0.0030	0.0337	0.0425	0.0021	0.0072	0.0028	0.0065	0.0031	0.0427	0.0375	0.0724	0.0664	0.0035	0.0026	0.0800	0.0288	0.4349
16	0.2970	0.1855	0.4621	0.3087	0.2856	0.2973	0.3007	0.2969	0.2986	0.3331	0.9543	1.1284	0.2965	0.2992	1.1799	0.1453	7.0688
14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Water																	
feed (1)	0.1317	0.1314	0.1319	0.1280	0.1313	0.1317	0.1316	0.1317	0.1315	0.1320	0.1202	0.1215	0.1317	0.1315	0.1153	0.1309	2.0640
5	0.0005	0.0032	0.0069	0.0014	0.0034	0.0004	0.0007	0.0004	0.0008	0.0242	0.0434	0.0191	0.0004	0.0014	0.0557	0.0225	0.1845
Re_out	5.5921	5.5827	5.5716	5.4382	5.5724	5.5926	5.5892	5.5923	5.5833	5.5324	4.9820	5.1391	5.5919	5.5833	4.7112	5.5104	87.1648
10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
16	0.2708	0.2618	0.2496	0.2681	0.2656	0.2710	0.2712	0.2709	0.2716	0.1646	0.1235	0.2454	0.2711	0.2716	0.1433	0.1706	3.7907
14	0.0049	0.0209	0.0399	0.1642	0.0273	0.0043	0.0073	0.0047	0.0127	0.1468	0.7309	0.4749	0.0049	0.0123	0.9744	0.1655	2.7959
Methanol																	
feed (1)	3.9047	1.6342	0.9532	2.0208	3.3008	3.9864	3.7413	3.9767	2.9389	1.2977	0.3423	0.3108	3.8593	3.8070	0.2654	1.6530	37.9924
5	0.1175	0.4113	0.6550	0.1152	0.1537	0.0803	0.1036	0.0818	0.1374	0.3645	0.4419	0.5575	0.1113	0.0728	0.4338	0.2415	4.0791
Re_out	1.1138	1.5457	0.5564	0.5902	0.9550	1.1496	1.0927	1.1470	0.8713	0.4696	0.4193	0.9128	1.1536	1.1216	0.2695	0.3406	13.7088
10	0.0881	0.4015	0.3146	0.0310	0.1768	0.0823	0.1791	0.0920	0.9363	0.3681	0.2161	0.1853	0.1004	0.0737	0.1956	0.3463	3.7872
16	0.6299	1.7475	3.2326	0.6515	0.7274	0.5707	0.6769	0.5602	0.8312	2.0566	2.4997	2.8189	0.6319	0.5700	2.5931	1.3293	22.1274
14	0.1460	0.2599	0.2882	2.5912	0.6863	0.1307	0.2064	0.1423	0.2847	1.4436	2.0807	1.2148	0.1434	0.3549	2.2426	2.0893	14.3050
Level																	
LCon1	0.2008	0.1712	0.1783	0.0143	0.2968	0.0590	0.2528	0.2108	0.1289	0.5418	0.1019	0.1100	1.8983	0.1618	0.0999	0.0609	4.4874
Lcon2	2.6294	0.3860	0.1619	0.1481	1.7701	0.7573	1.9946	2.8136	1.1099	0.9574	0.0865	0.1205	1.3277	2.5394	0.0786	0.1706	17.0515
LReboil1	1.0485	3.1919	3.4013	2.8050	1.4701	0.2566	1.5104	0.8160	2.4369	0.0007	2.4208	2.7141	0.6836	0.9072	2.2457	3.3666	29.2753
LReboil2	0.1212	0.2508	0.2585	1.0327	0.4630	2.9272	0.2423	0.1596	0.3243	2.5001	1.3909	1.0554	0.0904	0.3916	1.5758	0.4020	13.1858

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Table C.6 IAE Result of Stage Temperature Deviation at Product Column

