การออกแบบโครงสร้างการควบคุมกระบวนการแยกน้ำของไอโซโพรพิลแอลกอฮอล์

นายศิวภัทร โทจันทร์

วิทยานิพนธ์นี้เป็นส่วนหนึ่งของการศึกษาตามหลักสูตรปริญญาวิศวกรรมศาสตรมหาบัณฑิต
สาขาวิชาวิศวกรรมเคมี ภาควิชาวิศวกรรมเคมี
คณะวิศวกรรมศาสตร์ จุฬาลงกรณ์มหาวิทยาลัย
ปีการศึกษา 2555

ลิขสิทธิ์ของจุฬาลงกรณ์มหาวิทยาลัย
บทคัดย่อและแฟ้มข้อมูลฉบับเต็มของวิทยานิพนธ์ตั้งแต่ปีการศึกษา 2554 ที่ให้บริการในคลังปัญญาจุฬาฯ (CUIR)
เป็นแฟ้มข้อมูลของนิสิตเจ้าของวิทยานิพนธ์ที่ส่งผ่านทางบัณฑิตวิทยาลัย
The abstract and full text of theses from the academic year 2011 in Chulalongkorn University Intellectual Repository(CUIR)
are the thesis authors' files submitted through the Graduate School.

CONTROL STRUCTURE DESIGN OF ISOPROPYL ALCOHOL DEHYDRATION PROCESS

Mr. Siwapat Tochan

A Thesis Submitted in Partial Fulfillment of the Requirements

for the Degree of Master of Engineering Program in Chemical Engineering

Department of Chemical Engineering

Faculty of Engineering

Chulalongkorn University

Academic Year 2012

Copyright of Chulalongkorn University

Thesis Title	CONTROL STRUCTURE DESIGN OF ISOPROPYL ALGOHOL DEHYDRATION PROCESS
By	Mr. Siwapat Tochan
Field of study	Chemical Engineering
Thesis Advisor	Assistant Professor Montree Wongsri, D.Sc.

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

.....Dean of the Facuty of Engineering

(Associate Professor Boonsom Lerdhirunwong, Dr.Ing.)

THESIS COMMITTEE

.....Chairman

(Associate Professor Muenduen Phisalaphong, D.Eng)

.....Thesis Advisor

(Assistant Professor Montree Wongsri, D.Sc.)

.....Examiner

(Assistant Professor Amornchai Armornwichanop, D.Eng)

.....External Examiner

(Chaiyapop Siraworakun,D.Eng)

ศิวภัทร โทจันทร์ : การออกแบบโครงสร้างการควบคุมกระบวนการแยกน้ำของไอโซโพรพิลแอลกอฮอล์. (CONTROL STRUCTURE DESIGN OF ISOPROPYL ALCOHOL DEHYDRATION PROCESS) อ.ที่ปรึกษาวิทยานิพนธ์หลัก : ผศ.ดร. มนตรี วงศ์ศรี, 92 หน้า.

วิทยานิพนธ์นี้เสนอการออกแบบโครงสร้างการควบคุมทั้งโรงงานสำหรับกระบวนการแยกน้ำของไอโซโพรพิลแอลกอฮอล์โดยใช้วิธีการออกแบบของวงศ์ศรี ซึ่งวิธีการออกแบบประกอบด้วย 8 ขั้นตอน โดยเน้นการออกแบบในระดับทั้งโรงงานซึ่งทำให้เกิดโรงงานคงที่ การจัดการกับสิ่งรบกวนเพื่อการควบคุมคุณภาพ โดยกระบวนการประกอบด้วยหอกลั่นสามหอและใช้หอกลั่นแบบสกัดแยกโดยอาศัยสารหนัก สายป้อนและสายรีไซเคิลของเอทิลีนไกลคอลถูกควบคุมและเป็นอัตราส่วนกับสายไอโซโพลพิลแอลกอฮอล์ สารแต่ละตัวถูกควบคุมตามตำแหน่งของตัวบ่งบอกปริมาณของสารนั้นๆ ตัวแปรรบกวนที่เป็นมวลสารถูกนำไปยังเส้นทางของสารที่ต้องการเพื่อการควบคุมคุณภาพ โครงสร้างการควบคุมกระบวนการทั้งโรงงานที่ออกแบบได้มีสมรรถนะเทียบได้กับการออกแบบของ Luyben

ภาควิชา วิศวกรรมเคมี ลายมือชื่อนิสิต .....

สาขาวิชา วิศวกรรมเคมี ลายมือชื่อ อ.ที่ปรึกษาวิทยานิพนธ์หลัก .....

ปีการศึกษา 2555

5470395121 : MAJOR CHEMICAL ENGINEERING

KEYWORDS : PLANTWIDE CONTROL / ISOPRORYL ALCOHOL / FIXTURE PLANT

SIWAPAT TOCHAN : CONTROL STRUCTURE DESIGN OF ISOPROPYL ALCOHOL DEHYDRATION PROCESS. ADVISOR : ASST. PROF. MONTERR WONGSRI, D.Sc., 92 pp.

The design of plantwide control structure for isopropyl alcohol dehydration process using Wongsri's design procedure is presented. The 8-step design procedure emphasis on design at plantwide level which are establishing a fixture plant, disturbance management for quality control. The process features three distillation columns and uses an extractive distillation with a heavy entrainer. The combined feed of fresh and recycled ethylene glycol is regulated and is ratioed to IPA feed. Each component is handled at their quantifiers. The material disturbances are directed to the desired material pathways for quality control. The designed plantwide control structure performance is comparable to the Luyben's design.

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

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

Academic Year : 2012 .....

ACKNOWLEDGEMENTS

Foremost, I would like to express my most sincere thank to my advisor, Assistant Professor Montree Wongsri, for this guidance and support.

Many thank for all of the members in the process control laboratory and friends who have contributed to the accomplishment of the work. I would like to express my gratitude to all those who encouraged over the years of my study at Chulalongkorn University.

Finally, I would like to express the highest gratitude to all of family for their love, inspiration, encouragement and financial support throughout this study.

CONTENTS

page

page

LIST OF TABLES

page

page

LIST OF FIGURES

page

page

page

CHAPTER I

INTRODUCTION

This chapter consists of importance and reasons for research, objectives of research, scopes of research, contributions of research, procedure plan and research contents.

1.1 Importance and Reasons for Research

Plantwide process control involves the systems and strategies required to control an entire chemical plant consisting of many interconnected unit operations

One of most common, importance, and challenging control tasks confronting chemical engineers are: How to design the control loops and system needed to run processes? In general, most industrial processes contain a complex flowsheet with several recycle streams, energy integration and any different unit operations. Given a complex, integrated process and a diverse assortment of equipment, we must devise the necessary logic, instrumentation and strategies to operate the plant safety and achieve its design objectives.

Plantwide process control design procedure is classified in to two main schools, heuristics and optimization. Each methodology has its own advantages and drawbacks. For example, rigorous optimization methodologies can be computationally expensive and subject to model accuracies, while heuristics-based methodologies normally require experience and insight.

In this work present a general heuristic design procedure (Wongsri's plantwide control design procedure). The eight steps of the design procedure center around the fundamental principles of plantwide control. Application of the procedure is illustrated with isopropyl alcohol dehydration process.

Solvents are widely used in many industries. The need to recover the solvent requires the subsequent separation of the solvent from components that have been produced during reactions. One importance example is an organic solvent that must be separate from water. The separation is frequently made difficult by the occurrence of complex vapor-liquid equilibrium that generates azeotropes.

The commercial process simulation program Aspen HYSYS is used in the work. The UNIQUAC physical property package is used

1.2 Research Objectives

This research aims to design the plantwide process control structures for an isopropyl alcohol dehydration process by using Wongsri's plantwide control design procedure (Wongsri 2012).

1.3 Scopes of Research

The scopes of this research are as follows:

1. The detail of isopropyl alcohol dehydration is given by Sommer and Melin (2004).

2. The simulations of the isopropyl alcohol dehydration in steady state and dynamic behavior are achieved via the commercial process simulator HYSYS.

3. Wongsri's plantwide control design procedure (Wongsri 2012) is considered for obtaining the new control structures of isopropyl alcohol dehydration.

1.4 Contribution of Research

The contribution of this research is the effectively control structure of an isopropyl alcohol dehydration by using Wongsri's plantwide control design procedure (Wongsri 2012).

1.5 Research Procedures

The research procedures are following steps:

1. Study of the plantwide process control structure design methodology, the isopropyl alcohol dehydration process and the relevant information.

2. Research the Luyben's control structure design of an isopropyl alcohol dehydration process.

3. Simulate the process in steady state and dynamic behavior by commercial process simulator HYSYS.

4. Design new control structures using Wongsri's plantwide control design procedure (Wongsri 2012).

5. Simulate the new control structure in dynamic behavior and compare with the base case, Luyben's control structure (Luyben 2006).

6. Analyze the result of the new design control structure simulation.
7. Summarize the research studied

1.6 Research Contents

This thesis is divided into six parts as follows:

Chapter I: An introduction of research consisting of reasons, objectives, scopes, contributions and procedure of the research.

Chapter II: Review of the earlier researches of plantwide control, control structure design, plantwide control procedure and related researches.

Chapter III: Background information of Luyben's plantwide control theory and Wongsri's plantwide control design procedure (Wongsri 2012).

Chapter IV: Description of the isopropyl alcohol dehydration process via the commercial process simulator.

Chapter V: Description of the designed control structures, dynamic simulation results and comparison of the control structures with Luyben's control structure (Luyben 2006).

Chapter VI: Conclusion of this research and Recommendations.

1.7 Research Plan

1. Study of plantwide process control theory, the isopropyl alcohol dehydration process.
2. Steady state modeling and simulation of isopropyl alcohol dehydration process.
3. Study the Wongsri's plantwide control design procedure
4. Design control structure for isopropyl alcohol dehydration process.
5. Dynamic simulation for isopropyl alcohol dehydration process with control structure design
6. Assessment of the dynamic performance of the control structure.
7. Analysis of the design and simulation.
8. Conclusion of the thesis.

CHAPTER II

LITERATURE REVIEW

This chapter presented the literature reviews of this process which illustrated the previous work on plantwide control procedure and control structure design.

2.1 Plantwide control

Luyben (1996) presented the number of parameters or variables that must be specified to complete the defined steady-state process, called design degree of freedom (DOF). DOF can be calculated by subtracting the number of equations from the number of the number of variables. For complex process DOF equal to the number of manipulated variables (the number of control valves in the process). The complexity of the phase equilibrium and the physical properties does not affect DOF.

Luyben, Tyreus and Luyben (1997) presented a general heuristic design procedure. 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. This procedure was illustrated with three industrial examples: the vinyl acetate monomer process, Eastman process and HDA process.

Skogestad (2000) presented the method is related to finding a simple and robust way of implementing the economically optimal operating policy. The goal is to find a set of controlled variables which, when kept at constant setpoints, indirectly lead to near-optimal operation with acceptable loss. Since the economics are determined by the overall plant behavior, it is necessary to take a plantwide perspective. A systematic procedure for finding suitable controlled variables based on only steady-state information is presented. Important steps are degree of freedom analysis, definition of optimal operation (cost and constraints), and evaluation of the loss when the controlled variables are kept constant rather than optimally adjusted. A case study yields very interesting insights into the control and maximum throughput of distillation columns. The focus in this paper has not been on finding the optimal operation policy, but rather on how to implement it in a simple manner in the control system. The idea is to find a set of controlled

variables c which, when kept at constant setpoints, indirectly lead to near-optimal operation (with acceptable loss). This is denoted “self-optimizing” control.

Skogestad (2004) proposed two main systematic procedures for control structure design of complete chemical plants (plantwide control); top-down analysis and bottom-up design. Top-down analysis is used to determine definition of operational objectives, manipulated variables and degrees of freedom, primary controlled variables and production rate. While, bottom-up design is used to identify regulatory control layer, supervisory control layer, optimization layer and validation. The studied also presented inventory and production rate control, decentralized versus multivariable control, loss in performance by bottom-up design and a definition of a “complexity number”

Suntisrikomol (2008) used “Fixture Point Theorem” to develop the Hydrodealkylation process (HDA) by Select the suitable set of controlled variable. The theorem states that the most disturbed points must be controlled before other controlled variables. The manipulate variables were selected and paired controlled variables by maximum gain. The performance of design process was illustrated in the IAE values. The Result compared with 2 reference structures (Araújo *et al*., 2006, Luyben, 1998) which the design control structure via Fixture Point Theorem give responded faster and more effective.

Detjareansri (2009) used Wongsri’s plantwide control procedure (Wongsri 2009) for designed control structure and developed the alkylation process. The dynamic performance of the design control structures are evaluated and compared with Luyben (2002) by inserted two types of disturbances; material and thermal disturbances. The designed control structures has good performance and economic.

2.2 Control structure design

Konda, Rangaiah and Krishnaswamy (2005) presented a simple effective procedure to find control degree of freedom (CDOF). The key idea is to define ‘restraining number’ (i.e., the minimum number of flows that cannot be manipulated along with others in a unit, which is also an inherent characteristic of that unit) of a unit. The study show that the restraining number is equal to the number of independent and overall material balances with no associated inventory in

that particular unit. The concept of restraining number is then used to find CDOF of not only simple units but also highly integrated processes.

Dorneau, Bildea and Grievink (2007) proposed a new approach exploiting advantage of fundamental structure that fit in a chemical plant in the form of units or groups of units connected together via material and energy streams. The recommended procedure is to employ model reduction, then to link these reduced-order models. The procedure is flexible and accurate due to its algorithm and variation from one unit to another. The time for solving solution is drastically reduced. The performance of the approach is verified by means of a case study.

James J. Downs, Skogestad (2011) presented the concept of process control design based on a holistic, the variety of procedures and approaches to the design problem has illustrated the difficulty of a unified approach. Using examples, the need and advantages of using a systematic approach based on considering the plant economics are highlighted. The examples deal with disturbance rejection, throughput maximization and economic optimization of plants consisting of parallel units.

2.3 Isopropyl Alcohol Dehydration Process

Sommer and Melin (2004) studied the steady-state economics of several alternative processes for the dehydration of liquid organic solvents. They explored both conventional distillation processes and hybrid processes that combine distillation with membranes. Their objective is to use the strengths of membrane separation to complement the weaknesses of distillation

Luyben (2006) presented plantwide control of a isopropyl alcohol dehydration process by used parameters and condition from Sommer and Melin's work. He proposes temperature control in the extractive column and the solvent recovery column by ratio the flow rate of the extractive solvent to the feed flow rate.

CHAPTER III

PLANWIDE CONTROL PRINCIPLE

In favor of designing the control system methodology, the unit-based system is generally used to design the entire plant. The highly integrated processes (material and energy) are tightly strict to the environmental regulations, safety and economic conditions. Wherefore, the plantwide process control strategies are used to obtain satisfactory products, process performance and stability.

3.1 Plantwide Control Issues

Most continuous processing plants contain many units, such as reactor, heat exchangers, and distillation columns. The goal of process design is to minimize capital costs while operating with optimum utilization of materials and energy. Unfortunately, achieving lower plant capital costs and higher processing efficiencies inevitably makes the individual units interact more with each other and thus makes tem harder to control. The process control engineer deals with these unit-to-unit interactions by designing a control system that counteracts disturbances before they propagate from their source to other units. Therefore, there are general problems a control system is called on to satisfy.

3.1.1 Integrated Process

Three fundamental characteristics of integrated chemical processes are necessary to be considered for control system of the entire plant:

1. The effect of material recycles,
2. The effect of energy integration,
3. The need to account for chemical component inventories.

These issues are concerned if we have to deal with a complex plantwide control.

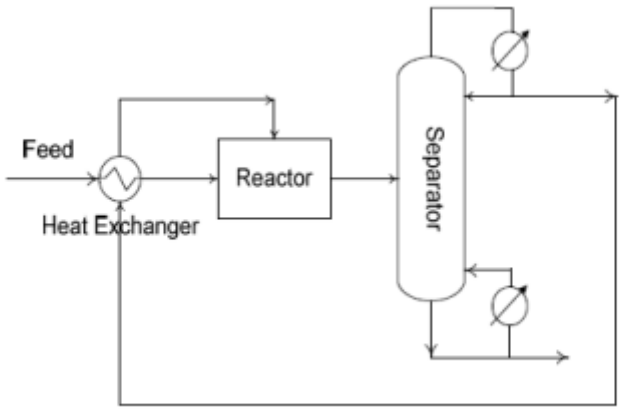


Figure 3.1 Integrated Process Flowsheet

3.1.1.1 Material Recycle

The effects of material recycle are important for six fundamental issues.

1. *Increase conversion*

For chemical processes involving reversible reactions, conversion of reactants to products is bounded by thermodynamics equilibrium constraints. Consequently the reactor effluent by the essential contains both reactants and products. To obtain economical viable, separation and recycle of reactants are necessary.

2. *Improve economics*

In most systems it is simply cheaper to build a reactor with incomplete conversion and recycle reactants than to reach the necessary conversion in one reactor or several in series. A reactor followed by a stripping column with recycle streams is much reasonable in price than using one large reactor or three reactors in series.

3. *Improve yields*

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

4. *Provide thermal sink*

In adiabatic reactors or reactors where cooling is difficult and exothermic heat effects are large, it is often essential to feed excess material to the reactor to prevent large amount of temperature increase in the reactor. High temperature can cause several unpleasant events: it can

lead to thermal runaways, it can deactivate catalysts, it can create undesirable side reactions, it can cause mechanical failure of equipment, etc. Therefore the heat of reaction is absorbed by the sensible heat required to raise the temperature of the excess material in the stream flowing through the reactor.

5. Prevent side reactions

A large excess of one the reactant is often used in order to keep the concentration of the other reactant low. If a limiting reactant is not kept in low concentration, it could react to produce undesirable products. Therefore the excess reactant must be separated from the product components in the reactor effluent stream and recycled back to the reactor.

6. Control properties

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

3.1.1.2 Energy Integration

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

3.1.2 Chemical Component Inventories

The Chemical species in plants can be characterized into three types: reactants, products and inerts. A material balance for each of these components must be satisfied. The actual problem typically appears when reactants are considered (because of recycle) and accounted for their inventories within the entire process. Because of their value, it is necessary to minimize the loss of reactants exiting the process since this represents a yield penalty. So reactants are prevented from leaving. This means we must ensure that every mole of reactant fed to the process is consumed by the reactions.

This is an important, from the viewpoint of individual unit, chemical component balancing is not a problem because exit streams from the unit automatically adjust their flows and composition. However, when we connect units together with recycle streams, the entire system behaves almost like a pure integrator in terms of reactants. If additional reactant is fed into the system without changing reactor conditions to consume the reactants, this component will build up gradually within the plant because it has no place to leave the system.

3.2 The Plantwide Control Obstacle

3.2.1 Internal Feedback of Material and Energy

Processes that include recycle systems have an important design requirement-namely, that there must be an exit path for every chemical species. For example, inert components can be dept at reasonable levels by “bleeding off” a small portion of the recycle stream. However, Luyben (1994) discussed a subtle problem with recycle systems, the snowball effect, which is characterized by a large sensitivity of one or more of the variables in a recycle loop to small changes in a disturbance variable. This problem arises from both a small reactor holdup and a particular control structure.

In particular, if changes in fresh feed composition “load the reactor excessively”-that is, beyond its ability to provide the required conversion-then the separator/recycle system will be called on to make up the deficiency. For the case where extra reactor capacity is available through an increase in the reactor level, the particular choice of level/flow control structure within the recycle loop can radically affect plant gains (sensitivities). In the following, we assume that the reactor design is fixed and its level is set at less than full capacity. The question to be considered is how alternative designs of level and flow loops mitigate the effect of fresh feed flow rate or composition disturbances.

3.2.1.1 Steady-state Behavior: The Snowball Effect

Firstly, an interesting observation that has been made about recycle system is their tendency to exhibit large variations in the magnitude of the recycle flows. Plant operators report extended periods of operation when very small recycle flows occur. It is often difficult to turn the

equipment down to such low flow rates. Then, during other periods when feed conditions are not very different, recycle flow rates increase drastically, usually over a considerable period of time. Often the equipment cannot handle such a large load.

This high sensitivity of the recycle flow rates to small disturbances called the snowball effect. It is important to note that this is not a dynamic effect; it is a steady-state phenomenon, it can be analyzed by considering a steady-state model. However, it does have dynamic implications for disturbance propagation and for inventory control. There is nothing to do with closed-loop stability. Furthermore, this does not imply that it is independent of the control structure. On the contrary, the extent of the snowball effect is very strongly dependent upon the control structure used. The large swings in recycle flow rates are undesirable in a plant because they can overload the capacity of the separation section or move the separation section into a flow region below its minimum turndown. Therefore, it is important to select a plantwide control structure that avoids this effect.

3.2.1.2 Transient Behavior: The Slowdown in Overall System Dynamics

A second characteristic of using material recycle and/or heat integration is that the plant may respond to disturbances much more slowly than would be anticipated based on the time constants of individual units.

3.2.1.3 Propagation and Recirculation of Disturbances

A third major effect often encountered with recycle and heat integration is the propagation of disturbances form unit to unit, and their recirculation around internal process flow paths. This understanding why this plantwide control issue differs so substantially from single-unit issues.

In a single unit regulation, one desirable effect of using feedback control to attenuate disturbances in a process unit is to transfer these variations to a utility stream. For example, if a reactor temperature is disturbed, the cooling water flow rate will be changed by the temperature

controller so as to reduce the effect of the disturbance. Even so, some variation in reactor temperature inevitably will remain, and this will propagate to downstream units as a disturbance.

A minor side effect of these actions is that the supply header temperature itself will change slightly as cooling water demand is raised/ lowered by actions of a reactor temperature controller. Although utility supply systems are built with their own internal controllers, and these are designed to attempt to regulate the utility outputs in the face of process disturbances, it is not possible to attenuate utility disturbances entirely. These propagate directly throughout the plant.

In older plants, surge tanks were used to damp flow variations between units. Material holdup in a surge tank can also serve as a thermal capacitance and thus reduce effluent temperature variations; only reduced flow and temperature variations propagate to downstream units. In today's more highly integrated plants, containing material recycle and/or heat integration but little surge capacity, unattenuated disturbances propagate directly to downstream units, even to adjacent (coupled) units and to upstream units.

3.2.2 Interaction of Plant Design and Control System Design

In the past, when the continuous processing plants were designed to be much less interaction than now, it was possible to complete the plant design before considering control system design. After the proposed plant's flowsheet and equipment specifications were completed, process control engineers were responsible for specifying instruments and controllers. By providing an excess of measurements (instruments) and control valves, plus a feedback controller for every important process variable, the control system designer was reasonably sure that the new plant could be started up and controlled. Continuous processing plants designed or retrofitted today no longer can utilize a sequential design process in which plant design is followed by control system design (Keller and Bryan, 2000), nor can designers specify redundant equipment, except for safety purposes.

Without careful attention to design, highly integrated plants may have too few control degrees of freedom, which makes them difficult to start up and operate safely. For example, in designing the heat exchanger and related equipment for heat integration the heating and cooling loads first must be approximately balanced by the process designer. Then the designer must establish whether the approach temperatures are satisfactory to meet the heat transfer

requirements with a reasonably sized heat exchanger; in this step, adjustment of column operating pressures may be required (Seider et al., 2003). Because the energy supply capability in one unit usually will not balance the demand in another unit exactly, a “trim exchanger” (small heat exchanger sized to make up the difference in heating/cooling capability) generally has to be provided in the steady-state design.

A heat integration scheme also causes two control degrees of freedom to be “lost”: the cooling water flow rate control valve that would have been located in the Column 1 condenser, plus the steam control valve that would have been used in the Column 2 reboiler. If process control engineers are not involved in the plant design process from the beginning, the critical process dynamic and control evaluations may be omitted that would provide such information and an opportunity to resolve any problems (Keller and Bryan, 2000). In short, a suitably sized trim unit must be available to make up for any steady-state heating/cooling deficiency plus lost control degrees of freedom necessary for normal operations. It also can assist in start-up and shutdown operations.

The control system designer must determine whether a proposed plant design will be controllable and operable (Fisher et al., 1988b; Downs and Ogunnaike, 1995). For example, highly integrated distillation columns can cause problems in a number of ways:

1. One or both column products cannot be controlled at the desired set point(s).
2. Disturbances in the Column 1 overhead cannot be prevented from propagating to Column 2.
3. The “lost degrees of freedom” from plant integration need to be restored by the addition of one or two trim heat exchangers operated and controlled using plant utility supplies.
4. The plant cannot be started up easily because of the need to have Column 1 “hot” before Column 2 can be brought into service.

Consequently, there are three main functions of the control system, namely, disturbance rejection. It is the main objective in installing control system. The external disturbance is uncertain so the operator cannot monitor the changing in process. As a result, the control system must be installed to follow the changing of process and manipulate the process variable to compensate for the disturbance from external factors. Another function is to maintain stability.

The stability is necessary for every process. As a result the control system is set to improve the process stability for the guarantee of quality of product, safety to equipment of process and plant. The other is to keep the process performing highest efficiency. Besides rejecting disturbance and maintaining stability, the control system can achieve the great profit because it losses less energy and raw materials during the operating. Moreover, the product will meet the required specification and have high production rate.

3.3 Fundamental Procedures for Plantwide Control

3.3.1 Buckley Basic

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

1. Material balance control
2. Production quality control

He suggested looking first at the flow of material through the system. A logical arrangement of level and pressure control loop is established, using the flow rates of liquid and gas process streams. Note that most level controllers should be proportional-only (P) to achieve flow smoothing. He then proposed establishing the product-quality control loops by choosing appropriate manipulated variables. The time constants of the closed-loop product-quality loops are estimated. He tries to make these as small as possible so that good, tight control is achieved, but stability constraints impose limitations on the achievable performance.

3.3.2 Douglas doctrines

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

1. Minimize losses of reactants and products.
2. Maximize flow rates through gas recycle systems.

The first idea implies that the tight control of stream compositions exiting the process to avoid losses of reactants and product. The second rests on the principle that yield is worth more than energy. Recycles are used to improve yields in many processes. The economics of improving

yields (obtaining more desired products from the same raw materials) usually overbalance the additional energy cost of driving the recycle gas compressor.

3.3.3 Drowns drill

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

3.3.4 Luyben laws

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

1. To prevent the snowball effect, all recycle loops should be controlled flow.

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

3. If the final product from process comes out at the top of a distillation column, the column feed should be liquid. If the final product comes out the bottom of a column, the feed to the column should be vapor. Changes in feed flow rate or feed composition have less of a dynamic effect on distillate composition than they do on bottoms composition if the feed is

saturated liquid. The reverse is true if the feed is saturated vapor: bottom is less affected than distillate.

3.3.5 Richardson rule

Bob Richardson (1988) proposed the heuristic that the largest stream should be selected to control the liquid level in a vessel. This makes good sense because it provides more muscle to achieve the desired control objective. An analogy is that it is much easier to maneuver a large barge with a tugboat than with a life. The point is that the bigger the handle you have to affect a process, the better you can control it. This is why there are often fundamental conflicts between steady-state design and dynamic controllability.

3.3.6 Shinkey schemes

Greg Shinskey (1988) proposed a number of "advanced control" structures that permit improvements in dynamic performance. These schemes are not only effective, but they are simple to implement in basic control instrumentation. Liberal use should be made of ratio control, cascade control, override control, and valve-position (optimizing) control.

3.3.7 Tyreus tuning

The use of P-only controllers for liquid levels, tuning of a P controller is usually trivial: set the controller gain equal to 1.67. This will have the valve wide open when the level is at 80 percent and the valve shut when the level is at 20 percent. For other control loops, suggest the use of PI controllers. The relay-feedback test is a simple and fast way to obtain the ultimate gain (K_u) and ultimate period (P_u). Then either the Ziegler-Nichols settings or the Tyreus-Luyben (1992) settings can be used:

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

The use of PID controllers should be limited to those loops where two criteria are both satisfied: (1) the controlled variable should have a very large signal-to-noise ratio and (2) tight dynamic control from a feedback control stability aspect is very crucial. The classical example of the latter is temperature control in an irreversible exothermic chemical.

3.4 Step of Plantwide Control Design Procedure

Plantwide control design procedure is importantly satisfying the principles of the overall conservation of energy and mass. Furthermore, economic criterion is essentially taken into account.

Luyben et al., (1997) proposed a nine-step heuristic design procedure for a workable plantwide control strategy. The nine-step of the design procedure essentially concentrate on: 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.

Step 1: Establish control objectives

Assess the steady-state design and dynamic control objectives for the process.

Step 2: Determine control degree of freedom

Count the number of control valves available.

Step 3: Establish energy management system

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

Step 4: Set production rate

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

Step 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.

Step 6: Control Inventories (pressures and levels) and fix a flow in every recycle loop

Fix a flow in every recycle loop and then select the best manipulated variables to control inventories.

Step 7: Check component balances

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

Step 8: Control individual unit operations

Establish the control loops necessary to operate each of the individual unit operations.

Step 9: Optimize economics or improve dynamic controllability

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

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

Wongsri's plantwide control design procedure:

Step 1: Established control objectives.

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

Step 3: Selected manipulated variables and measurements via degree of freedom analysis.

Step 4: Energy management via heat exchanger network.

Step 5: Selection of control configuration using various tools available.

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

Step 7: Selection of controller type

Step 8: Validation via rigorous dynamic simulation.

Fixture Point Theorem

Hagglund (1995) present the real-time oscillation detection by calculates the integrated absolute deviation (IAE) between successive zero crossing of controller error signal. Its motivation is automatic monitoring of control-loop performance. The concept of material and energy disturbance propagation controls lead to fixture point theorem.

Fixture point theorem analysis:

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

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

3. Study the magnitude of integral absolute error (IAE) of all process variables that deviates from steady state.

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

Recently, the plantwide control procedure of Wongsri (Wongsri 2012) presented the eight-step of design procedure which used Fixture Point theorem.

Step 1: Gather relevant plant information and control objective including constraints for control.

Step 2: Plant Analysis.

2.1 Control degree of freedom (CDOF).List manipulated variables (control degree of freedom, CDOF).

List all control variables:

I. An independent stream must have a control valve (1 DOF) you cannot place two control valves on a single stream.

II. A heater, cooler, pump, or compressor has one degree of freedom (to adjust heat load or duty or work load)

III. A process to process heat exchanger has one degree of freedom by adding a by-pass line.

IV. A reactor has zero or one degree of freedom depends on its type. For example, an isothermal reactor need heat input to keep its temperature constant, while an adiabatic reactor has zero degree of freedom.

V. A flash separator has two degree of freedom.

VI. A simple distillation column has five degree of freedom.

How to pair manipulated and controlled variables:

VII. A control and manipulated variables must have strong causal relationship (high gain)

VIII. The manipulated variables should not be far from the control variables (zero or minimal dead time)

IX. The time constant of the quality loops should be short and the time constant of the inventory loops should be longer.

X. The manipulated variables should not be saturated for the whole range of the disturbances.

The change of the manipulated variables should not or have little effect on others variables (low gains with the remainder of the variables)

2.2 Heat pathway.

Three different "heat pathways" introduced in Luyben (1997) is also useful in plant analysis from a plantwide perspective. The first pathway is from inside the process and flows out to the environment heat generated by exothermic reactions and by degradation of mechanical work. A second pathway carries heat from utilities into the process and to the environment. The third pathway is internal to the process. The heat flow is circular and its magnitude depends upon the heating and cooling needs and the amount of heat integration implemented.

The level of heat circulated of the third pathway can be adjusted to optimize the energy used (step 7). The heat pathway is used to design control loops to reject the disturbances or to maintain the product qualities. This is done in step 4.1.

2.3 Material pathways.

The concept of material pathway is introduced here. The pathway is the flow path of a component from an entry point or an originated point to an exit point or an end point. The material pathway is useful for component balance and in control design as discussed in section 3 and section 4.2.

2.4 Material quantifiers.

The notion of material quantifier is also introduced here. In order to regulate a component balance in a process plant, a place representing the amount of material in the plant must be identified to provide its handle. A material balance for each component must be satisfied. From the viewpoint of individual units, chemical component balancing is not a problem because exit streams from the unit automatically adjust their flows and compositions. However, when we connect units together with recycle streams, the entire system behaves almost like a pure integrator in terms of the reactants. For example, we want to minimize the loss of reactants exiting the process since we would lose its value. This means we must ensure that most of

reactant fed to the process is consumed by the reactions. If reactants increase, the reactor conditions must be adjusted to consume more reactants. In the case that increasing one reactant composition will decrease the other reactant composition with an uncertain net effect on reaction rate. If this case is not effectively handled, the process will shut down when manipulated variable constraints are encountered in the separation section. Luyben (1997) gives a more complete discussion of this phenomenon. The features of material recycle and chemical component inventories mentioned above which have profound implications for a plant's control strategy (Luyben, 1997) can be handled quite readily by making use of quantifiers. By locating a quantifier, we can regulate the quantity of a component quantified by using its handle.

2.5 Reaction section (where to measure the extent of reaction)

2.6 Separation section (disturbances tests to find the best place to detect the disturbances.)

The sensitivity test is suggested to be done on the changing of composition, total flow, temperature, and component flow while keeping the reboiler heat duty and reflux flow or reflux ratio constants. This sensitivity test is to spot the tray with the largest changes in temperature from the initial steady state. This is the exact situation happen right after the disturbance entering the column, when the column temperatures are controlled by manipulating the reboiler heat duty and reflux flow or reflux ratio. Hence the trays with largest changes may be good locations to control.

Step 3: Establish fixture plant.

The principal idea of establishing a fixture plant is first to have an entire plant fluid-filled and a material-balanced. This idea is similar to creating *hydraulic* control structure proposed by Buckley (1964). By establishing a fixture plant we mean creating a material-balanced process plant:

3.1 Keep the materials entered and reentered fixed.

$$q_i(t)+q_r(t) = \text{constant} \tag{1}$$

This leaves the recycle streams free to adjust; one degree of freedom is added to the process.

If the composition of the recycle streams differ from the fresh feed stream significantly, each stream are separately controlled:

$$q_i(t) = \text{constant} \quad (2)$$

$$q_r(t) = \text{constant} \quad (3)$$

In this settlement, the flow of recycle stream cannot be used to regulate, e.g., the level of the reflux drum.

3.2 Regulate the production rate.

3.2.1) Consume the limiting reactant. The limiting reactant should be totally consumed at the reactor for the economic reason. Determine the most appropriate manipulate variable to control this, i.e. the reactor temperature, the reactor pressure, or the reactor holdup.

3.2.2) Regulate the production rate. The product rate can be regulated through 3.2.1. If this is done and the production rate does not reach the objective or the production demand, the limiting reactant feed rate must be increased. The reaction information about the accumulation or depletion of the limiting reactant must be used to determine the control strategy. However the design constraint may limit this strategy concerning increasing the reactant feed rate.

3.3 Adjust the flow of exit material streams (products, by-products, and inert) according to their accumulations.

$$q_o(t) = q_i(t) - dq/dt \quad (4)$$

3.4 Control the inventory of the rest of the component at their quantifiers, i.e. the indicators of the representative accumulation, for the rest of the components and design the control loops to regulate their inventories in the plant. The quantifier can be volume (mass), pressure, or flow rate.

$$q_p(t) = -dq/dt \quad (5)$$

$$q_p(t) = \text{constant} \quad (6)$$

In retrospect, the material balances are checked in this step, since the control loops generated accomplish the plantwide material balances. Therefore, it is guaranteed the plantwide inventory will be regulated.

Step 4: Disturbance Management.

In this step, the disturbances are handled by configuring the control loops employing the principle of disturbances management:

4.1 Heat Disturbance Management.

The Heat disturbance is divided into two categories. Heat Disturbance Category 1 (HDC1) is the heat disturbance that does not directly effect on product qualities, such as heat disturbance in a process stream toward to a heater, a cooler, or a process-to-process heat exchanger. Heat Disturbance Category 2 (HDC2) is the heat disturbance that will affect the product qualities, such as heat disturbance in a process stream toward to a reactor or a separator.

4.1.1 Direct the thermal disturbances that are not directly related to quality to the environment via the next and nearest exit points, usually heaters or coolers, to keep the thermal conditions of process stream fixed. The thermal condition of process stream is changed along the process plant, usually by heater or cooler of process to process heat exchanger.

4.1.2 Manage the thermal disturbance that related to quality in order to maintain the product specification constraints.

4.2 Material Disturbances Management.

The configuration of the control loops depend on the desired material pathways. The pathways can be obtained by analyzing the results of the material disturbance tests. The material disturbances can be generated at reactors and separators, besides coming with feeds and recycle streams. So if the feeds and recycle streams are fixed, the only places that alter the material (total or component) flow rates are the reactors and the separators. At reactor, its inlet temperature is adjusted in order to keep the reactor component flow rate or its composition in outlet stream. The decision of whether how to choose to control the component flow or the composition or not to control is based on the profit maximization or the smooth operation policies.

The control structure we select must reject the disturbance to the desired pathways. As in the case of heat disturbance management, we direct the material disturbances to the environment via the next and nearest exit points, usually separators, to keep the material conditions of process stream fixed.

How to direct material disturbance? At a splitter (e.g. a distillation column), we must decide which paths to push extra loads or disturbances to. It is depend on how we want to manage the extra loads to keep the plant running smoothly and the quality of the products. For example, we don’t want to push the extra loads to the product stream. It is always designable to reject the disturbance out of the process plant as soon as possible. Thirdly, we prefer to keep the recycle flow constant in the case that its composition differs significantly from the make-up feed.

However if this is not allowed, we must trade-off between pushing the extra loads to the recycle stream and keeping it constant.

Ratio Control on Feeds. Add ratio control to accommodate the variation of one of the fresh feed in the case that the two feeds must be proportional.

Single-end Control, **Which End?** Since the distillation columns, usually the one-point control is common. To control top or bottom temperatures, depend on the material disturbance rejection policy. The control policy of a distillation column is to reject or direct disturbances to the designated pathways. For example, a recycle distillate flow must be maintained. Product purity must be maintained.

Fixing Reflux flow, Reflux Ratio or Reflux to feed. To aid in making this choice, a series of dynamic simulation runs can be made in which the effects of changes in composition, temperature, total flow and component flow of distillation column feed.

Single-end or dual-end control. If there are two locations with large changes in the temperature profiles when the sensitivity test is performed (see Step 2.6), so it may be possible to use dual-temperature control if this structure is required.

Step 5: Design the rest of the control loops.

5.1 Design the control loops for the remaining control variables, i.e. the rest of the inventories.

5.2 Adding enhanced controls, i.e. cascade, feed forward controls.

Step 6: Energy management via heat exchanger networks.

If potential heat exchanger networks or alternative heat integrated processes (HIPs) exist, list additional control variables and manipulated variables.

Step 7: Optimize economics and/or improve control performance. For example, the controls scheme/structure of the reactor (e.g. temperature/composition sensor location), the control scheme of the distillation column (e.g. reflux to feed ratio control), the optimal operating temperatures of the reactors, the recycle flow rates, the sequence of separation, etc. If the opportunity of optimization exists, we might backtrack to the previous step as dictated.

Step 8: Validate the designed control structures by rigorous dynamic simulation. The measures can be costs, raw material and energy consumptions, control performances of the total plant or some selected loops, etc.

3.5 Heat Exchanger and Plantwide Energy Management

Another important issue in process design is energy conservation. Common ways to improve the conservation is to install feed-effluent heat exchangers (FEHEs) around rectors and distillation columns where one streams is heated, another must be cooled. For instance, in HDA process, the toluene fresh feed, the makeup hydrogen, the recycle toluene, and the recycle gas stream needed to be heated up to the required reaction temperature. And, the reactor effluent stream must also be cooled to the cooling water temperature to accomplish a phase split. So the energy integration is required to reduce the utility cost in addition to improve thermodynamic efficiency of the process.

3.5.1 Heat Exchanger Dynamics

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

3.5.2 Heat Pathway

In the process, the energy required for heating certain streams can be matched by similar amount of energy required for cooling other streams. Heat recover from cooling a stream could be recycling back to the process to heat another stream. This is the proposed of heat integration and heat exchanger networks (HENs).

From a plantwide perspective, the heat pathways in the process can be separated into three different paths as illustrate in Fig. 3.2. The first pathway shows the heat expend to the environment generated by exothermic reaction and by degradation of mechanical work. This pathway is from inside the process to outside. It is also possible to convert some of the heat to work as it is removed from high temperature in the process.

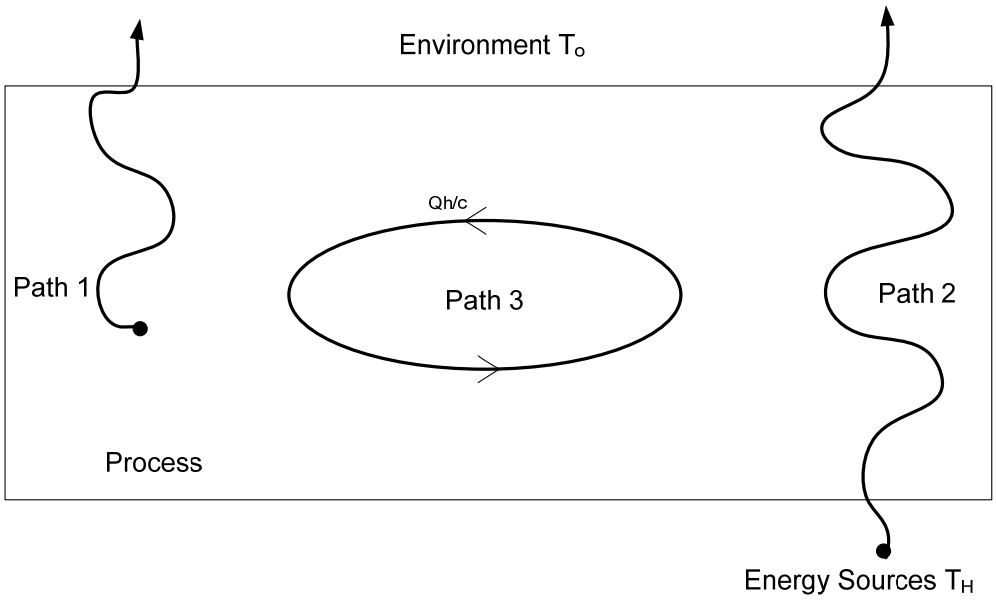


Figure 3.2 Heat Pathways.

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

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

3.5.3 Heat Recovery

The great improvements in the plant's thermal efficiency are made by recycling much of the energy needed for heating and cooling process streams. There is of course a capital expense associated with improved efficiency but it can usually be justified when the energy savings are accounted for during the lifetime of the project. The current context draws attention on how heat integration affects the dynamics and control of a plant and how energy in plants can be managed with a high degree of heat recovery.

3.6 Control of Process-to-Process Exchangers

Process-to-process (P/P) exchangers are employed for heat recover within the process. Two exit temperatures can be controlled provided that the two inlet flow rates can be manipulated separately. Though, these flow rates are normally unavailable to manipulate. Therefore two degrees of freedom are given up fairly easily. It is possible to oversize the P/P exchanger and provides a controlled bypass around it as in Fig. 3.3a. It is possible to combine the P/P exchanger with a utility exchanger as in Fig. 3.3b.

Figure 3.3 Control of P/P heat exchangers**;** (a) use of bypass; (b) use of auxiliary utility exchanger

3.6.1 Bypass Control

When the bypass method is employed for unit operation control, several choices about the bypass location and the control point are considered. Fig. 3.4 shows the most common options. The question like “Which option is best?” may arise. The best alternative depends on how “best” is defined. As many other examples, it reduce the trade-off between design and control. Design considerations might suggest that the cold side is measured and bypass 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 perspective the most important stream should be measured, regardless of temperature, and bypass on the same side. This minimizes the effects of heat exchanger dynamics in the loop. A large fraction of the controlled stream should be bypass as it improves control range. Hence a large heat exchanger is required.

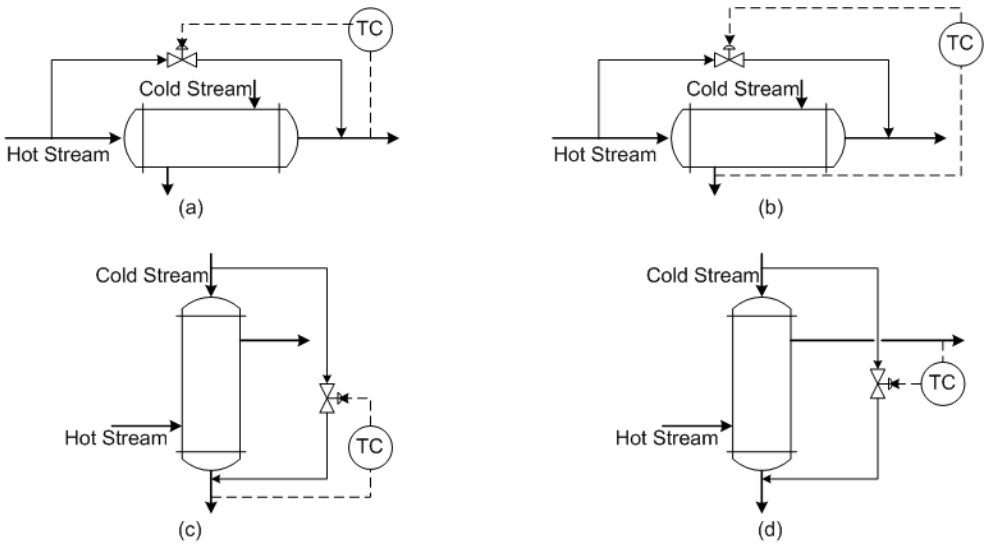


Figure 3.4 Bypass controls of process-to-process heat exchangers. (a) Controlling and bypassing hot stream; (b) Controlling cold stream and bypassing hot stream; (c) Controlling and bypassing cold stream; (d) Controlling hot stream and bypassing hot stream.

3.6.2 Use of auxiliary utility exchangers

There are a few design decisions to make, when the P/P exchanger is combined with a utility exchanger. First, the relative sizes between the recovery and the utility exchangers have to be established. From a design perspective, it is needed to make the recovery exchanger large and the utility exchanger small. This leads to the most heat recovery, and it is also the least expensive option from an investment standpoint. Though, a narrow control range and the inability to reject disturbances make this choice the least desirable from a control standpoint.

Second, decision on how to combine the utility exchanger with the P/P exchanger must be made. This could be done either in a series or parallel arrangement. Physical implementation

issues may dictate this choice but it could affect controllability. Finally, decision on how to control the utility exchanger for best overall control performance must be made.

Consider a distillation column that uses a large amount of high-pressure stream in its thermosiphon reboiler. Heat-integrate this column with the reactor is applied to reduce operating costs. A virtual way of doing this is to generate stream in a waste heat boiler connected to the reactor as suggested. Some or all of this steam can be used to help reboil the column by condensing the stream in the tubes of a stab-in reboiler. Nevertheless, the total heat from the reactor may not be enough to reboil the column, so the remaining heat must come from the thermosiphon reboiler, which serves as an auxiliary reboiler. The column tray temperature controller would manipulate the stream to the thermosiphon reboiler.

CHAPTER IV

ISOPROPYL ALCOHOL DEHYDRATION PROCESS

4.1 Introduction

Solvents are liquid organic compounds used on a large scale in industry in various functions. The coatings industry consumes nearly 50% of the world's solvent production. In the production and recycling of solvents, dehydration plays a major role. Solvents or solvent mixtures very often form azeotropes with water, which cannot be separated by simple distillation. Their treatment requires special thermal processing such as two-pressure, azeotropic, or extractive distillation.

4.2 Process Description

The feed stream is a binary mixture of 2-propanol (isopropyl alcohol, IPA) and water. This nonideal mixture has an azeotropes that makes it impossible to obtain complete separation in a single column. The composition of the minimum-boiling homogeneous azeotrope at a pressure of 1 atm is 87.6 wt % IPA, with a temperature of 353.4 K. The normal boiling point of IPA is 355.4 K.