Tdis1	VLV -100	VLV -101	VLV -102	VLV -103	VLV -104	VLV -105	VLV -106	VLV -107	Flow top1	Flow top2	Qevap	Qcooler	Qcon1	Qcon2	Qreboil1	Qreboil2	SUM IAE
state 1	0.5136	0.7118	0.4578	0.4643	0.7236	0.5467	0.9270	0.6207	2.5207	1.0186	0.6947	0.4083	0.4813	0.5474	0.7006	0.6737	12.0106
state 2	1.1868	1.9138	1.4941	0.9706	1.5304	1.3326	1.9153	1.4280	5.3928	2.4103	1.8617	1.1922	1.2776	1.3589	2.1273	1.4645	28.8569
state 3	0.8397	1.5384	1.2061	0.6712	1.0469	0.9559	1.3018	1.0105	3.2230	1.9294	1.7707	1.3506	0.8765	0.9810	2.8386	1.0252	22.5653
state 4	0.4612	0.7773	0.4994	0.3673	0.5547	0.5303	0.6931	0.5551	1.0738	0.9399	0.8828	0.8597	0.4833	0.5474	2.2038	0.5249	11.9541
state 5	0.3343	0.4877	0.2598	0.2676	0.3922	0.3863	0.4940	0.4021	0.3597	0.5544	0.4522	0.4344	0.4071	0.4003	1.1854	0.3553	7.1727
state 6	0.3038	0.4122	0.1970	0.2453	0.3543	0.3506	0.4481	0.3649	0.1906	0.4634	0.3303	0.2550	0.3835	0.3635	0.5890	0.3152	5.5667
state 7	0.3020	0.4005	0.1818	0.2470	0.3536	0.3473	0.4477	0.3619	0.1821	0.4469	0.3074	0.2027	0.3718	0.3597	0.3975	0.3139	5.2237
state 8	0.3085	0.4080	0.1690	0.2564	0.3659	0.3513	0.4629	0.3680	0.2298	0.4529	0.3154	0.1909	0.3588	0.3629	0.3607	0.3269	5.2884
state 9	0.3164	0.4225	0.1632	0.2679	0.3831	0.3543	0.4835	0.3744	0.2953	0.4702	0.3356	0.1905	0.3451	0.3642	0.3711	0.3470	5.4843
state 10	0.3156	0.4244	0.1637	0.2712	0.3924	0.3460	0.4925	0.3694	0.3077	0.4892	0.3537	0.1944	0.3355	0.3532	0.3863	0.3632	5.5585
state 11	0.2887	0.3833	0.1662	0.2407	0.3662	0.3072	0.4542	0.3332	0.1125	0.4699	0.3385	0.1820	0.3327	0.3111	0.3610	0.3473	4.9948
state 12	0.2906	0.3831	0.1683	0.2419	0.3676	0.3123	0.4554	0.3360	0.1132	0.4704	0.3403	0.1867	0.3369	0.3166	0.3603	0.3481	5.0276
state 13	0.2941	0.3842	0.1697	0.2438	0.3706	0.3159	0.4575	0.3391	0.1146	0.4724	0.3472	0.2206	0.3398	0.3201	0.3587	0.3505	5.0989
state 14	0.3063	0.3900	0.1689	0.2533	0.3808	0.3277	0.4654	0.3498	0.1192	0.4796	0.3716	0.3310	0.3476	0.3319	0.3547	0.3592	5.3370
state 15	0.3478	0.4101	0.1669	0.2917	0.4147	0.3701	0.4922	0.3872	0.1350	0.5008	0.4603	0.6123	0.3734	0.3745	0.3584	0.3885	6.0838
state 16	0.4770	0.4759	0.2636	0.4228	0.5207	0.4990	0.5755	0.5045	0.1863	0.5628	0.7355	1.1490	0.4564	0.5036	0.4470	0.4817	8.2612
state 17	0.8631	0.6990	0.6567	0.8402	0.8365	0.8850	0.8223	0.8637	0.3598	0.7419	1.3617	1.9107	0.7928	0.8902	0.7414	0.7650	14.0300
state 18	1.8357	1.3415	1.6769	1.9097	1.6384	1.8255	1.4671	1.7647	0.8282	1.1925	2.3168	2.7269	1.7597	1.8269	1.3780	1.5153	27.0039
state 19	3.4401	2.5245	3.3982	3.6864	2.9950	3.3357	2.5553	3.2185	1.6273	2.0085	3.0710	3.2511	3.3832	3.3202	2.1544	2.8942	46.8635
state 20	4.2597	3.3661	4.5859	4.6223	3.7241	4.0439	3.0912	3.9041	2.0925	2.4968	2.8638	3.0543	4.1859	4.0008	2.3079	3.8891	56.4882
state 21	3.1603	2.7434	3.7913	3.4673	2.8218	2.9222	2.3050	2.8270	1.6340	1.9197	1.7423	2.1068	3.0199	2.8654	1.5189	3.2193	42.0648
state 22	1.5547	1.4021	1.9954	1.7511	1.4666	1.3543	1.1933	1.3173	0.9020	1.5096	0.7466	0.9900	1.3510	1.3001	0.4990	1.7320	21.0650

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Table C.7 IAE Result of Stage Temperature Deviation at Recycle Column

Tdis2	VLV -100	VLV -101	VLV -102	VLV -103	VLV -104	VLV -105	VLV -106	VLV -107	Flow top1	Flow top2	Qevap	Qcooler	Qcon1	Qcon2	Qreboil1	Qreboil2	SUM IAE
state 1	1.0081	0.9283	0.9283	0.2680	1.0151	0.9082	0.9622	0.8078	0.4835	0.8921	0.9248	0.6067	0.6010	1.0037	0.9288	0.8821	13.1488
state 2	0.8600	0.9172	0.9172	0.2801	1.0474	0.7522	0.8753	0.8422	0.5027	0.9030	0.9356	0.6283	0.6277	0.9922	0.9377	0.8945	12.9133
state 3	0.8558	0.9301	0.9301	0.2845	1.0486	0.7474	0.8688	0.8392	0.5233	0.9147	0.9473	0.6566	0.6273	0.9779	0.9472	0.9080	13.0067
state 4	0.8440	0.9408	0.9408	0.2883	1.0453	0.7316	0.8701	0.8356	0.5535	0.9277	0.9600	0.6942	0.6239	0.9669	0.9572	0.9228	13.1027
state 5	0.8623	0.9537	0.9537	0.2921	1.0424	0.7485	0.8759	0.8110	0.5937	0.9415	0.9733	0.7440	0.6074	0.9584	0.9675	0.9387	13.2642
state 6	0.8961	0.9698	0.9698	0.2964	1.0406	0.7821	0.8848	0.7884	0.6497	0.9556	0.9860	0.8088	0.5993	0.9570	0.9769	0.9546	13.5159
state 7	0.8972	0.9894	0.9894	0.3015	1.0388	0.7769	0.8902	0.7320	0.7176	0.9677	0.9954	0.8821	0.5773	0.9639	0.9831	0.9683	13.6706
state 8	0.8958	1.0142	1.0142	0.3075	1.0390	0.7716	0.9002	0.6585	0.7874	0.9738	0.9987	0.9478	0.5500	0.9783	0.9839	0.9738	13.7948
state 9	0.8754	1.0458	1.0458	0.3143	1.0416	0.7457	0.9103	0.5978	0.8313	0.9733	0.9901	0.9773	0.5298	1.0002	0.9792	0.9596	13.8173
state 10	0.8567	1.0848	1.0848	0.3213	1.0480	0.7202	0.9248	0.5335	0.8159	0.9326	0.9542	0.9450	0.5415	1.0458	0.9571	0.9070	13.6732
state 11	0.8092	1.1285	1.1285	0.3272	1.0580	0.6647	0.9394	0.5137	0.7037	0.9082	0.9290	0.8071	0.5860	1.0288	0.9336	0.8903	13.3557
state 12	0.7534	1.1700	1.1700	0.3283	1.0701	0.5878	0.9500	0.5106	0.6781	0.9039	0.9245	0.7867	0.5939	1.0213	0.9300	0.8879	13.2665
state 13	0.7014	1.1759	1.1759	0.3201	1.0814	0.5222	0.9507	0.5010	0.6659	0.9033	0.9236	0.7773	0.6013	1.0116	0.9295	0.8880	13.1293
state 14	0.6418	1.1422	1.1422	0.3011	1.0948	0.4717	0.9354	0.4983	0.6555	0.9047	0.9239	0.7698	0.6126	0.9961	0.9303	0.8900	12.9104
state 15	0.5678	1.0791	1.0791	0.3033	1.1053	0.4853	0.8481	0.5004	0.6420	0.9097	0.9258	0.7807	0.6445	0.9639	0.9329	0.8955	12.6633
state 16	0.5563	1.0659	1.0659	0.2995	1.1023	0.4895	0.8331	0.4911	0.6500	0.9233	0.9311	0.8294	0.7385	0.8920	0.9397	0.9097	12.7174
state 17	0.5452	1.0568	1.0568	0.2869	1.0933	0.4892	0.8272	0.5077	0.8589	0.9584	0.9447	0.9544	0.9682	0.8501	0.9572	0.9460	13.3010
state 18	0.5376	1.0463	1.0463	0.2539	1.0770	0.4987	0.8197	0.9537	1.3784	1.0418	0.9778	1.2266	1.4886	0.7902	0.9993	1.0348	15.1708
state 19	0.5275	1.0298	1.0298	0.1931	1.0428	0.5274	0.7999	2.2836	2.4959	1.2146	1.0563	1.7645	2.5184	0.8635	1.0918	1.2281	19.6670
state 20	0.4970	0.9979	0.9979	0.3014	0.9627	0.5968	0.7460	3.6918	3.7352	1.4586	1.2196	2.4391	3.7461	1.7001	1.2503	1.5153	25.8558
state 21	0.4935	0.9311	0.9311	0.6679	0.7730	0.7938	0.6554	3.3189	3.3616	1.5397	1.4078	2.5444	3.3607	1.5001	1.3880	1.6262	25.2933
state 22	0.7675	0.8112	0.8112	1.8667	0.3539	1.3071	0.6822	1.4586	1.6532	1.3144	1.4027	1.8431	1.7760	0.5975	1.3279	1.3716	19.3445
state 23	2.0431	0.7752	0.7752	3.9362	0.6505	2.3809	1.4801	0.8078	0.4835	0.8921	0.9248	0.6067	0.6010	1.0037	0.9288	0.8821	19.1719
state 24	3.3890	1.0839	1.0839	5.7046	1.5107	3.5581	2.4307	0.8422	0.5027	0.9030	0.9356	0.6283	0.6277	0.9922	0.9377	0.8945	26.0248
state 25	3.0461	1.0614	1.0614	5.1499	1.3257	3.2366	2.1920	0.8392	0.5233	0.9147	0.9473	0.6566	0.6273	0.9779	0.9472	0.9080	24.4144
state 26	1.2721	0.6706	0.6706	2.8058	0.2918	1.7061	0.9476	0.8356	0.5535	0.9277	0.9600	0.6942	0.6239	0.9669	0.9572	0.9228	15.8064