The conventional process presented by Sommer and Melin (2004) uses extractive distillation with a heavy entrainer. The flowsheet features three distillation columns, as shown in Figure 4.1. The feed is 1875 kg/h of an 80 wt % IPA and 20 wt % water binary mixture at 350 K. It is fed on Stage 10 of a 19 stage (17 trays) column C1 operating at 1.1 atm condenser pressure, giving top and bottom temperatures of 356 and 379 K, respectively. The reflux ratio is 0.637. The bottoms product is 99.99 wt % water. The distillate is 86.14 wt % IPA, which is close to the azeotropic composition. Reboiler heat input is 724 kW.

The distillate is fed to Stage 30 of the 60-stage second column C2. An extractive solvent of ethylene glycol is fed on Stage 7 at a flow rate of 5,000 kg/h, and a temperature of 373 K. The reflux ratio is 2.04. The specifications for this column are a distillate product that is 99.91 wt % IPA, and a bottoms stream with ppm concentrations of IPA. The column pressure is 0.8 atm,

giving top and bottom temperatures of 350 and 437 K, respectively. The reboiler heat input is 1148 kW.

The bottoms of C2 are fed on Stage 12 of the third column C3, which has 25 stages and operates at 0.4 atm, giving top and bottom temperatures of 349 and 447 K, respectively. The distillate product is 99.98 wt % water. The bottoms stream is essentially pure ethylene glycol, which is cooled and recycled back to C2. The reboiler heat input is 440 kW, and the reflux ratio is 1.5. Temperature profile of each column are shown in Fig.4.2

Table 4.1 gives details of the design parameters and operating conditions at steady state.

Table 4.1 Design parameters for three columns

		C1	**C2**	**C3**
Reflux Ratio		0.637	2.04	1.5
Flow (kg/h)	D	1742	1501	241
	B	133	5241	5000
	F	1875		
	Solvent		5000	
Compositions				
(wt% IPA)	Feed	80		
(wt% IPA)	Distillate	86.1	99.91	0.02
(wt% H2O)	Bottoms	99.99	4.59	
Pressure (atm)		1.1	0.8	0.4
Diameter (m)		0.584	0.609	0.552
Total Stage		19	60	25

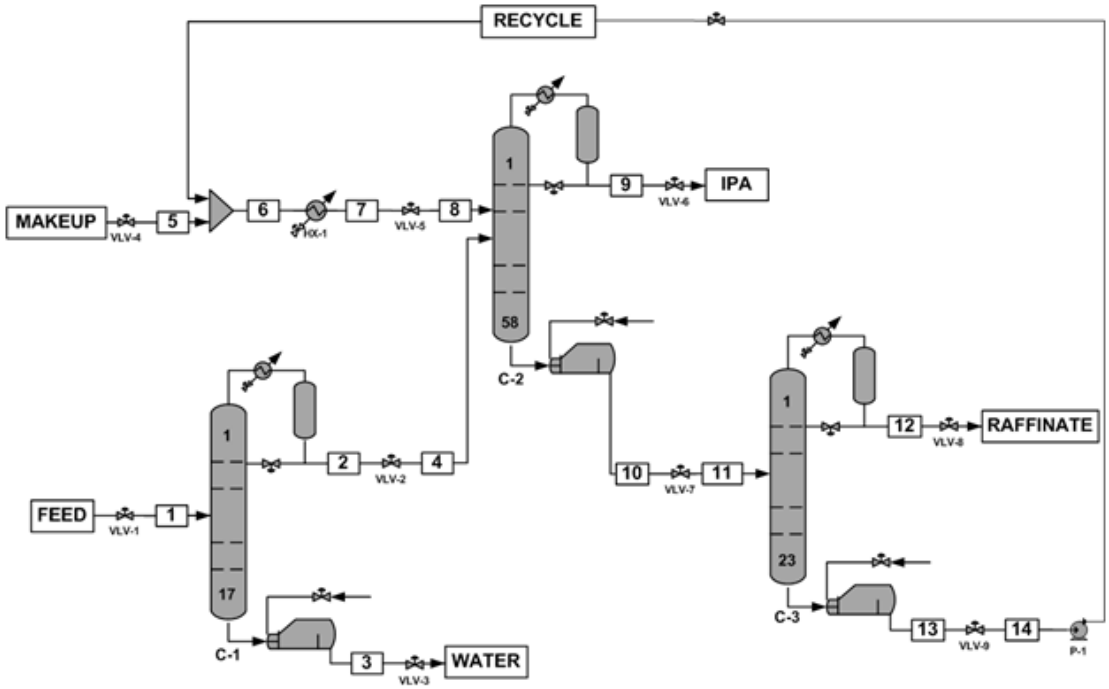
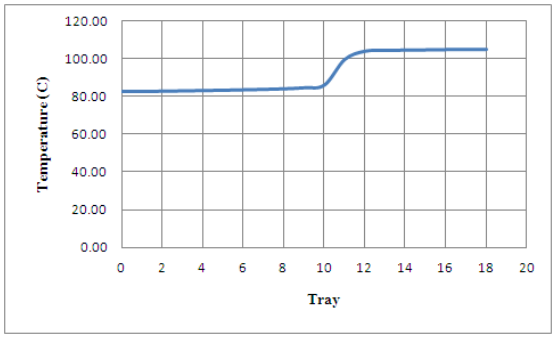
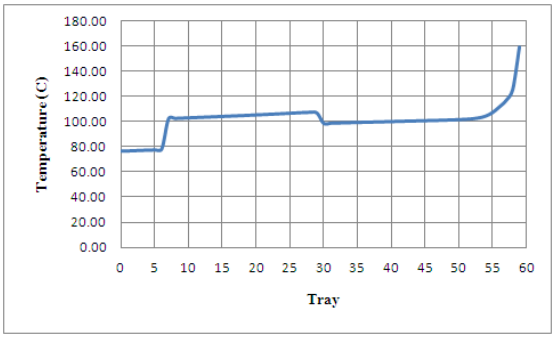


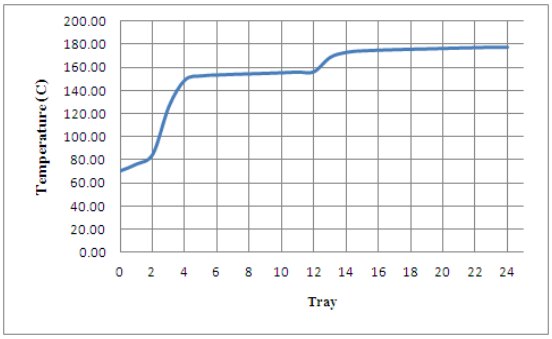
Figure 4.1 Isopropyl Alcohol Dehydration Process Flowsheet



(a)



(b)



(c)

Figure 4.2 Temperature profile. (a) Temperature profile of column C1.(b) Temperature profile of column C2. (c) Temperature profile of column C3.

CHAPTER V

CONTROL STRUCTURE DESIGN AND DYNAMIC SIMULATION

5.1 Control structure design

This research follows the 8-step of Wongsri's plantwide control design procedure for designing the control structure of isopropyl alcohol dehydration process.

Step 1*:* Starting with the process information given in section 4.2 and the process conditions given in Appendix A. The purity of the IPA product at 98 wt% is essentially requisites to meet process objectives. The essentially pure ethylene glycol in recycle stream is process constrain.

Step 2*:* Plant Analysis.

2.1 Control degree of freedom (CDOF).List manipulated variables (control degree of freedom, CDOF). The CDOF can be obtained using the guideline given in section 3.4.

Table 5.1 The control degree of freedom for the isopropyl alcohol dehydration process.

Unit	Manipulated Variables	Quantity	DOF
Independent Stream	Mass Flow Rate	2	2
Cooler	Cooler Heat Removal	1	1
Distillation Column	Reflux Flow		
	Distillate Flow		
	Bottom Flow	3	15
	Condenser Heat Removal		
	Reboiler Heat Input		
		Total	**18**

2.2 Heat pathway. Heat flows into the process carry from feed stream and makeup ethylene glycol are 1.4e7 kJ/h and 3.6e7 kJ/h, respectively. Heat flow out of the plant by bottom product of column C1, distillate product of column C2 and distillate product of column C3 are

2.1e6, 7.8e6 and 3.6e6 kJ/h, respectively. The heat pathways that circulate in the plant are shown in Fig.5.1.

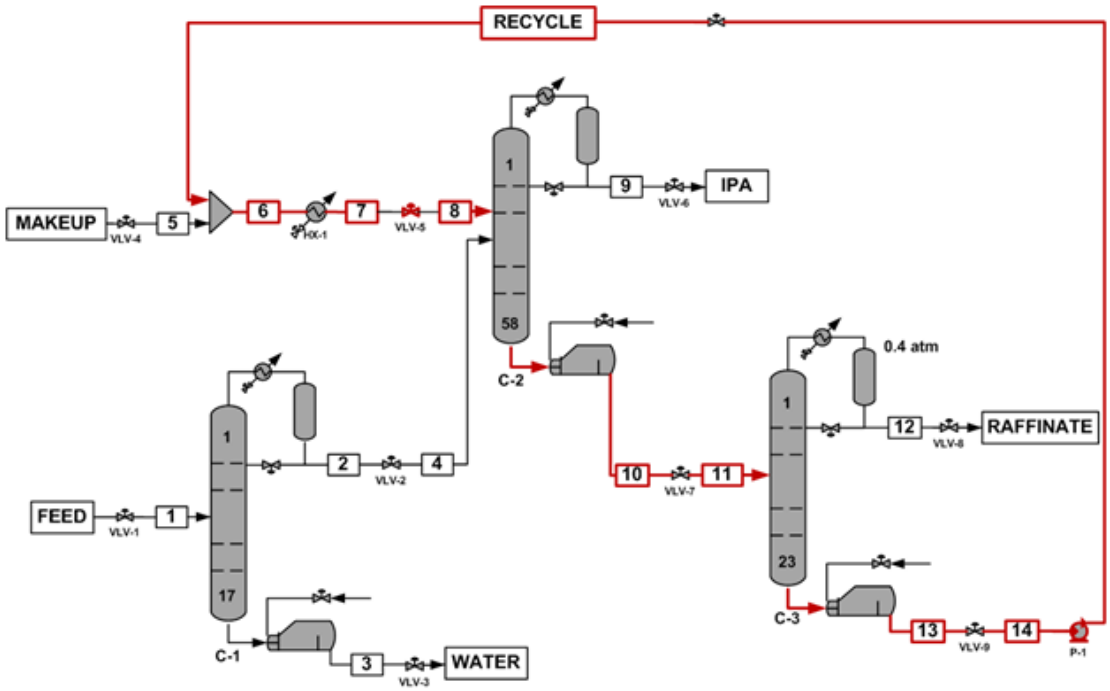


Figure 5.1 Heat pathway circulate in the plant

2.3 Material pathway. The pathway is the flow path of a component from an entry point or an originated point to an exit point or an end point. The material pathway is useful for component balance and in control design.

The water pathway is shown in Fig.5.2. The IPA pathway is presented in Fig.5.3. Finally, the ethylene glycol pathway is illustrated in Fig.5.4.

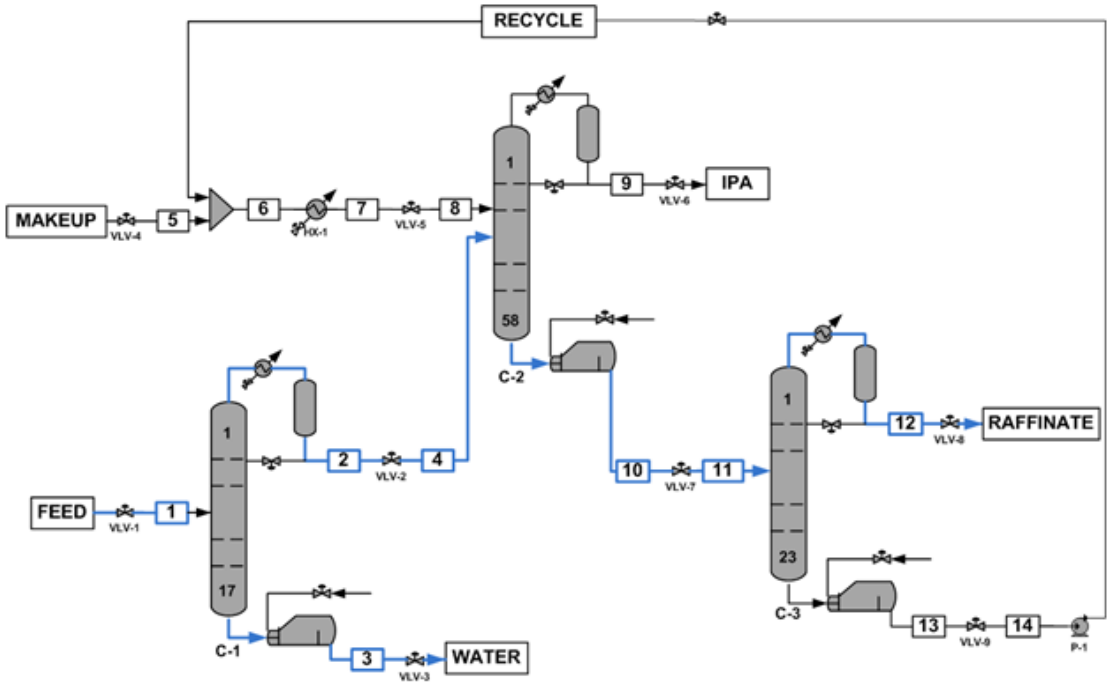


Figure 5.2 Water pathway

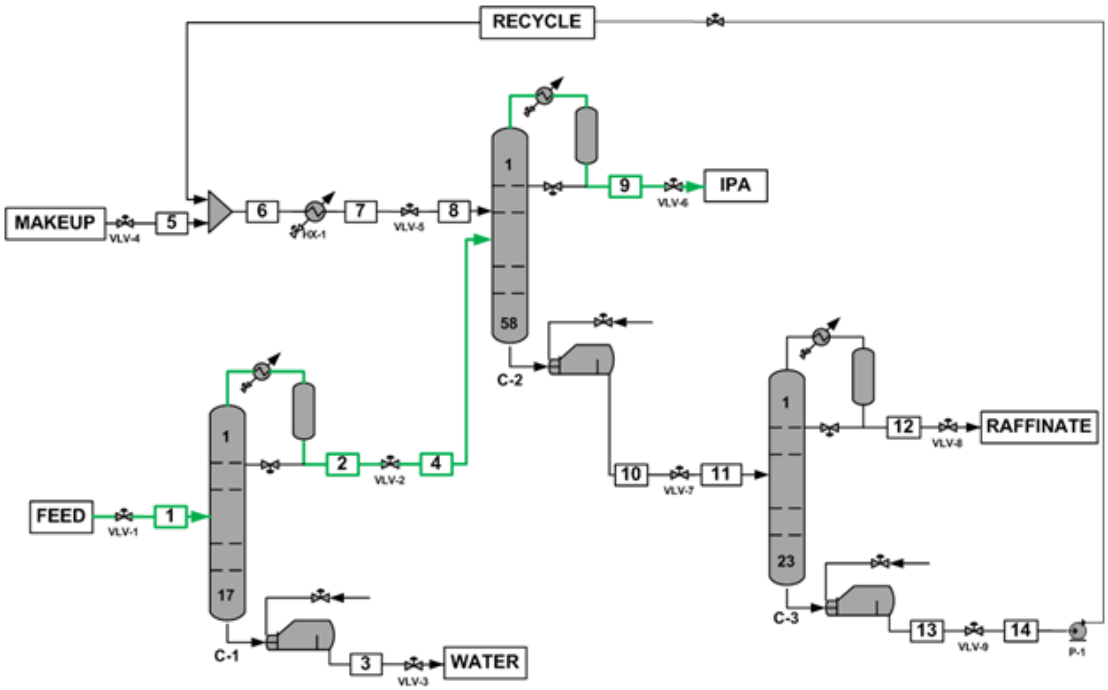


Figure 5.3 IPA pathway

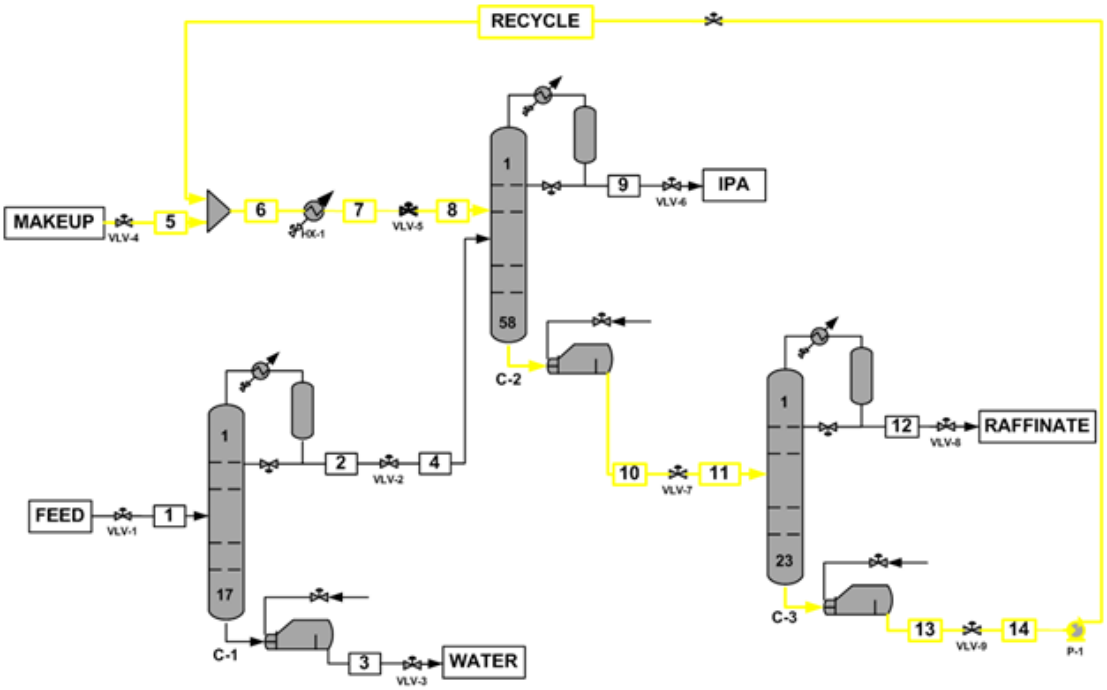


Figure 5.4 Ethylene glycol pathway

2.4 Material quantifiers. Material quantifiers is a place representing the amount of material in the plant must be identified to provide its handle.

The quantifiers of each component are show in Fig.5.5.The quantifier of water and IPA is the feed stream. The quantifier of ethylene glycol is combined of makeup and recycle stream.

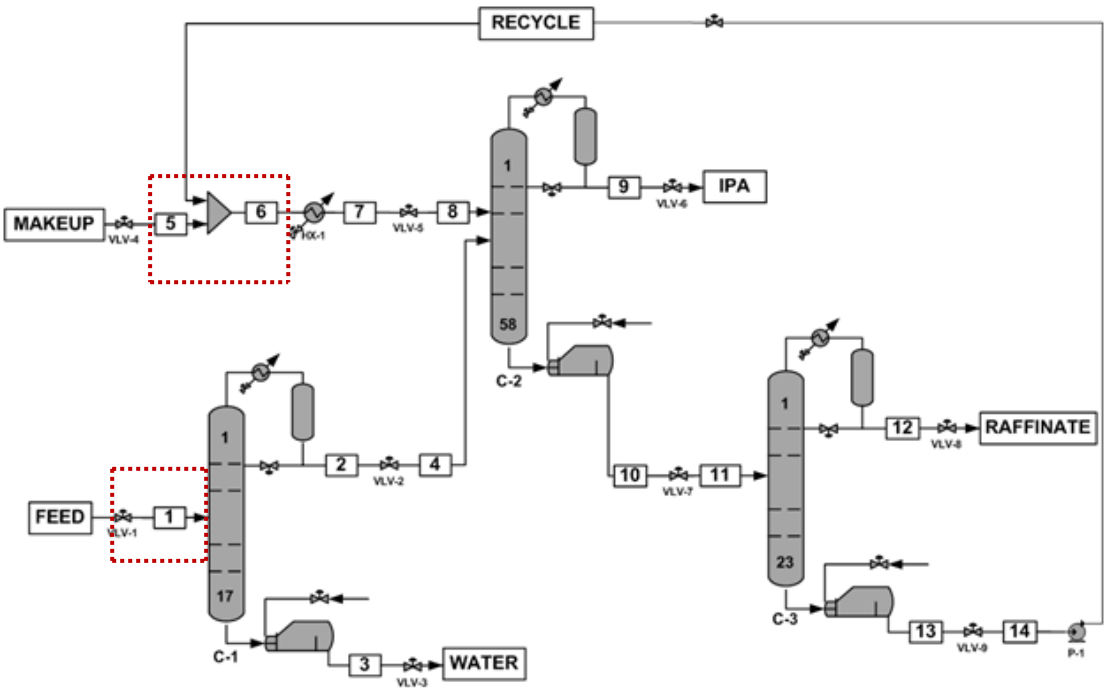


Figure 5.5 Material quantifiers

2.5 Reaction section. In this process there is no reaction section.

2.6 Separation section.

The main idea of Separation section analysis is spot the tray with the largest changes in temperature from the initial steady state by keep manipulated variables constant. The sensitivity test is suggested to be done on the changing of composition, component flow, total flow and temperature while keeping the reboiler heat duty and reflux flow or reflux ratio constants.

The sensitivity tests of Column C1 are show in Fig.5.6. The sensitivity tests of Column C2 are presented in Fig.5.7. Finally, the sensitivity tests of Column C3 are illustrated in Fig.5.8.

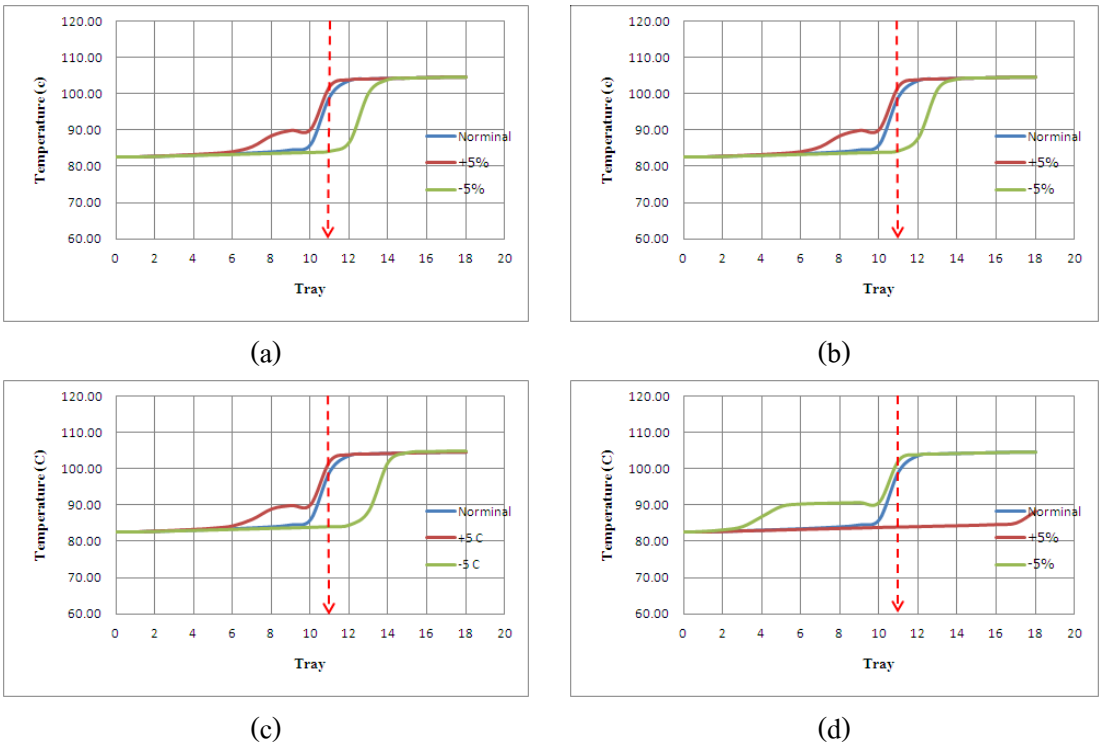


Figure 5.6 The sensitivity tests of Column C1. (a) Component flow change. (b) Composition change. (c) Temperature change. (d) Total flow change.