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Table C.8 IAE Result of Stage Temperature Deviation at Product Column

Stream	process variable	VLV -100	VLV -101	VLV -102	VLV -103	VLV -104	VLV -105	VLV -106	VLV -107	Flow top1	Flow top2	Qevap	Qcooler	Qcon1	Qcon2	Qreboil1	Qreboil2	SUM IAE
16	Pressure	1.9048	1.0545	1.6678	2.8706	2.0269	2.6106	0.7637	1.2746	1.3268	2.7486	2.2890	0.8637	1.0818	3.6914	1.9116	3.7964	31.8828
LReboil1	Level	1.0485	3.1919	3.4013	2.8050	1.4701	0.2566	1.5104	0.8160	2.4369	0.0007	2.4208	2.7141	0.6836	0.9072	2.2457	3.3666	29.2753
12	Flow rate	1.6499	0.4335	1.2852	6.0557	0.7149	0.2381	0.6060	0.9750	3.0876	2.0441	1.5398	1.3340	1.4070	1.6576	1.4971	1.9326	26.4580
com_hot_st	Temp	3.2518	1.3448	0.6178	1.5558	1.3106	3.4258	1.6683	2.8274	1.8347	0.3928	0.9252	1.0460	2.8576	2.2630	0.2698	0.4017	25.9930
4	Temp	1.2319	3.2685	1.0927	0.9942	1.2173	0.6352	2.7139	2.0313	1.1345	0.2603	1.4747	2.5174	1.3405	1.1989	0.1827	0.2944	21.5883
14	Temp	0.2030	0.5496	0.5469	4.9886	1.3055	0.4888	0.2882	0.2194	0.4401	2.4840	2.0364	0.4709	0.4503	1.2096	2.6775	2.5913	20.9500
10	Flow rate	1.1870	0.9236	3.2969	0.0728	0.8148	0.1270	0.9610	1.0300	3.0193	1.1742	0.9194	0.8862	3.4105	1.0782	0.8339	0.5512	20.2860
Re_out	Temp	2.4551	0.3141	0.2172	1.3978	1.0369	3.3287	1.2431	2.5568	1.6760	0.0892	0.0344	0.0606	2.6647	2.0652	0.0264	0.1169	19.2832
5	Temp	0.9376	2.4860	1.1987	0.4502	0.8946	0.2560	2.4630	1.1000	0.6437	0.6495	1.9609	2.1894	0.8188	0.5578	0.5482	0.5275	17.6819
Lcon2	Level	2.6294	0.3860	0.1619	0.1481	1.7701	0.7573	1.9946	2.8136	1.1099	0.9574	0.0865	0.1205	1.3277	2.5394	0.0786	0.1706	17.0515
8	Temp	1.2176	1.5786	0.8871	0.4601	1.2072	1.4586	1.2800	1.2375	0.7353	0.5287	0.4690	0.7506	1.3141	0.8323	0.4795	0.6102	15.0465
2	Flow rate	0.9119	0.9625	0.6678	0.1038	1.3633	0.0778	1.6219	0.8176	0.5889	1.1079	1.1556	1.1148	0.6573	1.3250	1.1775	1.1265	14.7801
13-1	Flow rate	0.8241	0.3455	0.6266	0.1303	1.9501	0.1355	0.5131	0.6552	0.4340	1.4435	1.3980	0.2636	0.6015	1.5918	1.5971	1.5518	14.0616
LReboil2	Level	0.1212	0.2508	0.2585	1.0327	0.4630	2.9272	0.2423	0.1596	0.3243	2.5001	1.3909	1.0554	0.0904	0.3916	1.5758	0.4020	13.1858
9	Flow rate	1.0359	1.8961	1.0190	0.0264	0.6189	0.1237	1.3000	1.4277	0.4874	0.3806	0.5419	1.2944	1.0447	0.4327	0.4318	0.4318	12.4931
10	Pressure	0.7499	0.7075	2.2730	0.1938	0.3403	0.8316	0.4913	0.7073	1.9139	0.3037	0.2060	0.5191	2.2581	0.3243	0.1303	0.1878	12.1379
Re_out	Flow rate	0.7496	2.3057	1.0330	0.0440	0.6198	0.0898	1.2360	1.2062	0.4646	0.3609	0.5351	1.3100	0.9023	0.4074	0.4239	0.4118	12.1000
11	Temp	0.9136	0.8296	1.4044	0.5695	0.8736	0.6633	0.6043	0.5549	1.6096	0.3895	0.4527	1.0730	0.5565	0.3755	0.5323	0.5416	11.9440
4-1	Flow rate	0.8944	1.2441	1.0222	0.0312	0.6146	0.0703	1.2799	1.3316	0.4476	0.3700	0.5388	1.2566	0.8960	0.2829	0.4291	0.4246	11.1341
4	Pressure	0.8437	1.1443	0.4677	0.4008	0.8317	0.6638	1.3082	0.8401	0.6742	0.4347	0.3942	1.0216	0.6342	0.5525	0.2767	0.5825	11.0708
Lcon1	Level	0.2008	0.1712	0.1783	0.0143	0.2968	0.0590	0.2528	0.2108	0.1289	0.5418	0.1019	0.1100	1.8983	0.1618	0.0999	0.0609	4.4874
14	Water Com	0.0049	0.0209	0.0399	0.1642	0.0273	0.0043	0.0073	0.0047	0.0127	0.1468	0.7309	0.4749	0.0049	0.0123	0.9744	0.1655	2.7959
10	DME Com	0.0030	0.0337	0.0425	0.0021	0.0072	0.0028	0.0065	0.0031	0.0427	0.0375	0.0724	0.0664	0.0035	0.0026	0.0800	0.0288	0.4349