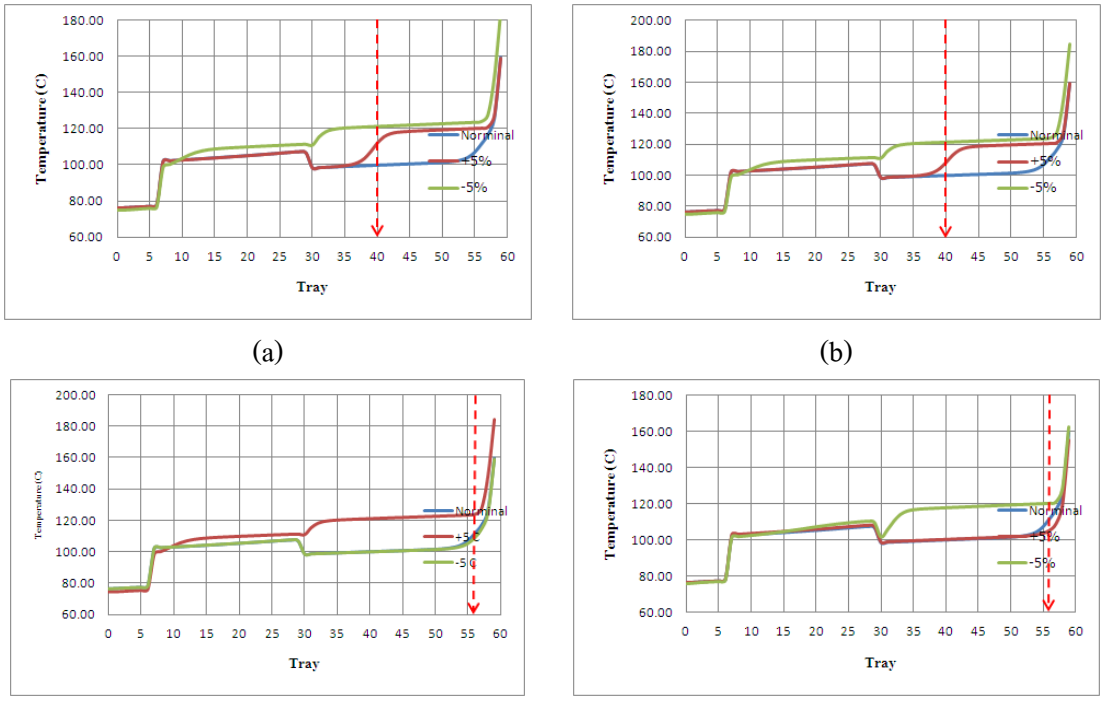


Figure 5.7 The sensitivity tests of Column C2. (a) Component flow change. (b) Composition change. (c) Temperature change. (d) Total flow change.

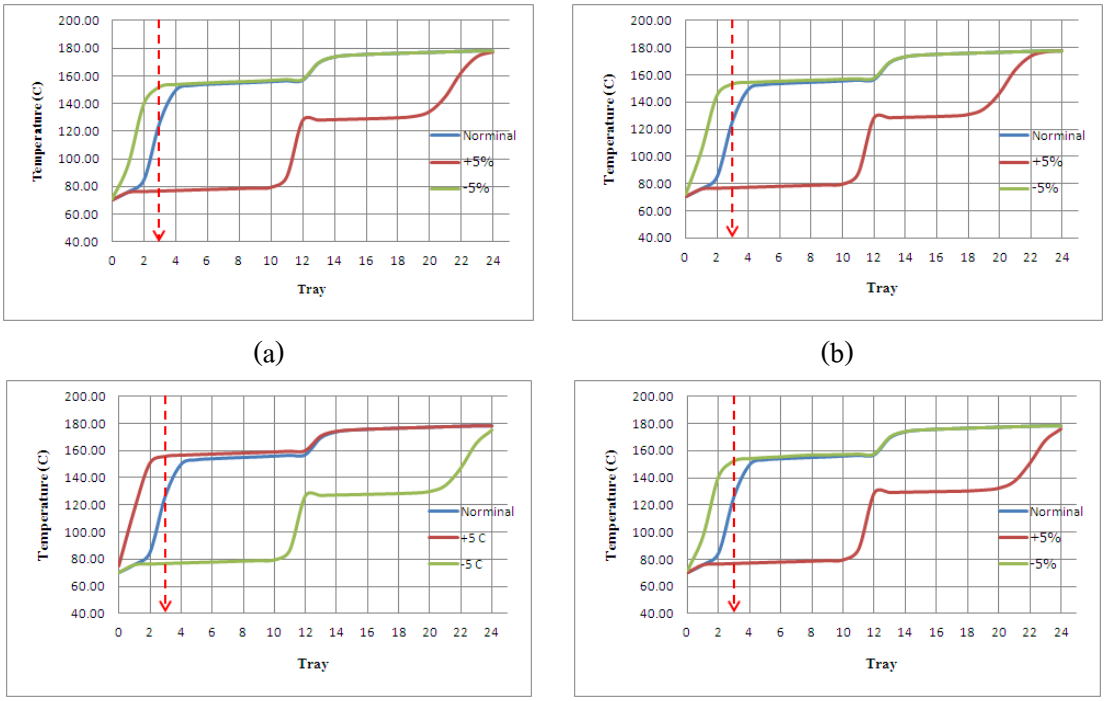


Figure 5.8 The sensitivity tests of Column C3. (a) Component flow change. (b) Composition change. (c) Temperature change. (d) Total flow change.

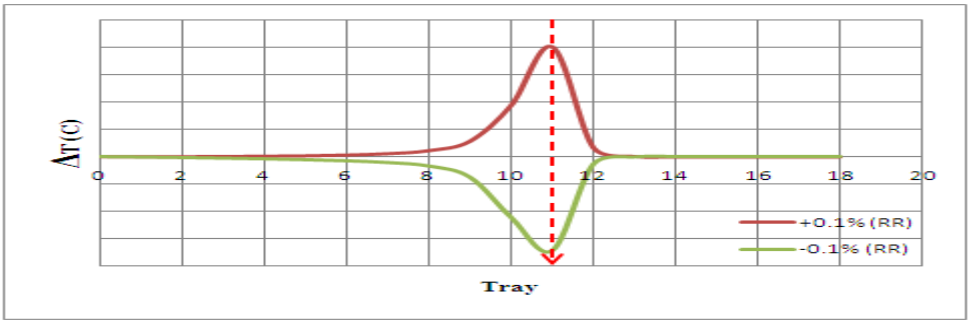
From the sensitivity tests of Column C1 (Fig 5.6) the largest change in temperature from the initial steady state is tray 11 so control tray of Column C1 is tray 11.The sensitivity tests of Column C2 (Fig 5.7) the largest change in temperature from the initial steady state is two difference point tray 40 and 56.From dynamic simulation by use tray 40 is control tray, the structure cannot remain composition of IPA product at specific constrain while control tray is tray 56 can remain the composition at specific constrain. So control tray of Column C2 is tray 56. Finally the sensitivity tests of Column C3 (Fig 5.8) the largest change in temperature from the initial steady state is Tray 3. Control tray of Column C3 is tray 3.

The sensitivity tests by changing of composition, total flow, temperature, and component flow while keeping the reboiler heat duty and reflux flow or reflux ratio constants are compared with design spec/vary tests.

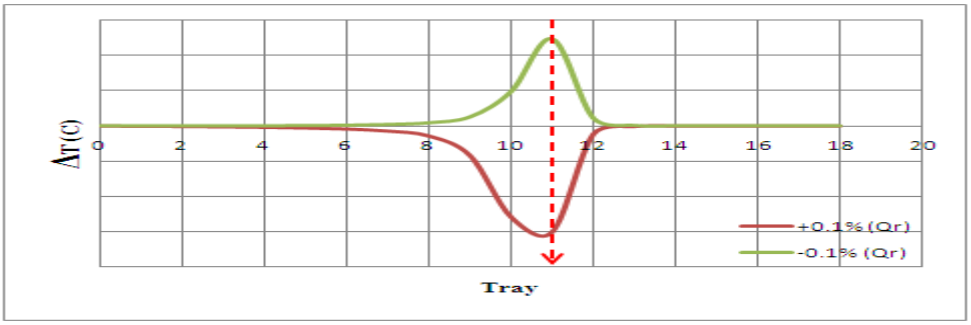
The design spec/vary tests is presented by Luyben. The main idea of design spec/vary tests is similar to sensitivity tests, find the largest change in temperature from the initial steady state by vary manipulated variables.

The design spec/vary test suggested by keeping the reboiler heat duty constant vary reflux ratio and vary the reboiler heat duty keep reflux ratio constant.

The design spec/vary test of Column C1 are show in Fig.5.9. The design spec/vary test of Column C2 are presented in Fig.5.10. Finally, the design spec/vary test of Column C3 are illustrated in Fig.5.11.

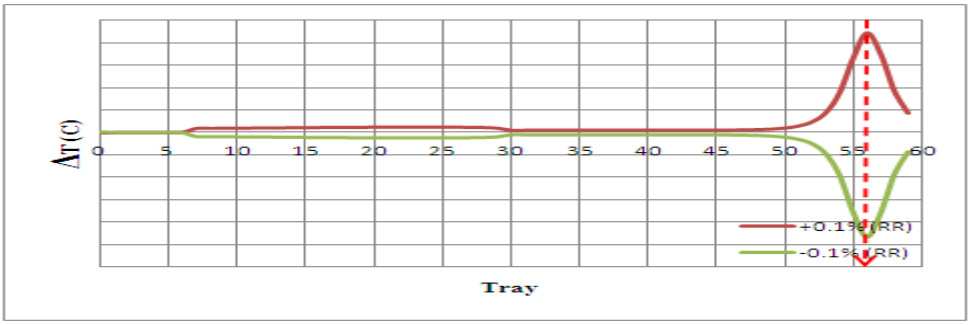


(a)

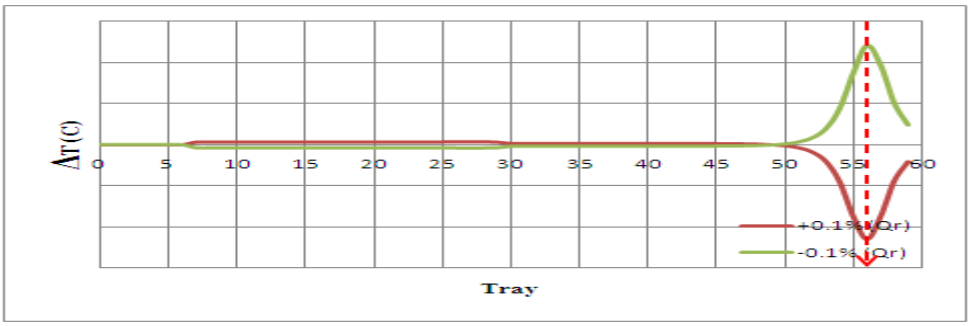


(b)

Figure 5.9 The design spec/vary test of Column C1. (a) Keep the reboiler heat duty constant vary reflux ratio. (b) Vary reboiler heat duty keep reflux ratio constant.

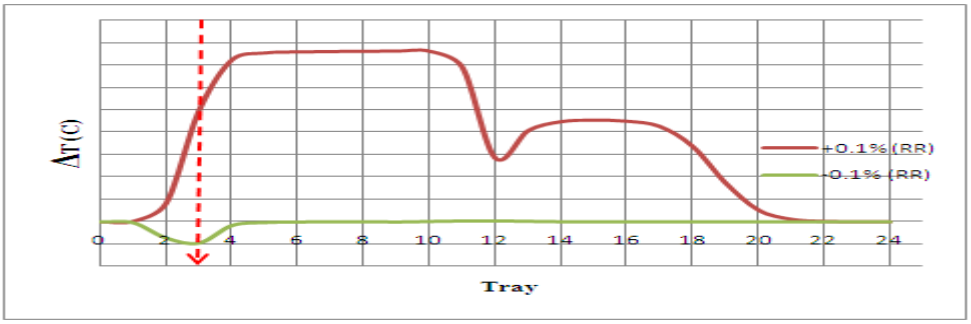


(a)

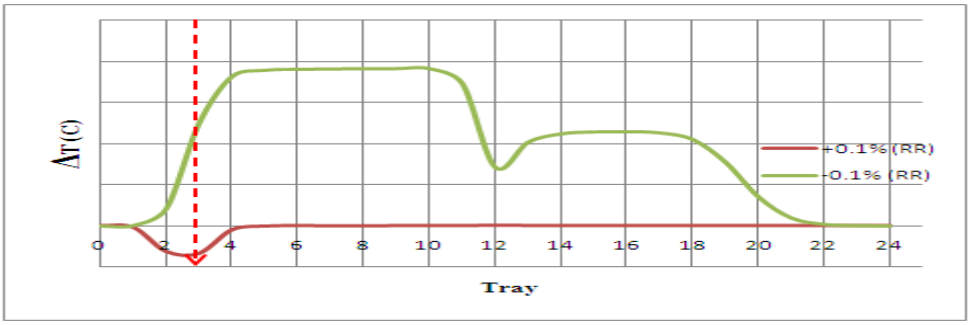


(b)

Figure 5.10 The design spec/vary test of Column C2. (a) Keep the reboiler heat duty constant vary reflux ratio. (b) Vary reboiler heat duty keep reflux ratio constant.



(a)



(b)

Figure 5.11 The design spec/vary test of Column C3. (a) Keep the reboiler heat duty constant vary reflux ratio. (b) Vary reboiler heat duty keep reflux ratio constant.

From the design spec/vary tests of Column C1 (Fig 5.9) the largest change in temperature from the initial steady state is tray 11. The design spec/vary tests of Column C2 (Fig 5.10) the largest change in temperature from the initial steady state is tray 56.Finally the design spec/vary tests of Column C3 (Fig 5.11) the largest change in temperature from the initial steady state is tray 3. From sensitivity tests and design spec/vary present the largest change in temperature of Column C1, C2 and C3 are the same point.

The difference point of the largest change in temperature from the sensitivity tests and design spec/vary in Column C2 cause by effect in vapor fraction of feed.

Step 3: Establish fixture plant.

3.1 Keep the materials entered and reentered fixed. The objective of this step is to keep the stream entered (fresh and recycled) the plant fixed; therefore the feed is regulated the by adjusting flow rate of the feed. The combined of ethylene glycol is regulated the by adjusting flow rate of the makeup feed. The control loops to keep the materials entered and reentered fixed are represented in Fig.5.12.

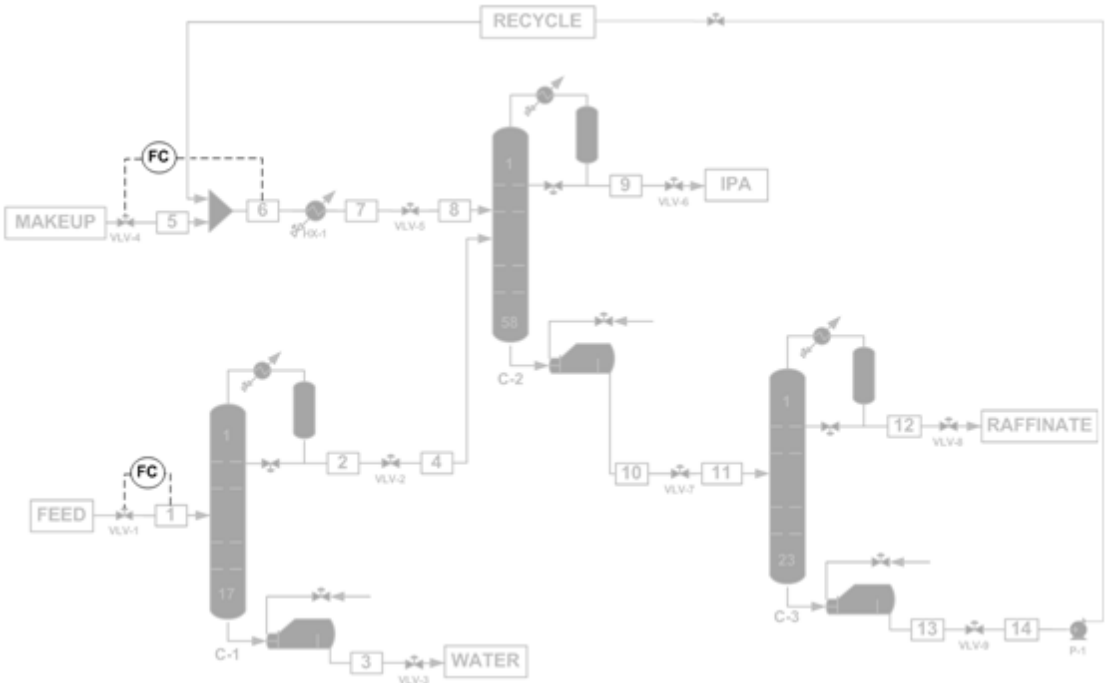


Figure 5.12 Control loops to keep the materials entered and reentered fixed

3.2 Regulate the production rate.

3.2.1) Consume the limiting reactant.

3.2.2) Regulate the production rate.

3.3 Adjust the flow of exit material streams. The exit material streams of the isopropyl alcohol dehydration process are the bottom product of column C1 (water), the distillate product of column C2 (IPA) and the distillate product of column C3 (water). The bottom product of column C1 is adjusted by the base level control of the column. The distillate product of column C2 is adjusted by the level control of reflux drum. The distillate product of column C3 is adjusted by the level control of reflux drum. The obtained control structure is shown in Fig.5.13.

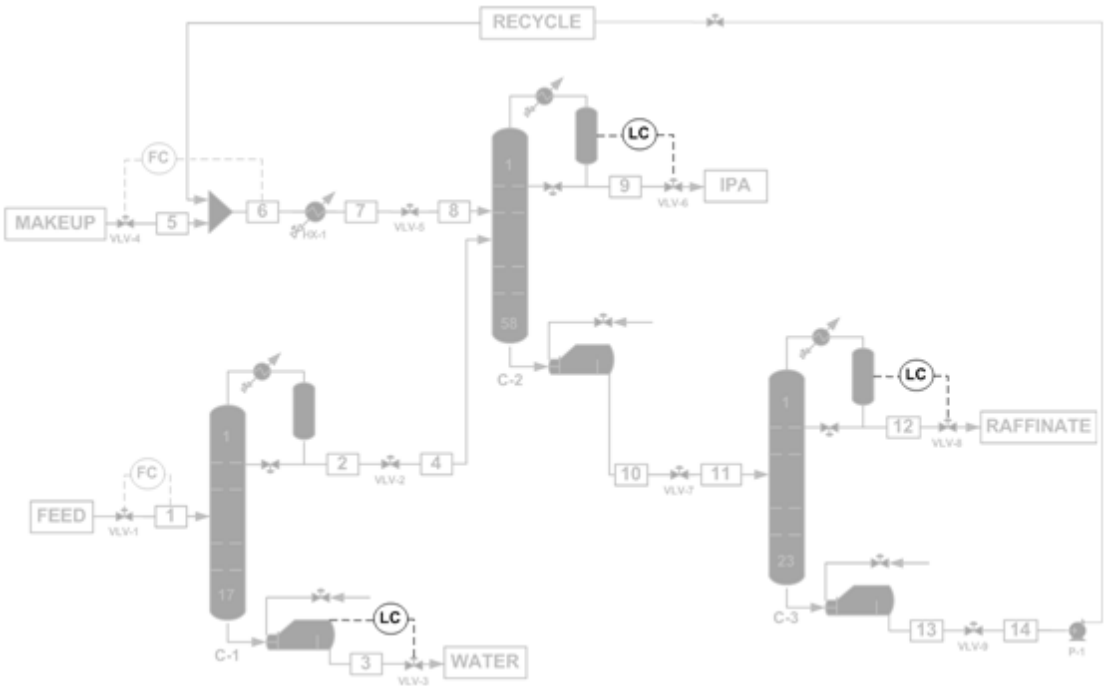


Figure 5.13 Control loops for adjust the flow of exit material streams

3.4 Control the inventory of the rest of the component at their quantifiers. All components have their quantifiers discuss in step 3.1 and 3.3.

Step 4: Disturbance Management.

4.1 Heat Disturbance Management. The thermal disturbance is divided into 2 categories: Heat Disturbances of Category1 (HDC1) is handled by HX-1 regulates the temperature of the combined ethylene glycol. Heat Disturbances of Category 2 (HDC2) are those presented in the feed to the separation columns. The three column temperatures are control by adjusting the reboiler heat input. The results control loops are shown in Fig.5.14.

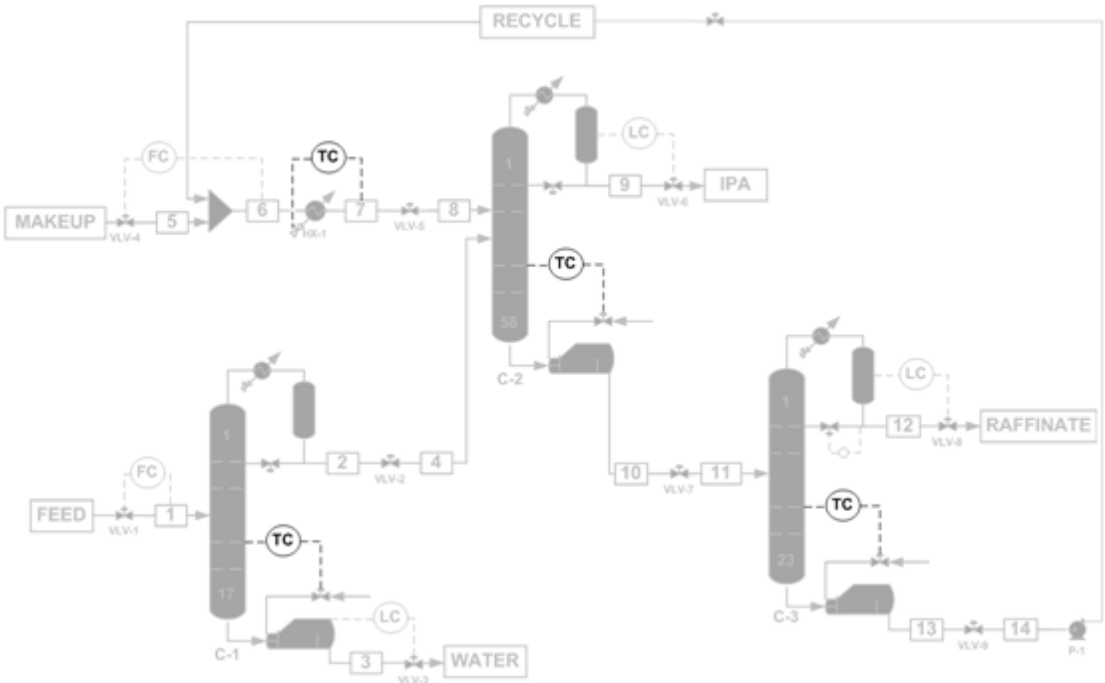


Figure 5.14 Control loops for handle heat disturbance

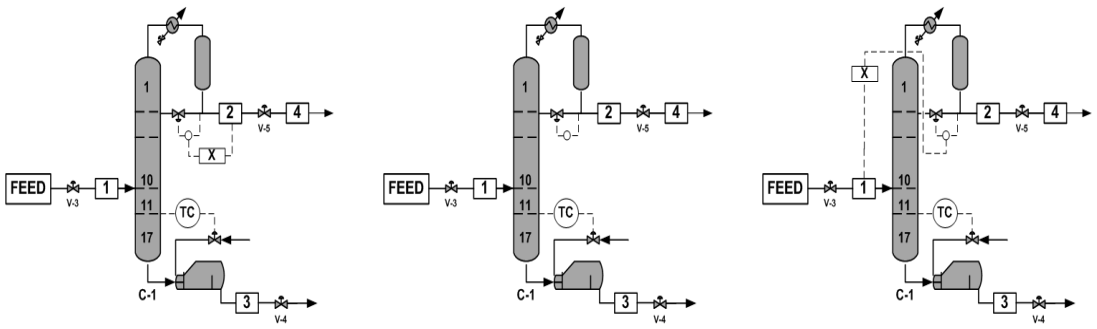
4.2 Material Disturbances Management.

The main ides of material disturbances management is find the configuration which direct extra load (disturbance) to their pathways. The configuration of the control loops depend on the desired material pathways. The pathways can be obtained by analyzing the results of the material disturbance tests.

The results of disturbance test of Column C1 are show in Fig.5.18-5.21. The results of disturbance test of Column C2 are presented in Fig.5.22-5.25. Finally, the results of disturbance test of Column C3 are illustrated in Fig.5.26-5.29.

Material disturbances management in Column C1: disturbance of Column C1 is water. IPA and Water are fed to Column C1 bottom product is 99.9 wt% water. So the configuration for material disturbances management of Column C1 must force the disturbance to bottom product.

The alternate control schemes of Column C1 are shown in Fig.5.15. Blue line represents reflux ratio constant, red line represents reflux flow constant and black line represents reflux to feed ratio.



(a) (b) (c)

Figure 5.15 Alternate control schemes of Column C1. (a) Reflux ratio constant (b) Reflux flow constant (c) reflux to feed ratio

From component flow and composition change in Column C1 (Fig. 5.18 and 5.19) all structure can force the disturbance to bottom product.

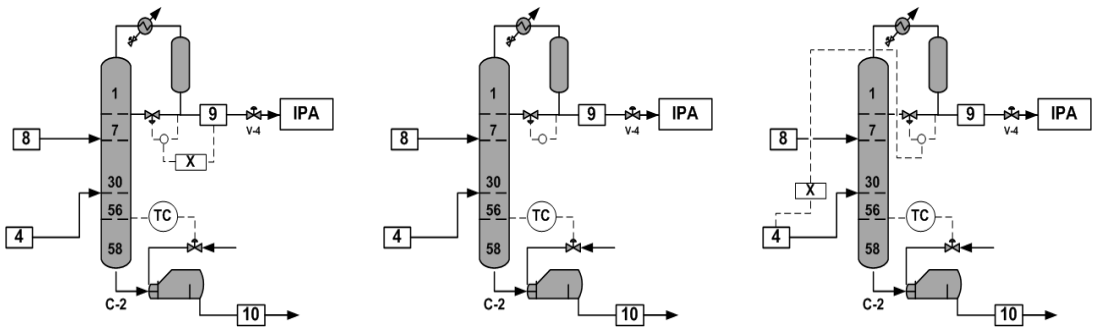
From temperature and total flow change in Column C1 (Fig. 5.20 and 5.21) reflux to feed ratio can keep composition of IPA in distillate product constant. So configuration of Column C1 is reflux to feed ratio.

Material disturbance management in Column C2: disturbance of Column C2 is water. The distillate product of Column C1 (IPA and water) is fed to Column C2. An extractive solvent ethylene glycol is fed on the top of the column. Distillate product is 99.91 wt% IPA. So the configuration for material disturbances management of Column C2 must force the disturbance to bottom product.

The alternate control schemes of Column C2 are shown in Fig.5.16. Blue line represents reflux ratio constant, red line represents reflux flow constant and black line represents reflux to feed ratio.

From component flow and composition change in Column C2 (Fig. 5.22 and 5.23) all structure can force the disturbance to bottom product.

From temperature and total flow change in Column C2 (Fig. 5.24 and 5.25) reflux to feed ratio can keep composition of IPA in distillate product constant. So configuration of Column C2 is reflux to feed ratio.

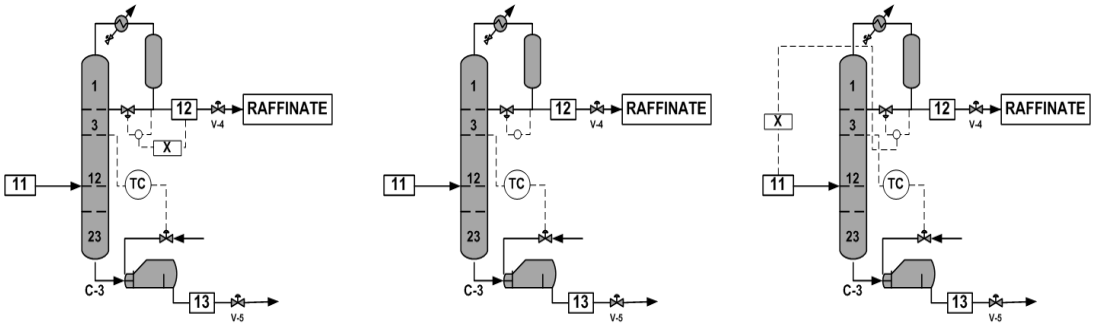


(a) (b) (c)

Figure 5.16 Alternate control schemes of Column C1. (a) Reflux ratio constant (b) Reflux flow constant (c) reflux to feed ratio

Material disturbance management in Column C3: disturbance of Column C3 is water. The bottom product of Column C2 (water and ethylene glycol) is fed to Column C3. Bottom product is 99.99 wt% ethylene glycol. So the configuration for material disturbances management of Column C3 must force the disturbance to distillate product.

The alternate control schemes of Column C3 are shown in Fig.5.17. Blue line represents reflux ratio constant, red line represents reflux flow constant and black line represents reflux to feed ratio.



(a) (b) (c)

Figure 5.17 Alternate control schemes of Column C1. (a) Reflux ratio constant (b) Reflux flow constant (c) reflux to feed ratio

From component flow and composition change in Column C3 (Fig. 5.26 and 5.27) all structure can force the disturbance to bottom product.

From temperature and total flow change in Column C3 (Fig. 5.28 and 5.29) reflux ratio constant can keep composition of water in distillate product and composition of ethylene glycol in bottom product constant. So configuration of Column C3 is reflux ratio constant.