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Table C.9 Control Variable and Manipulate Variable are Selected for Control Structure 3 and Control Structure 4

stream	Type of Variable	Control Variable	Manipulate Variable for CS3	Manipulate Variable for CS4
16	Pressure	Recycle column pressure	Qcon2	Qcon2
Lreboil1	Level	Product column level reboil	Cascade with VLV-103	VLV-101
12	Flow rate	Recycle column feed flow rate	VLV-103	VLV-103
4	Temperature	Vaporizer outlet temperature	Q26	Q26
14	Temperature	Recycle column temperature	Qreboiler 2	VLV-105
5	Temperature	Reactor inlet temperature	VLV-100	VLV-100
Lcon2	Level	Condenser level of recycle column	Reflux flow dis2	Reflux flow dis2
8	Temperature	Product column inlet temperature	Q27	Q27
2	Flow rate	Inlet flow rate stream	VLV-106	VLV-106
13-1	Flow rate	Recycle flow rate	VLV-104	VLV-104
Lreboil2	Level	Recycle column level reboil	VLV-105	Qreboil2
9	Flow rate	Product column feed flow rate	VLV-101	-
10	Pressure	Product column pressure	Qcon1	Qcon1
11	Temperature P	roduct column temperature	Qreboil1	Qreboil1
4-1	Flow rate	FEHE feed flow rate	VLV-107	VLV-107
4	Pressure	Vaporizer outlet pressure	Cascade with flow in, VLV-106	Cascade with flow in, VLV-106
LCon1	Level	Condenser level of product column	VLV-102	VLV-102
10	DME Com	Product purity	Reflux flow dis1	Reflux flow dis1

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VITA

Mr. Lersak Bunnachai was born in Nakhonsrithammarat, Thailand on October 13, 1982. He graduated Bachelor Degree in Chemical Engineering from King's Mongkut Institute of Technology North in 2005. Then he had been worked at Michelin Siam Company until 2007, after that he studied for Master degree in Chemical Engineering and joined Control and Systems Engineering research group at Chulalongkorn University and completed in 2009.