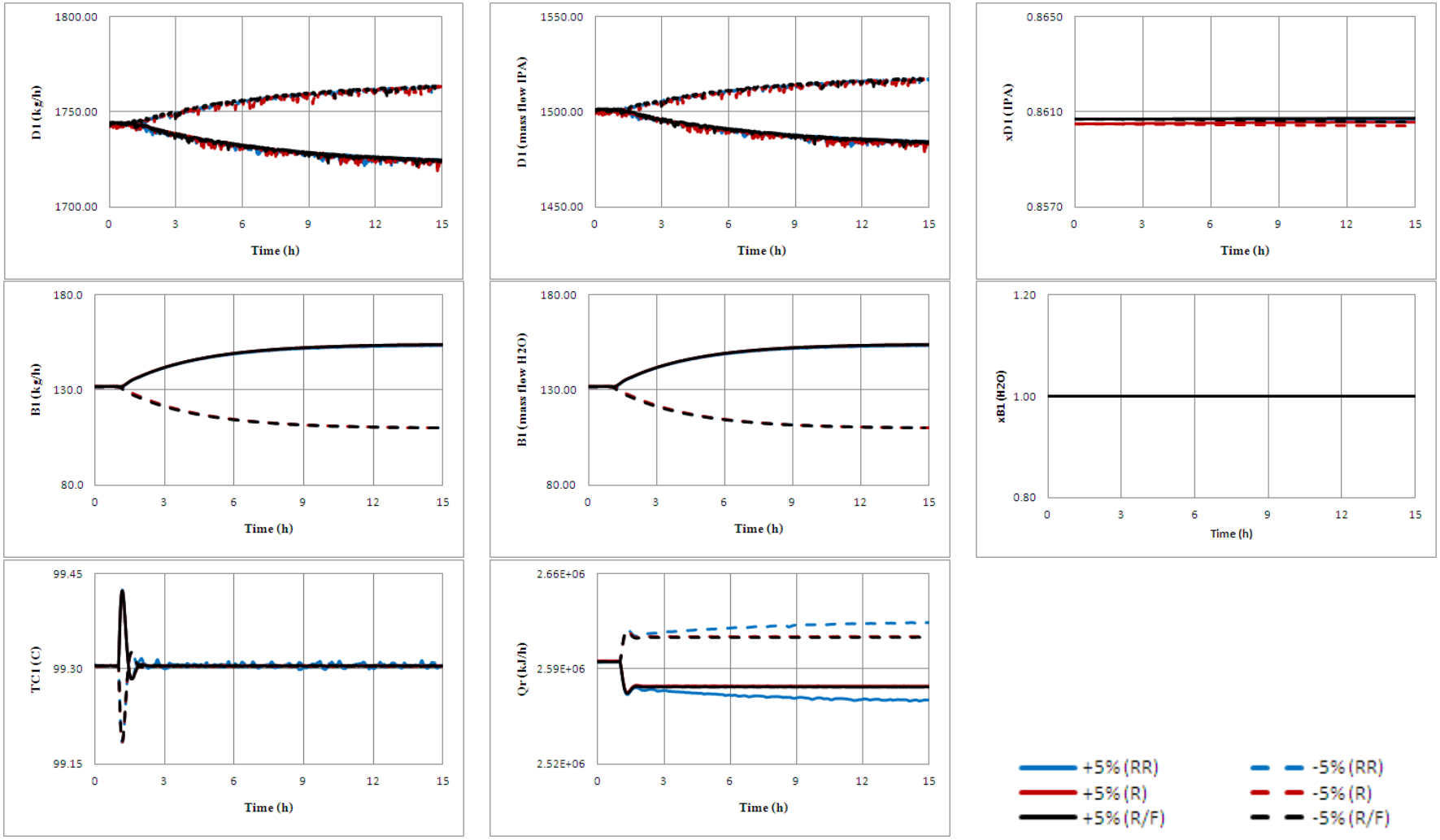


Figure 5.18 Result of component flow change in Column C1

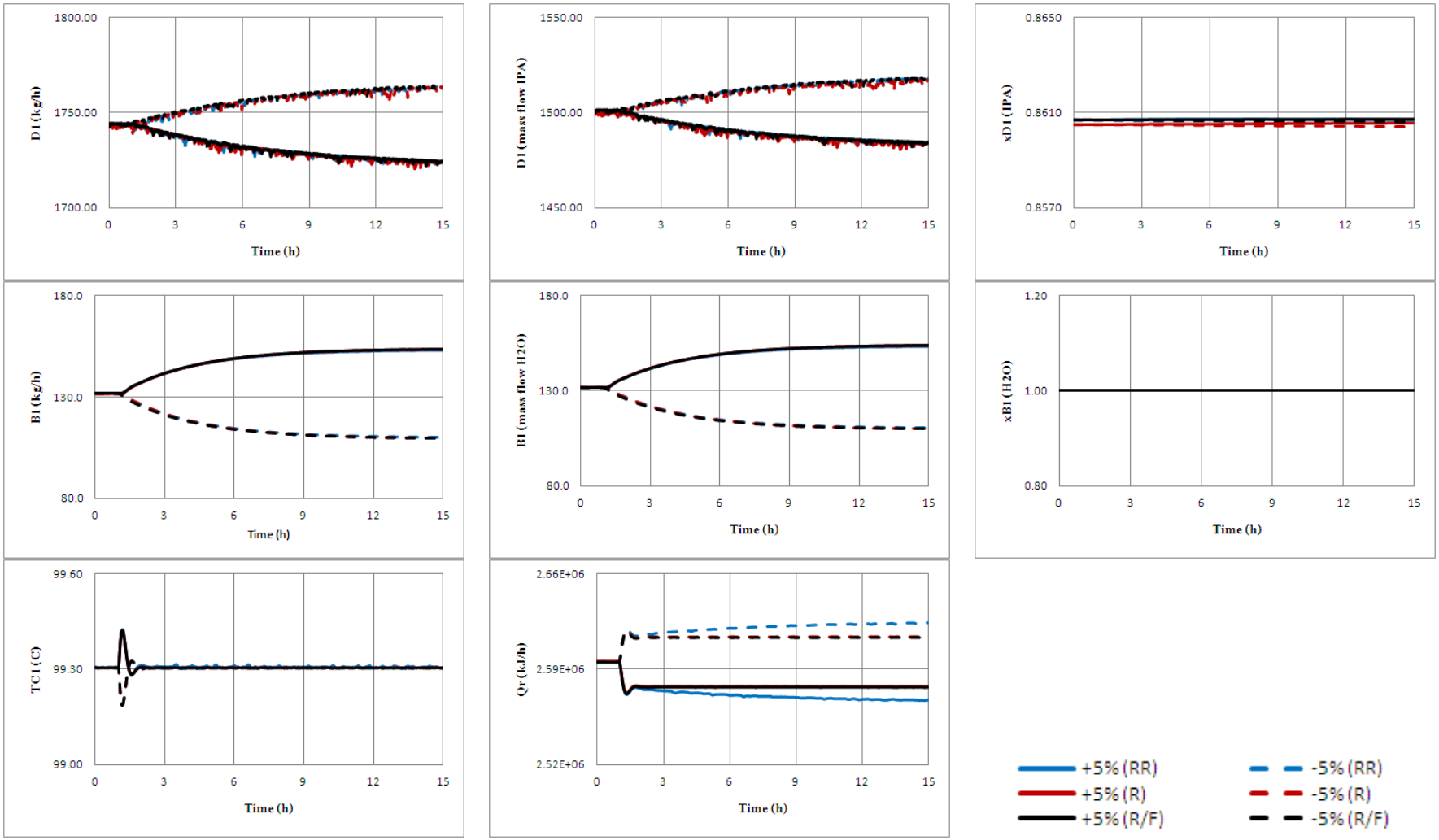


Figure 5.19 Result of composition change in Column C1

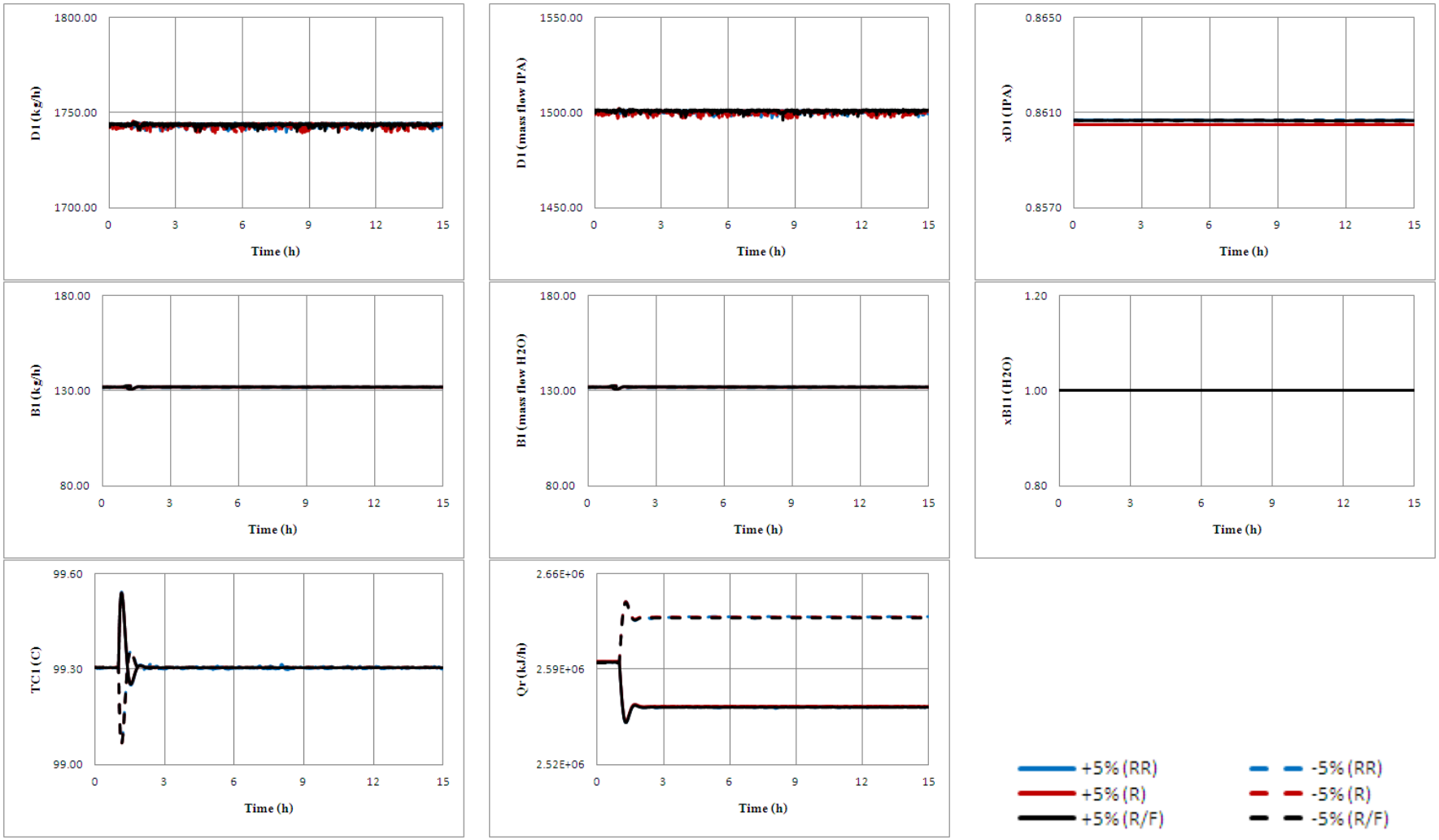


Figure 5.20 Result of temperature change in Column C1

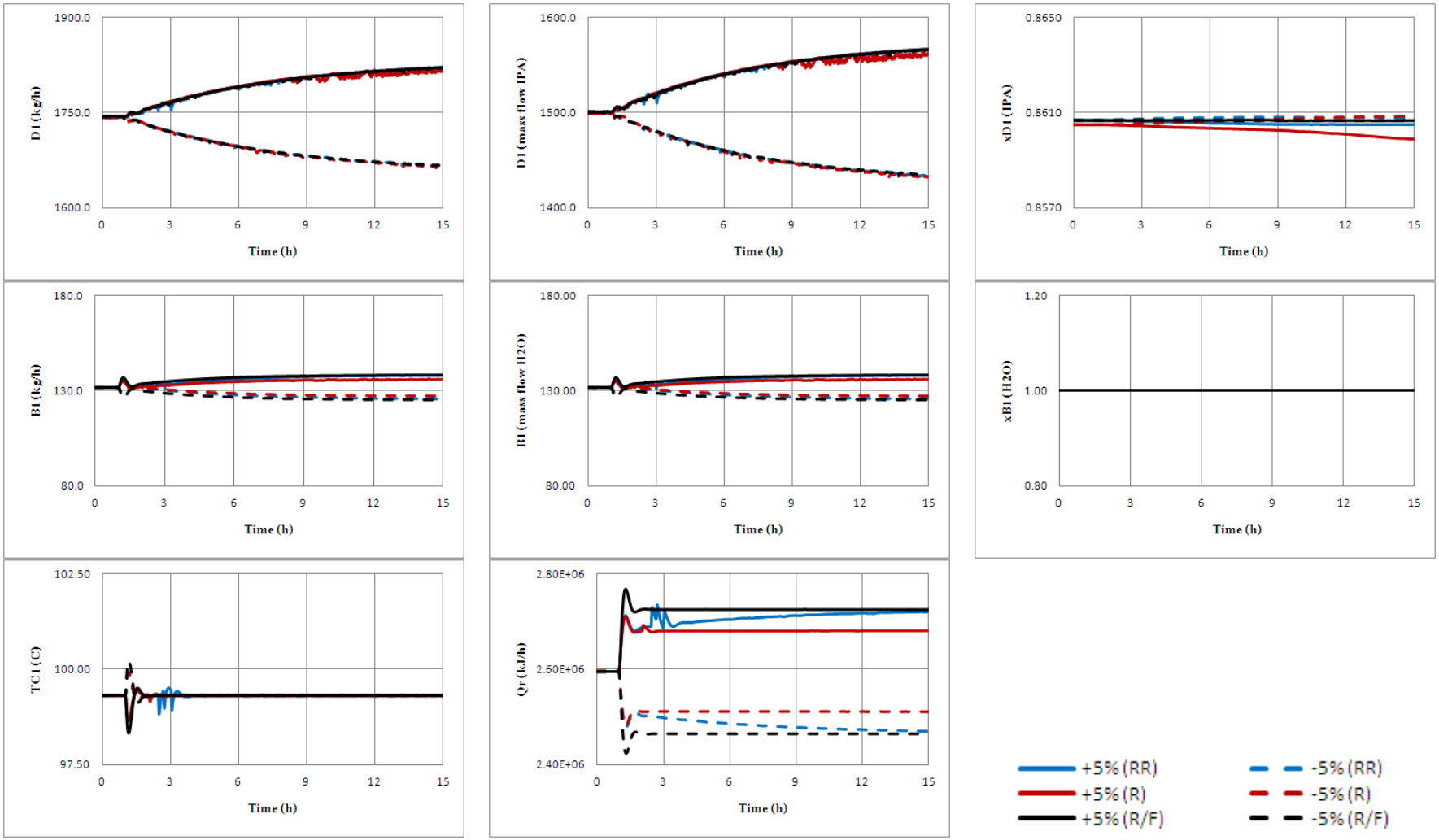


Figure 5.21 Result of total flow change in Column C1

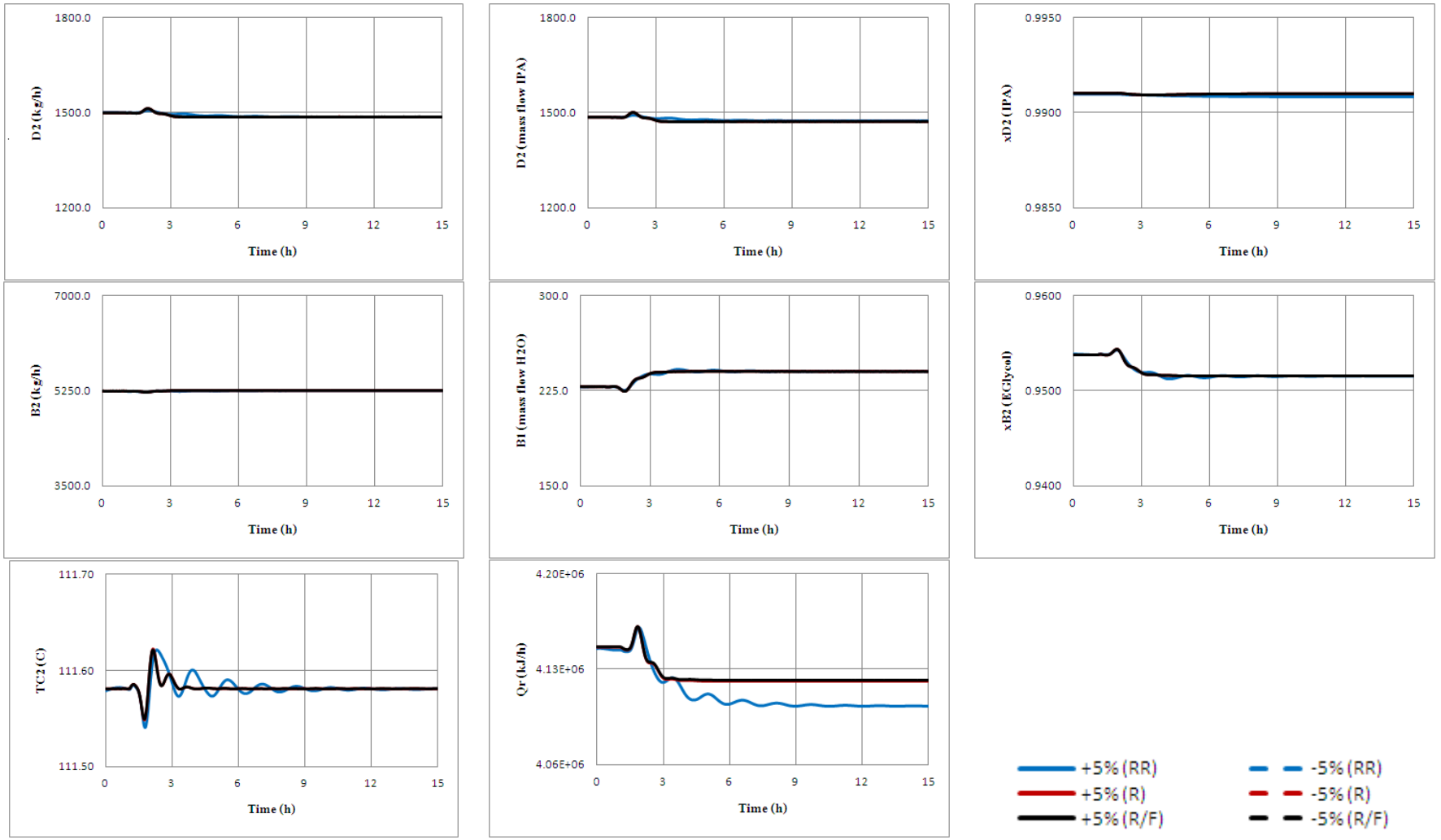


Figure 5.22 Result of component flow change in Column C2

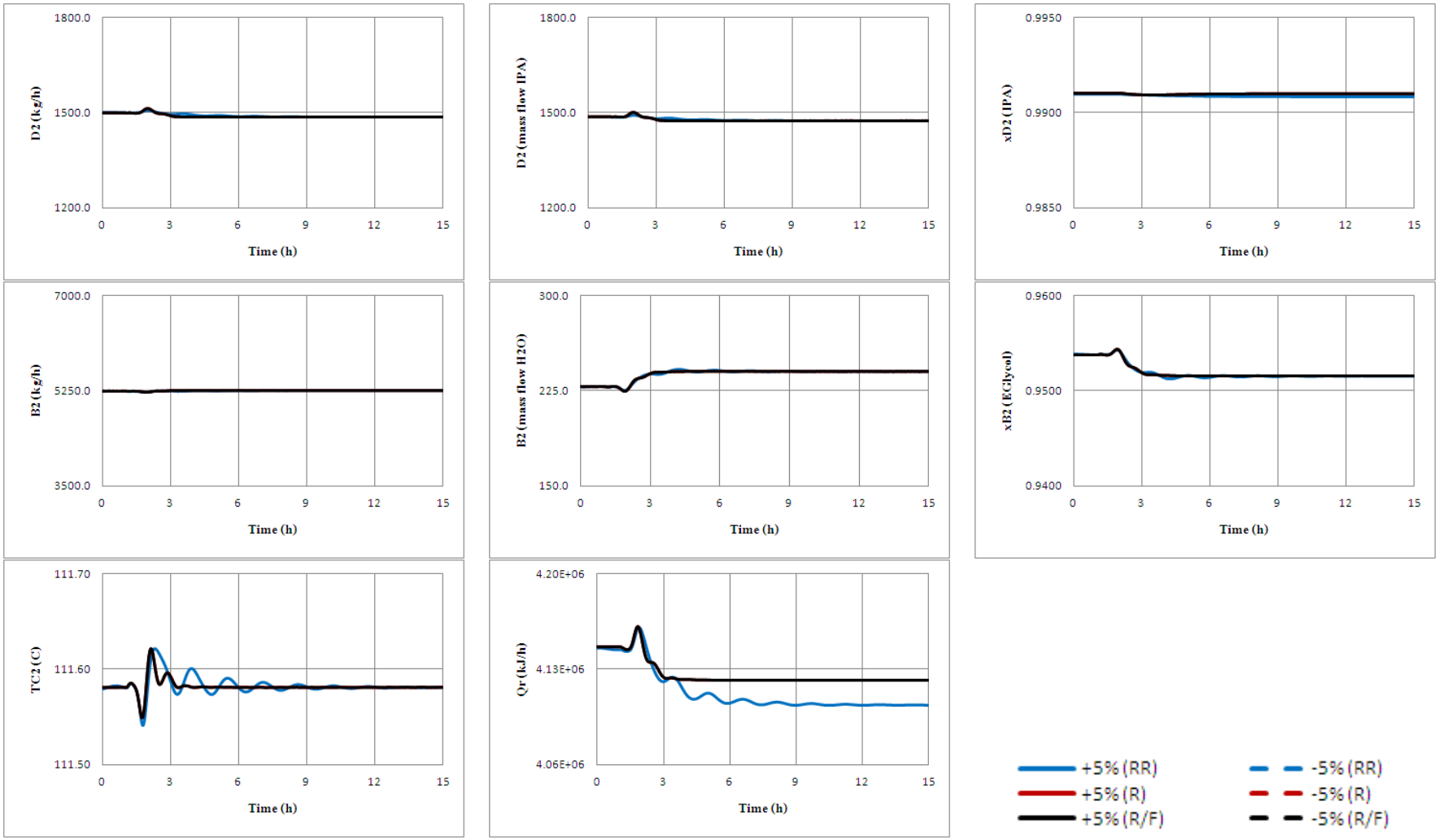


Figure 5.23 Result of composition change in Column C2

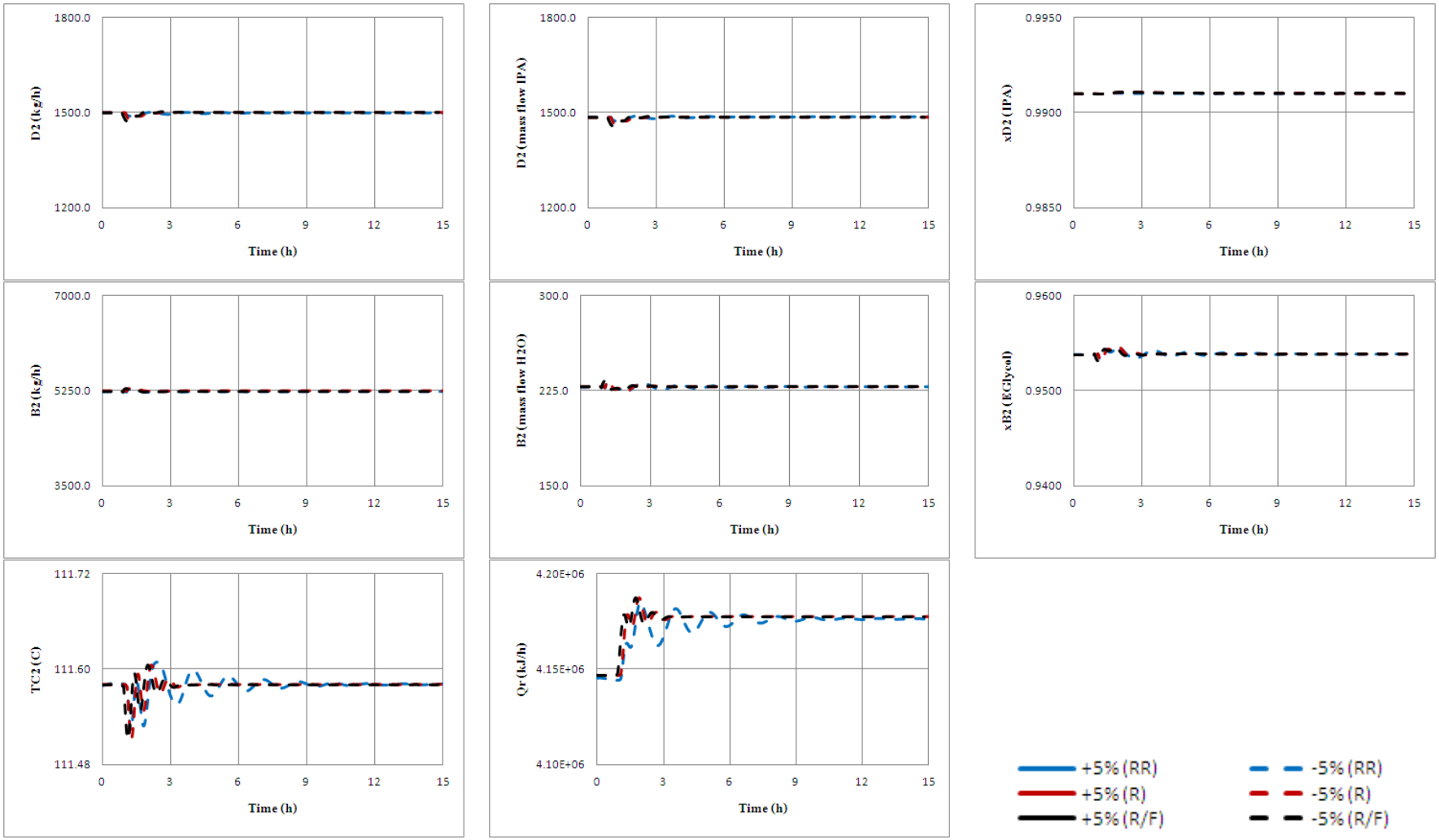


Figure 5.24 Result of temperature change in Column C2

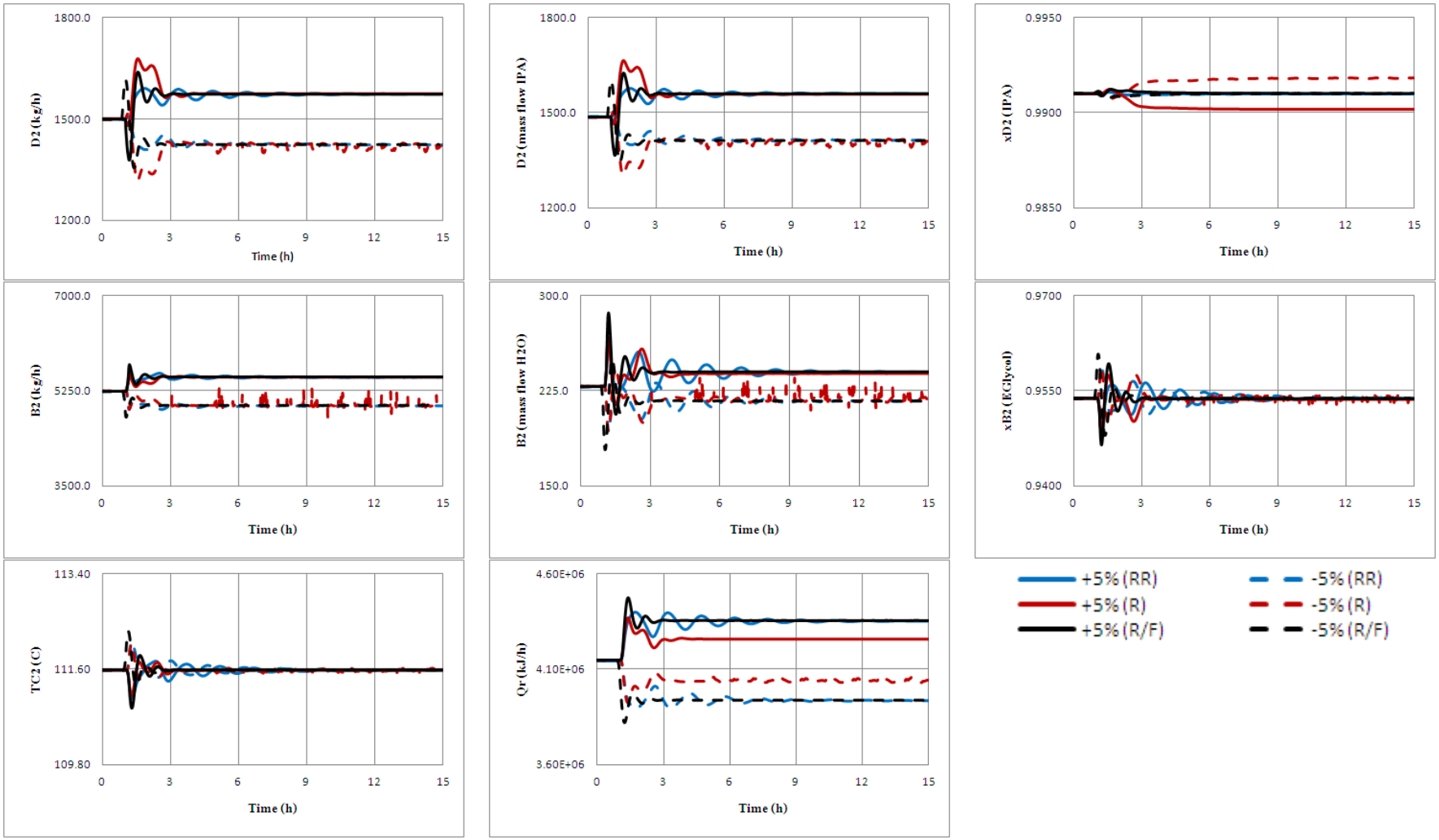


Figure 5.25 Result of total flow change in Column C2

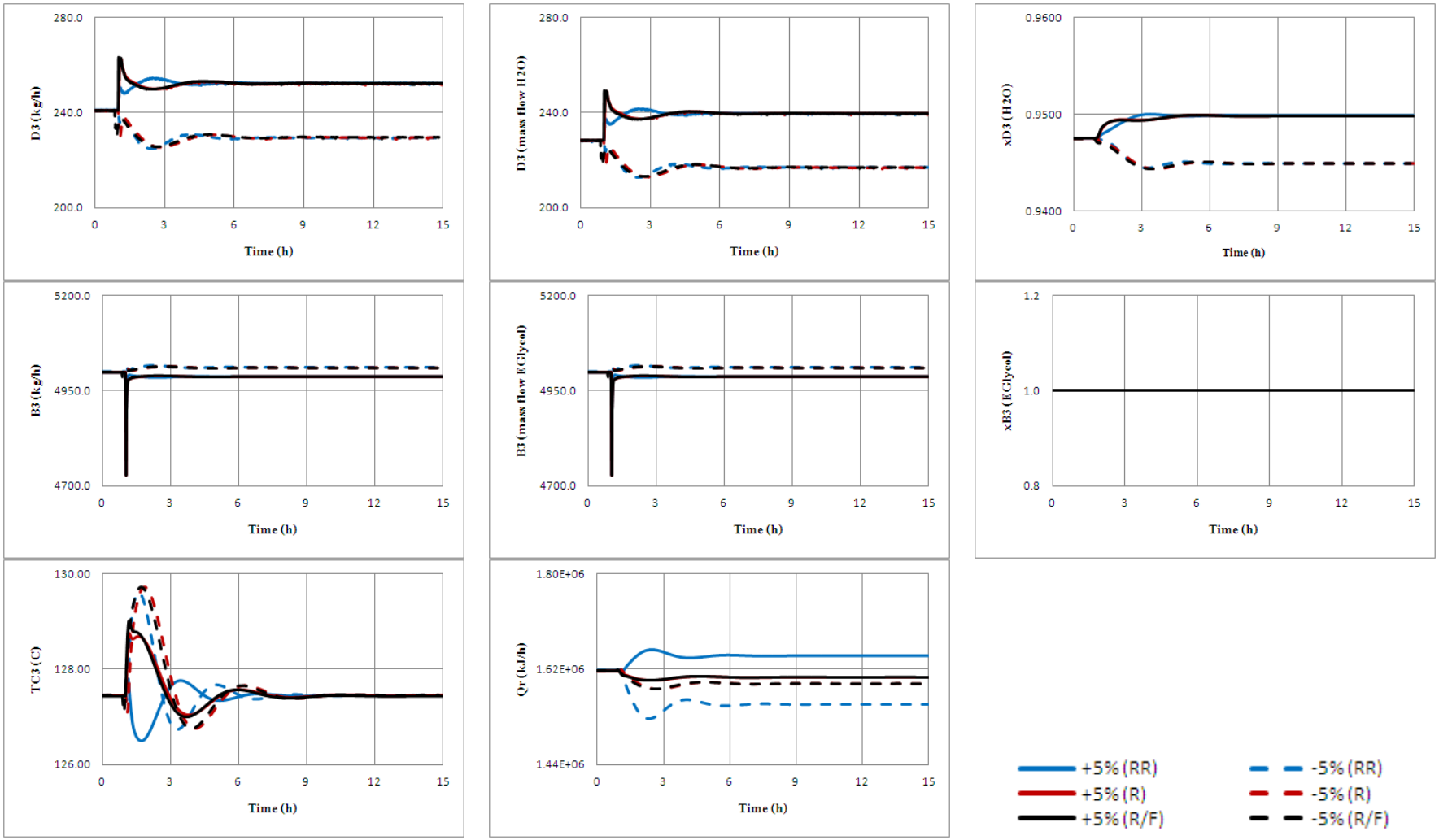


Figure 5.26 Result of component flow change in Column C3

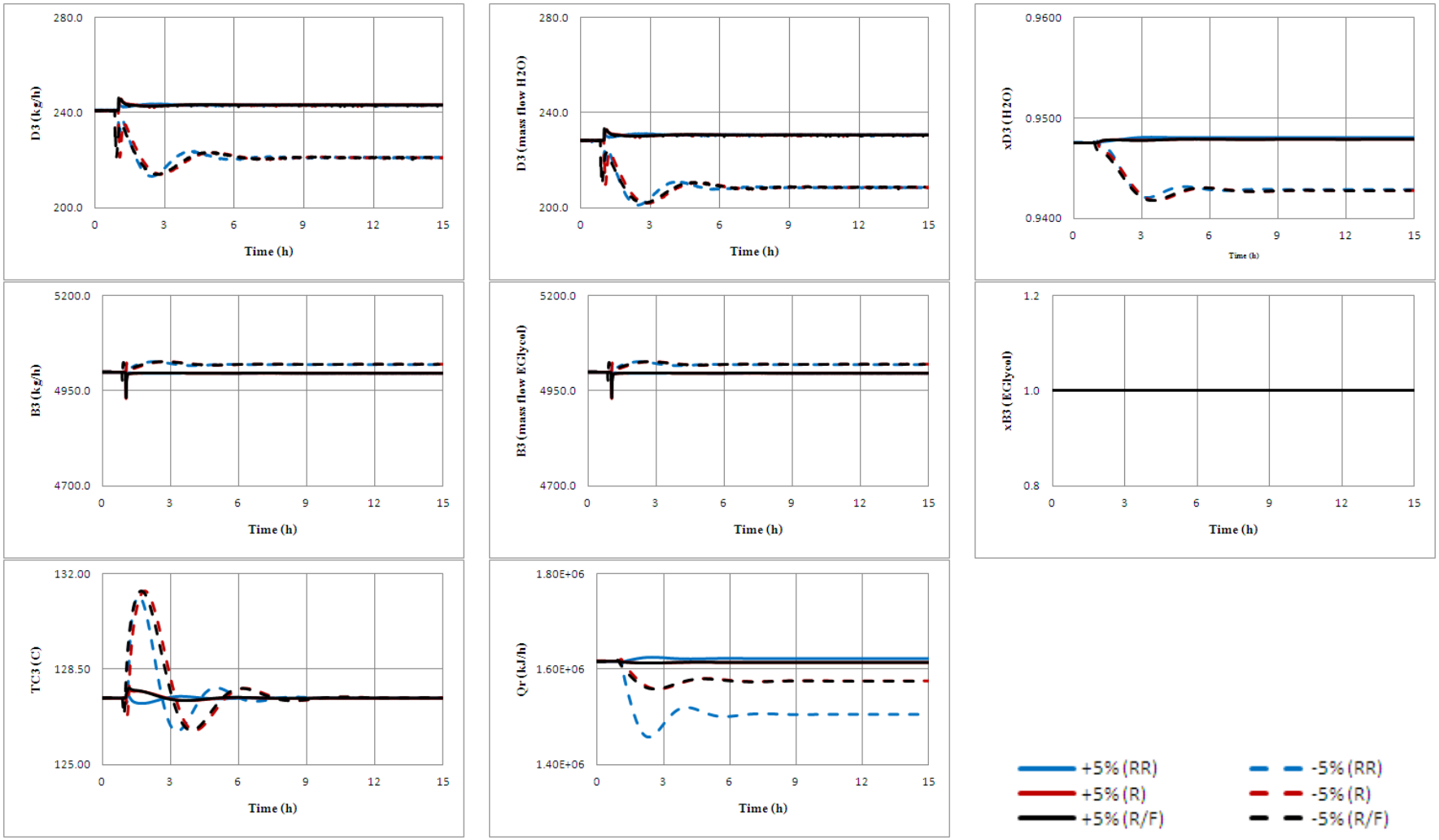


Figure 5.27 Result of composition change in Column C3

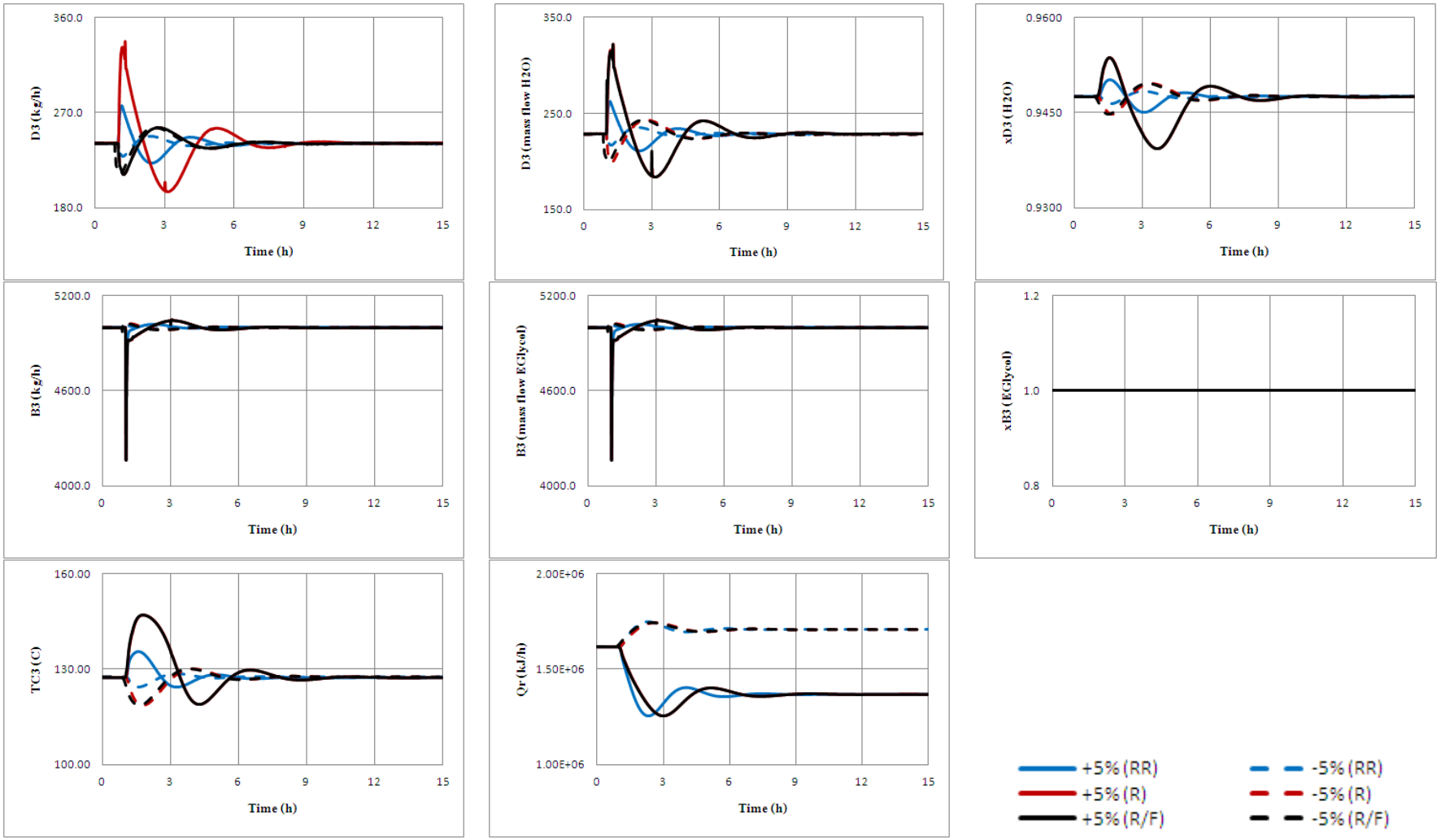


Figure 5.28 Result of temperature change in Column C3

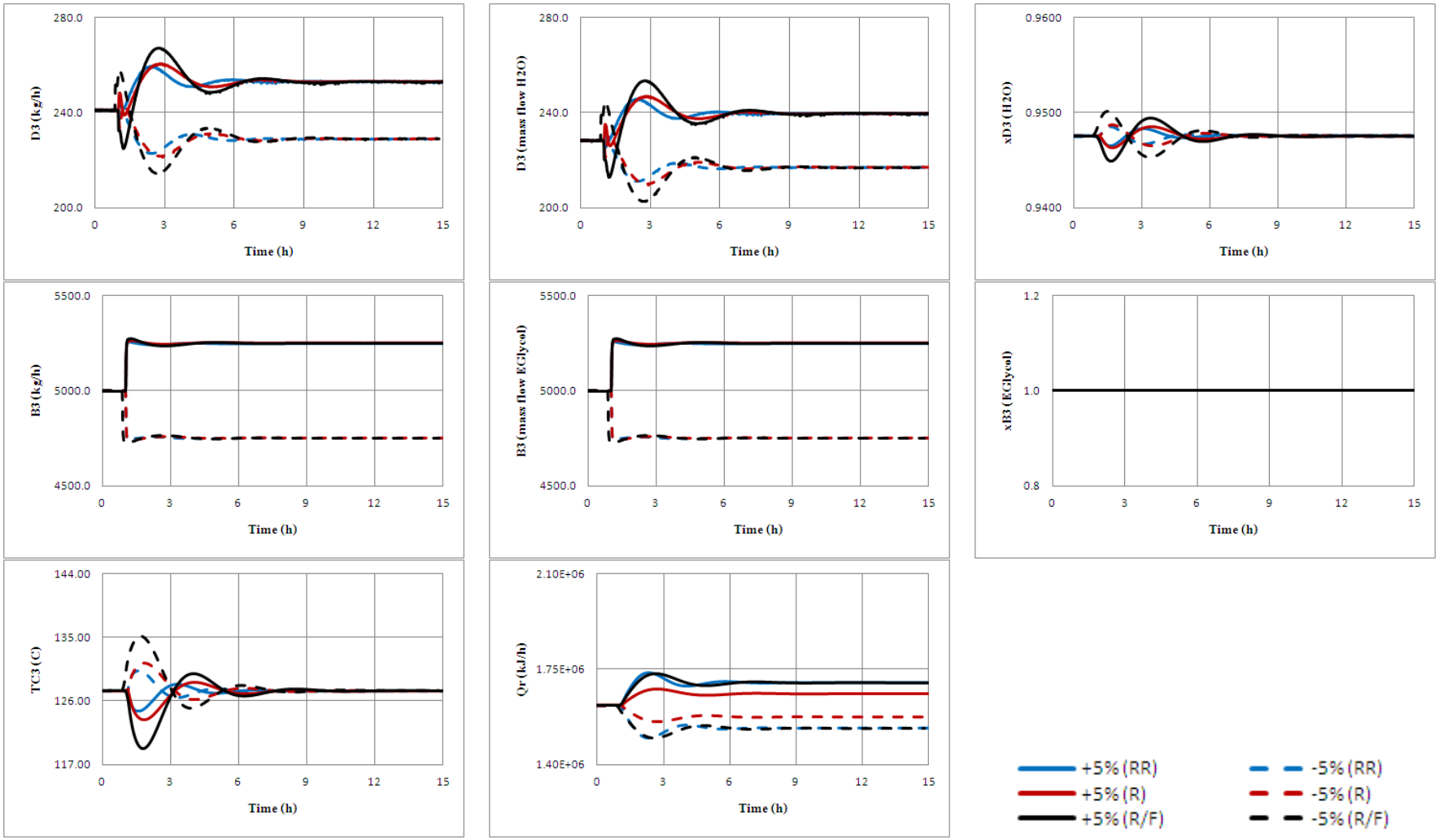


Figure 5.29 Result of total flow change in Column C3

The control structure of water pathways is shown in Fig.5.30. The control structure of IPA pathways is presented in Fig.5.31. Finally, the control structure of ethylene glycol pathways is illustrated in Fig.5.32.

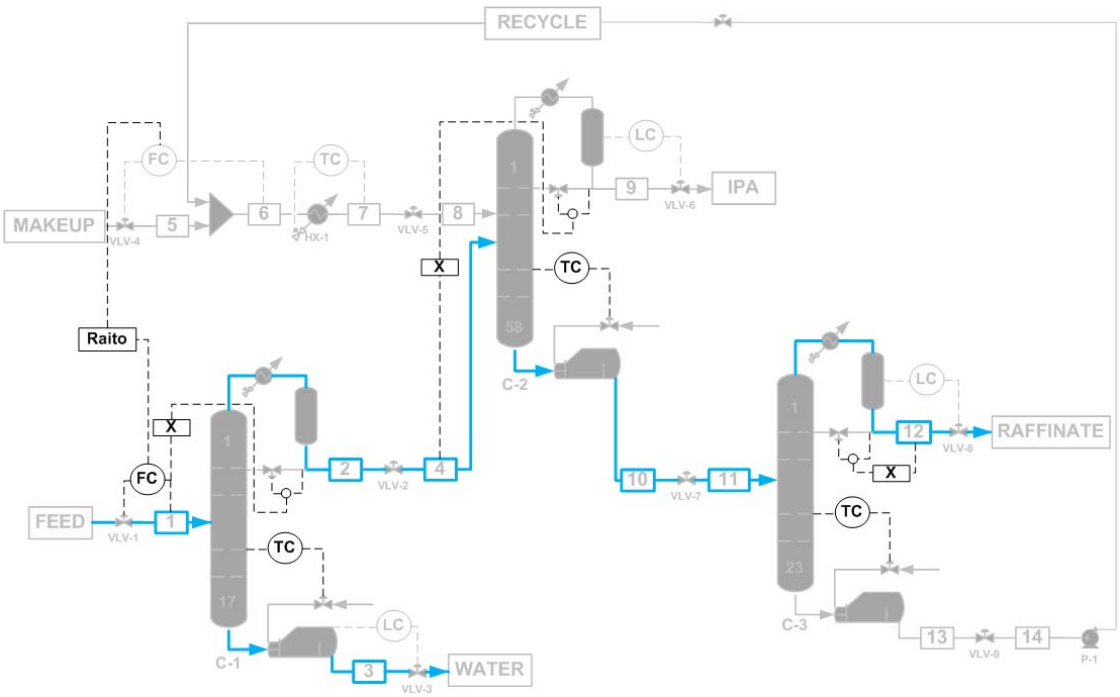


Figure 5.30 Control structure of water pathway

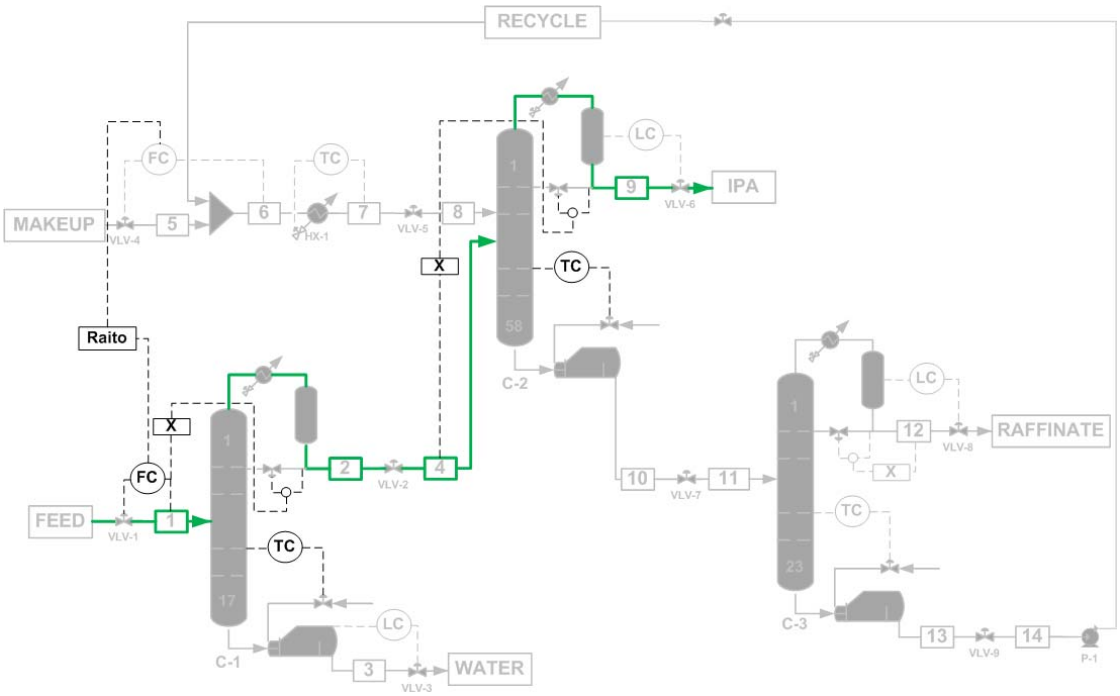


Figure 5.31 Control structure of IPA pathway

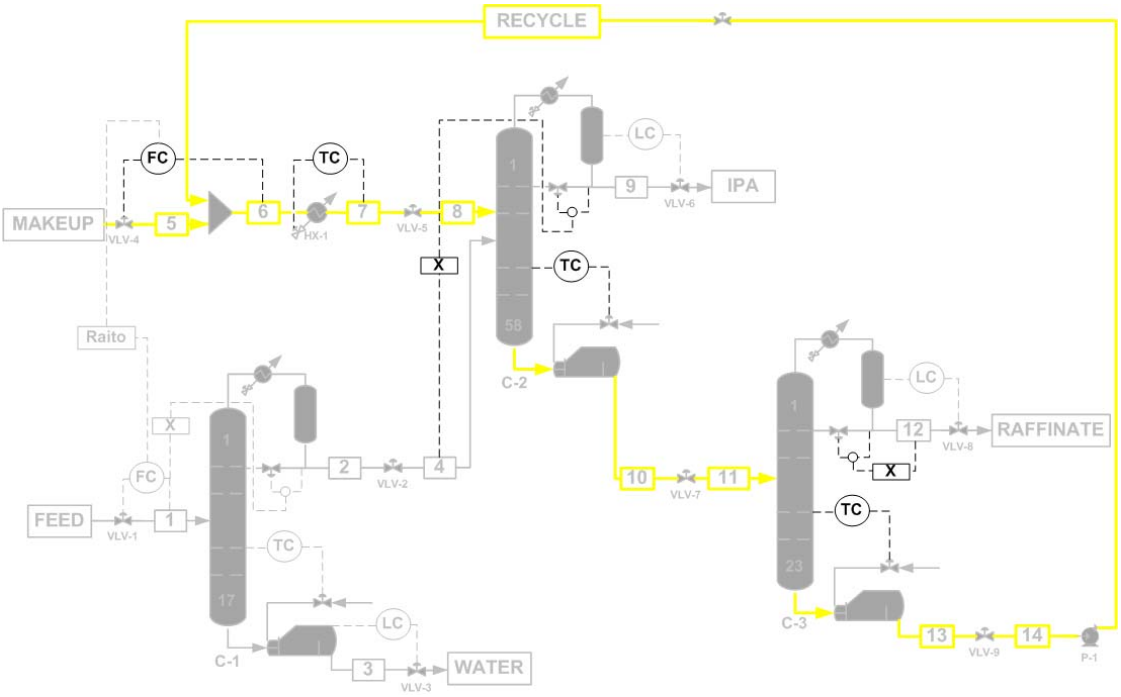


Figure 5.32 Control structure of ethylene glycol pathway

Step 5: Design the rest of the control loops. The control loops for the remaining control variables are shown in Fig.5.33.

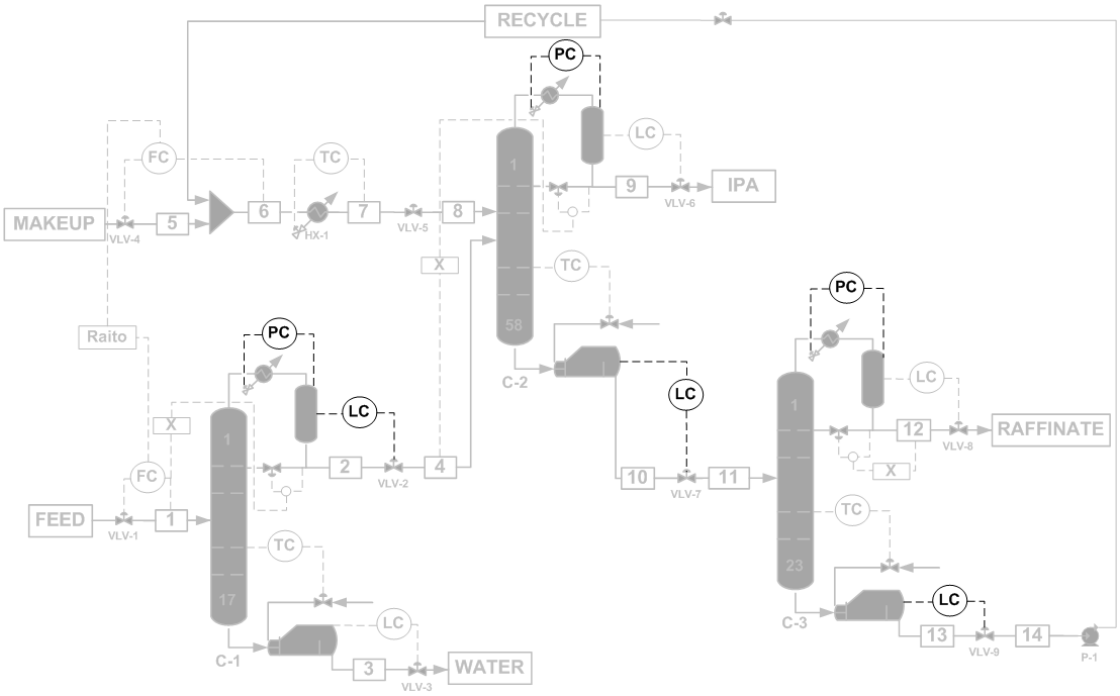


Figure 5.33 Control loops for remaining control variables

Step 6: Energy management via heat exchanger networks. In this process there are no process-to-process heat exchanger and no heat integration. Disturbance can then be rejected to the plant utilities system.

Step 7: Optimize economics and/or improve control performance. We have now established the basic regulatory plantwide control strategy. We have constructed a rigorous dynamic simulation of isopropyl alcohol dehydration process

Control structure of base case.

Column C1:

1. Feed is flow controlled.
2. Stage 11 temperature is controlled by reboiler heat input.
3. Pressure is controlled by condenser heat removal.
4. Reflux ratio is held constant.
5. Reflux drum level is controlled by distillate flow.
6. Base level is controlled by bottom flow.

Column C2:

1. The extract solvent flow is flow controlled.
2. The temperature of the solvent is controlled by manipulating the heat removal in the cooler.
3. Stage 56 temperature is controlled by reboiler heat input.
4. Pressure control by condenser heat removal.
5. Reflux ratio is held constant.
6. Reflux drum level is controlled by distillate flow.
7. Base level is controlled by bottom flow.

Column C3:

1. Stage 3 temperature is controlled by reboiler heat input.
2. Pressure control by condenser heat removal.
3. Reflux ratio is held constant.
4. Reflux drum level is controlled by distillate flow.
5. Base level is controlled by flow rate of the makeup ethylene glycol.

Base case control structure is show in Fig 5.34 and control structure list are show in Table 5.2 and 5.3.

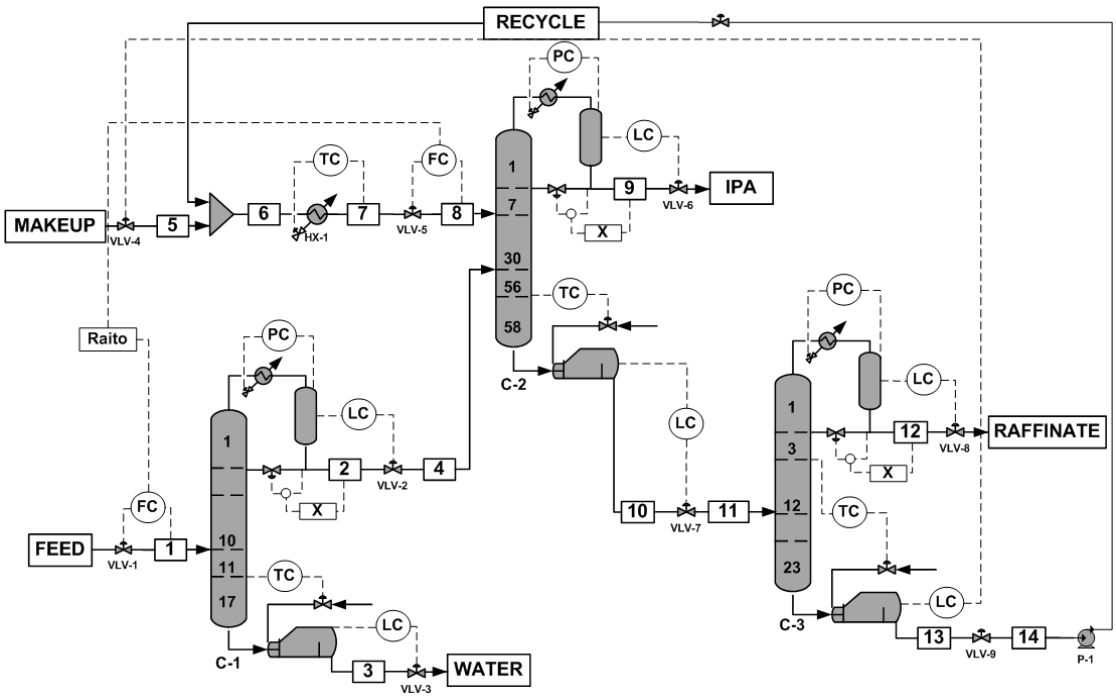


Figure 5.34 Control structure of base case

Table 5.2 Control structure list of base case

Equipment	Controller	Controlled variables	Manipulated variables	Type	Action
Feed	FC1	Mass flow rate	Feed flow rate	PI	Reverse
EGlycol feed	FC2	Total flow rate	EGlycol flow rate	PI	Reverse
	RATO1	Mass flow rate	Set point of FC2	PI	Reverse
	TC4	EGlycol temperature	Cooler heat removal	PI	Direct
Column C1	TC1	Stage 11 temperature	Reboiler heat input	PI	Reverse
	PC1	Condenser pressure	Condenser heat removal	PI	Direct
	LC1	Reflux drum level	Distillate flow	P	Direct
	LC2	Base level	Bottom flow	P	Direct
	RATO11	Reflux flow rate	Set point of reflux flow control	PI	Reverse

Table 5.3 Control structure list of base case (Continue)

Equipment	Controller	Controlled variables	Manipulated variables	Type	Action
Column C2	TC2	Stage 56 temperature	Reboiler heat input	PI	Reverse
	PC2	Condenser pressure	Condenser heat removal	PI	Direct
	LC3	Reflux drum level	Distillate flow	P	Direct
	LC4	Base level	Bottom flow	P	Direct
	RATO21	Reflux flow rate	Set point of reflux flow control	PI	Reverse
Column C3	TC3	Stage 3 temperature	Reboiler heat input	PI	Reverse
	PC3	Condenser pressure	Condenser heat removal	PI	Direct
	LC5	Reflux drum level	Distillate flow	P	Direct
	LC6	Base level	Makeup flow rate	P	Reverse
	RATO31	Reflux flow rate	Set point of reflux flow control	PI	Reverse

New design control structure.

Column C1:

1. Feed is flow controlled.
2. Stage 11 temperature is controlled by reboiler heat input.
3. Pressure is controlled by condenser heat removal.
4. Reflux to feed ratio is held constant.
5. Reflux drum level is controlled by distillate flow.
6. Base level is controlled by bottom flow.

Column C2:

1. The total extract solvent flow is controlled by the flow rate of makeup ethylene glycol
2. The temperature of the solvent is controlled by manipulating the heat removal in the cooler.

3. Stage 56 temperature is controlled by reboiler heat input.
4. Pressure control by condenser heat removal.
5. Reflux to feed ratio is held constant.
6. Reflux drum level is controlled by distillate flow.
7. Base level is controlled by bottom flow.

Column C3:

1. Stage 3 temperature is controlled by reboiler heat input.
2. Pressure control by condenser heat removal.
3. Reflux ratio is held constant.
4. Reflux drum level is controlled by distillate flow.
8. Base level is controlled by bottom flow.

New design control structure is show in Fig 5.35 and control structure list are show in Table 5.4 and 5.5.

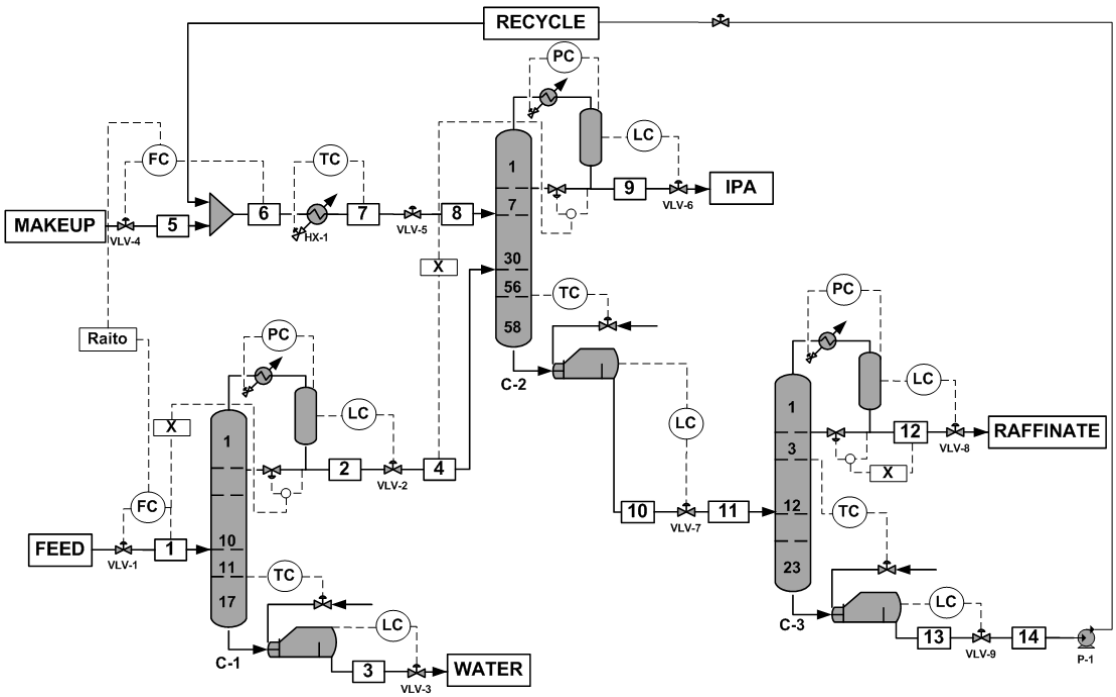


Figure 5.35 Control structure of new design

Table 5.4 Control structure list of new design.

Equipment	Controller	Controlled variables	Manipulated variables	Type	Action
Feed	FC1	Mass flow rate	Feed flow rate	PI	Reverse
EGlycol feed	FC2	Total flow rate	Makeup flow rate	PI	Reverse
	RATO1	Mass flow rate	Set point of FC2	PI	Reverse
	TC4	EGlycol Temperature	Cooler heat removal	PI	Direct
Column C1	TC1	Stage 11 temperature	Reboiler heat input	PI	Reverse
	PC1	Condenser pressure	Condenser heat removal	PI	Direct
	LC1	Reflux drum level	Distillate flow	P	Direct
	LC2	Base level	Bottom flow	P	Direct
	RATO11	Reflux flow rate	Set point of reflux flow control	PI	Reverse

Table 5.5 Control structure list of new design (Continue)

Equipment	Controller	Controlled variables	Manipulated variables	Type	Action
Column C2	TC2	Stage 56 temperature	Reboiler heat input	PI	Reverse
	PC2	Condenser pressure	Condenser heat removal	PI	Direct
	LC3	Reflux drum level	Distillate flow	P	Direct
	LC4	Base level	Bottom flow	P	Direct
	RATO21	Reflux flow rate	Set point of reflux flow control	PI	Reverse
Column C3	TC3	Stage 3 emperature	Reboiler heat input	PI	Reverse
	PC3	Condenser pressure	Condenser heat removal	PI	Direct
	LC5	Reflux drum level	Distillate flow	P	Direct
	LC6	Base level	Bottom flow	P	Direct
	RATO31	Reflux flow rate	Set point of reflux flow control	PI	Reverse

Step 8: Validate the designed control structures by rigorous dynamic simulation. This step, the disturbances are feed flow rate and feed composition.

Table 5.6 Disturbances to the effect of dynamic simulations for base case and new design control structures.

No.	Description	Nominal	Disturbance
Dyn1	Feed rate (kg/h)	1,875 kg/h	± 5 %
Dyn2	Feed composition (H2O wt%)	20 wt%	± 5 %
Dyn3	Feed temperature (°C)	76.85 °C	± 5 °C

Note: that the disturbances are applied 1 hr after the beginning of each simulation run.

Table 5.7 Dynamic responses with feed rate change.

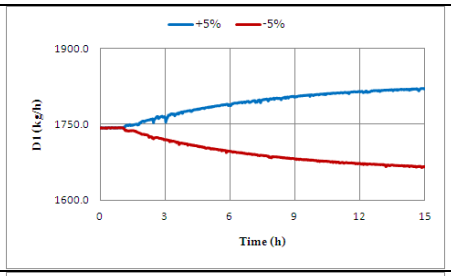
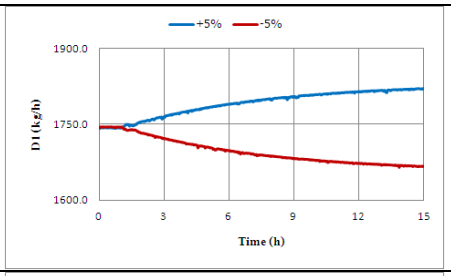
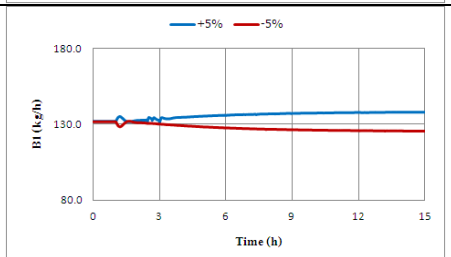
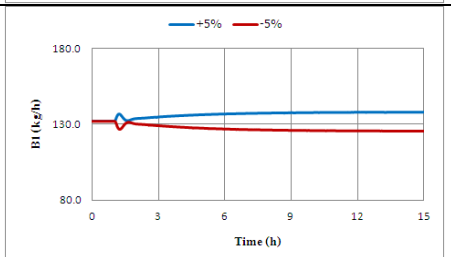
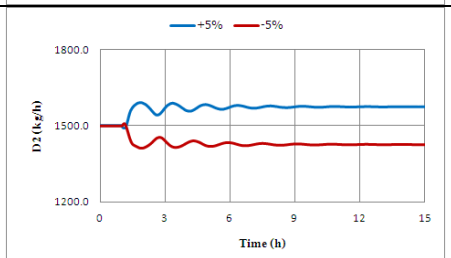
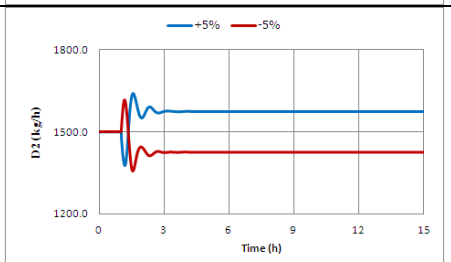
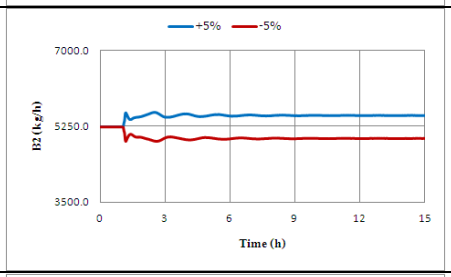
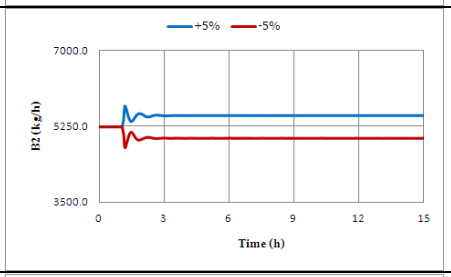
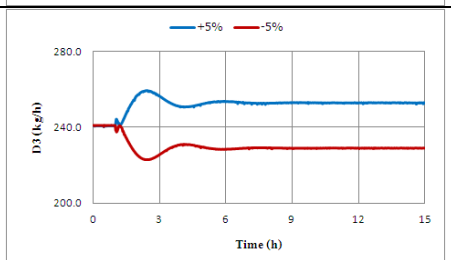
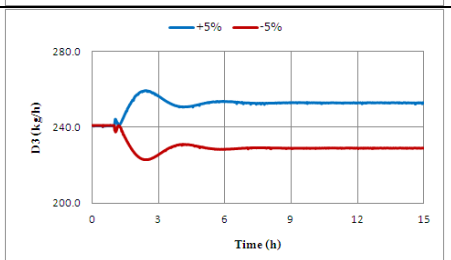
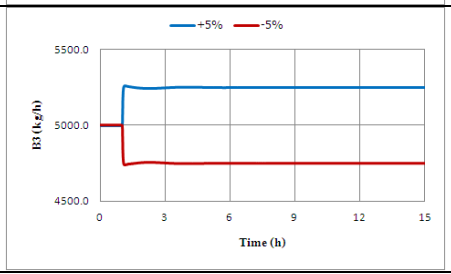
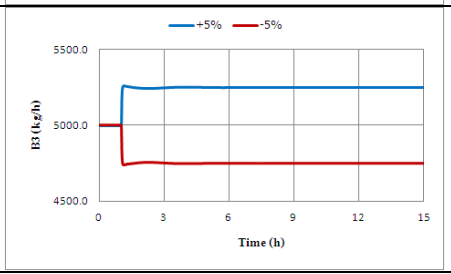
	BC	New design
a.		
b.		
c.		
d.		
e.		
f.		

Table 5.8 Dynamic responses with feed rate change. (Continue)

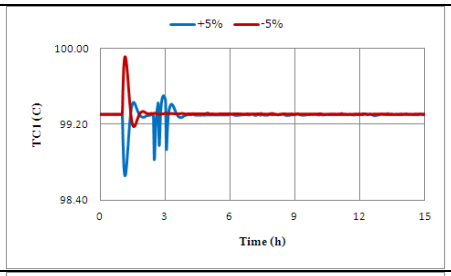
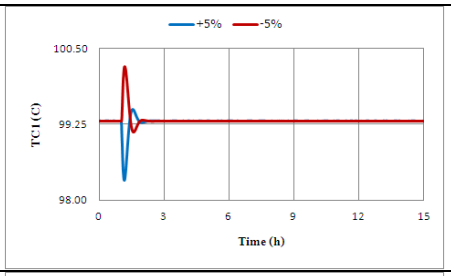
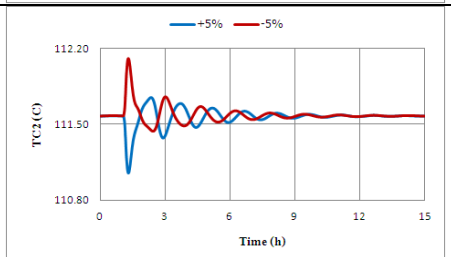
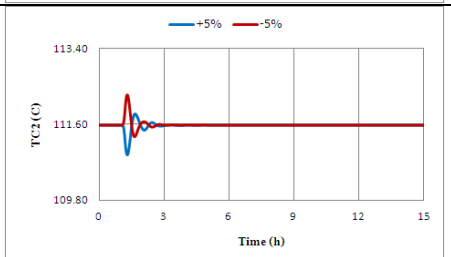
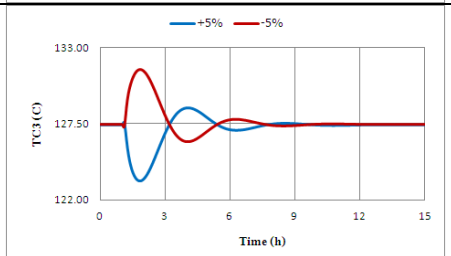
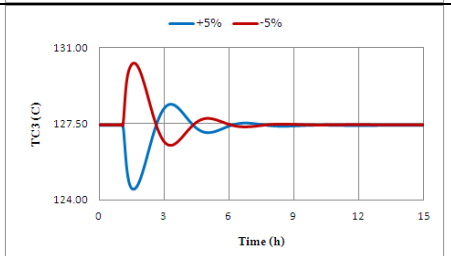
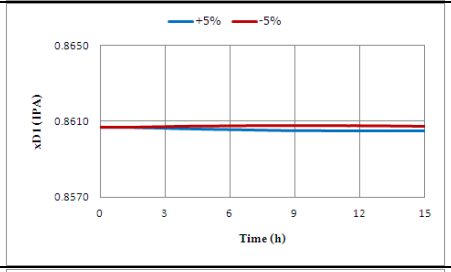
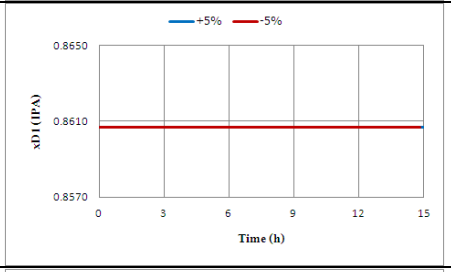
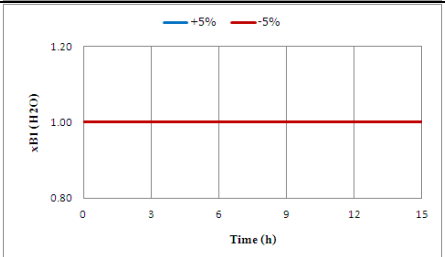
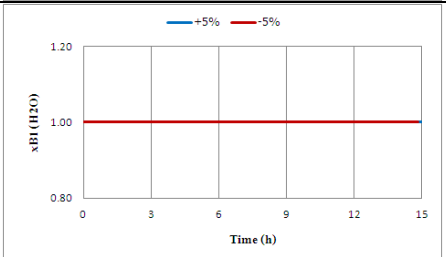
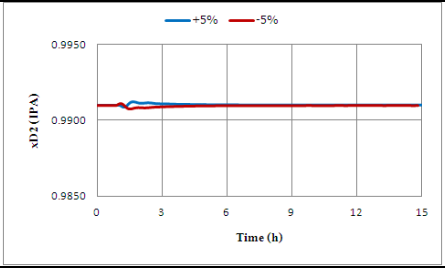
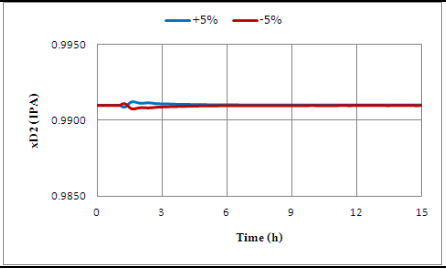
	BC	New design
g.		
h.		
i.		
j.		
k.		
l.		

Table 5.9 Dynamic responses with feed rate change. (Continue)

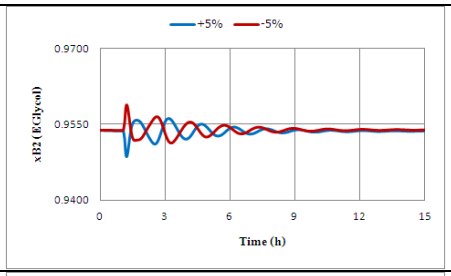
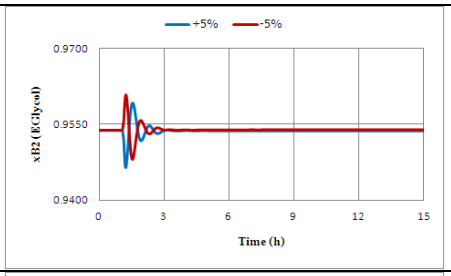
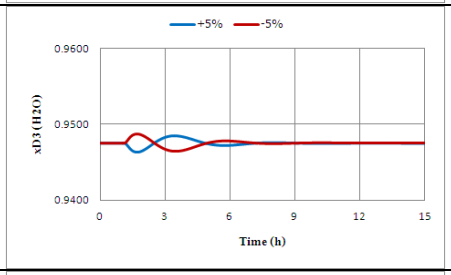
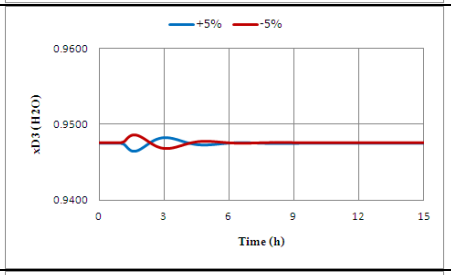
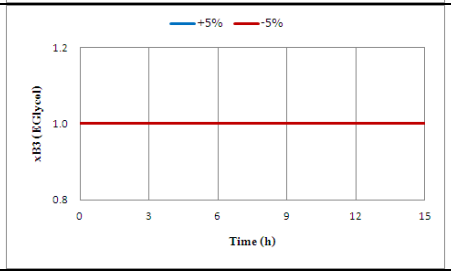
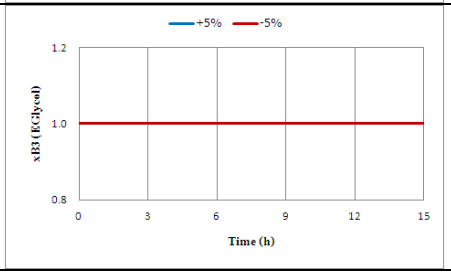
	BC	New design
m.		
n.		
o.		

Table 5.10 Dynamic responses with feed composition change.

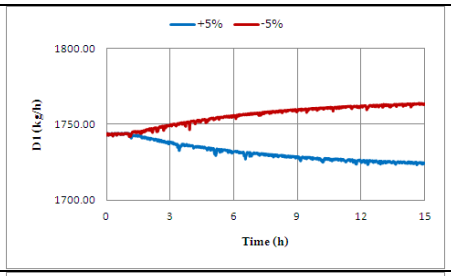
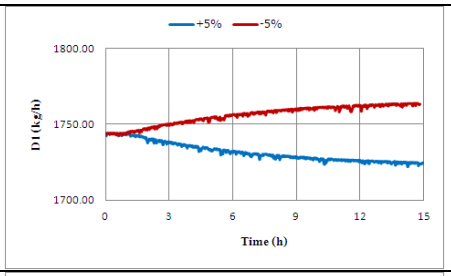
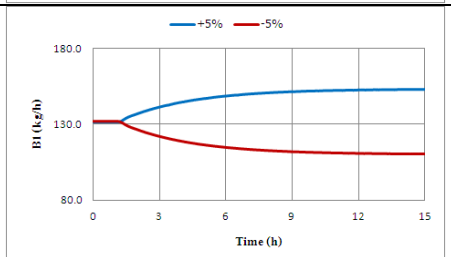
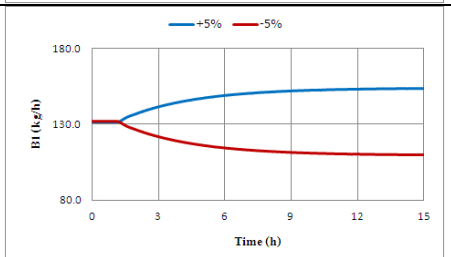
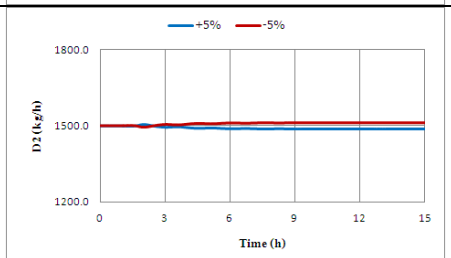
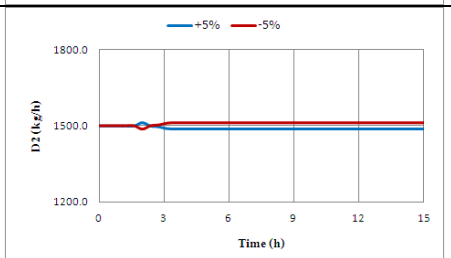
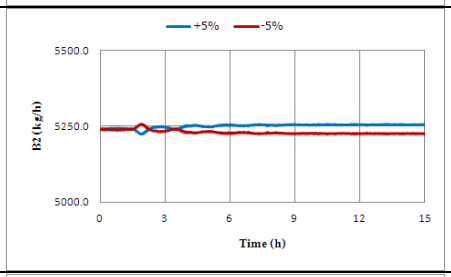
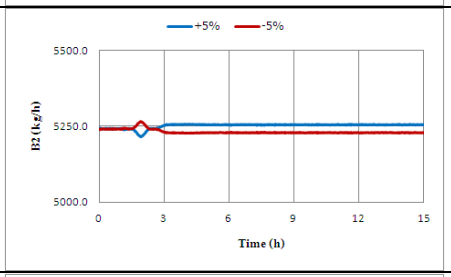
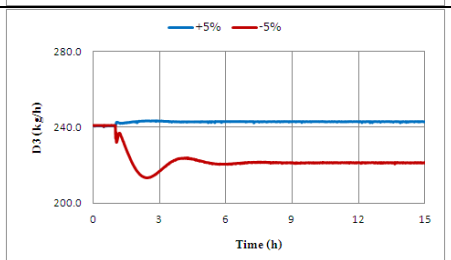
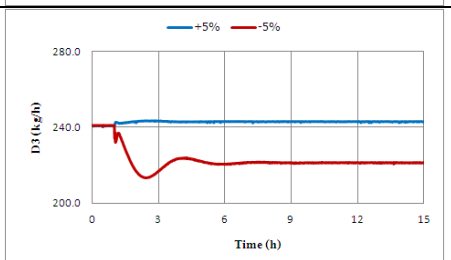
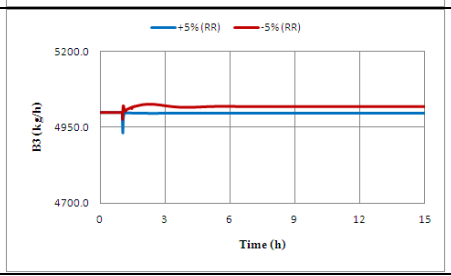
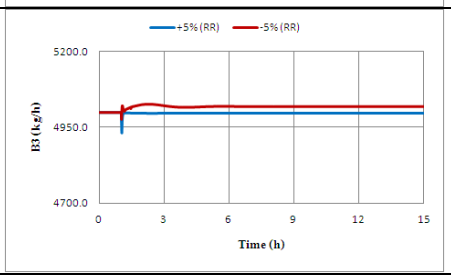
	BC	New design
a.		
b.		
c.		
d.		
e.		
f.		

Table 5.11 Dynamic responses with feed composition change. (Continue)

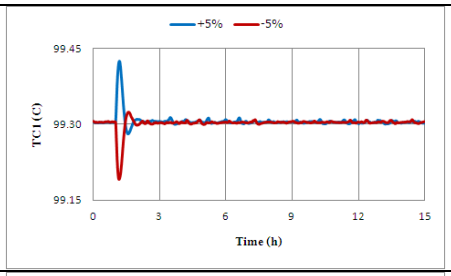
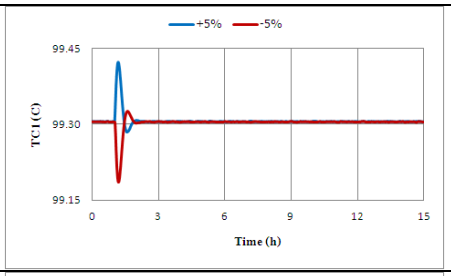
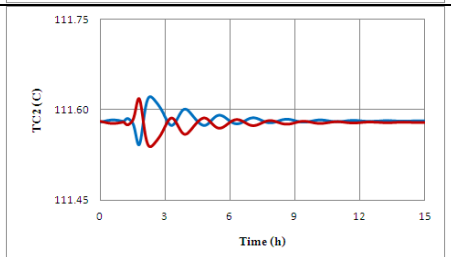
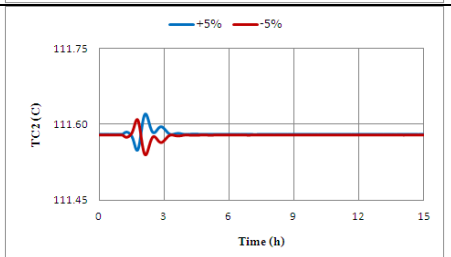
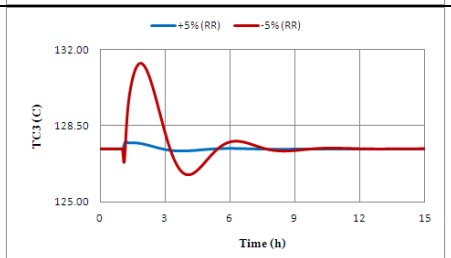
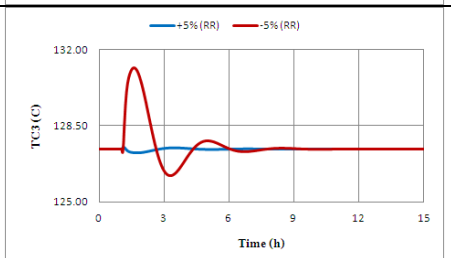
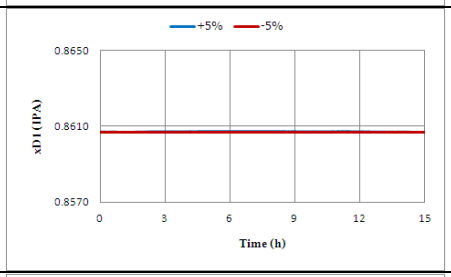
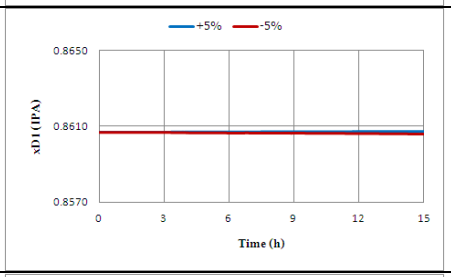
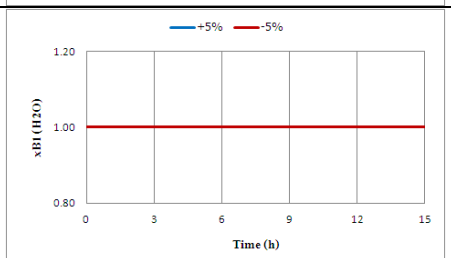
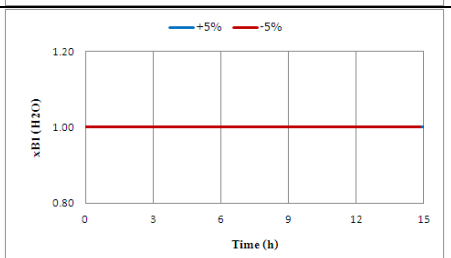
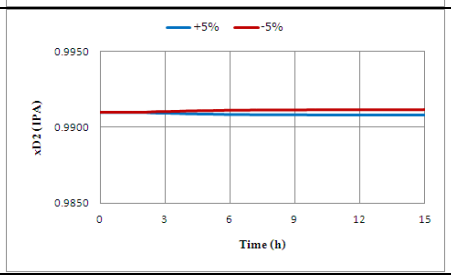
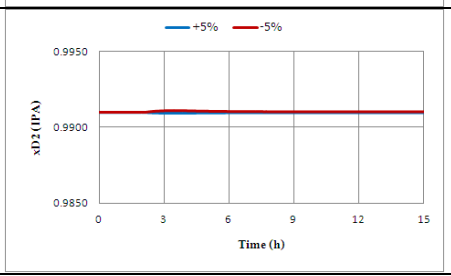
	BC	New design
g.		
h.		
i.		
j.		
k.		
l.		

Table 5.12 Dynamic responses with feed composition change. (Continue)

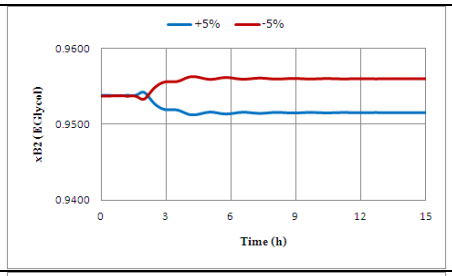
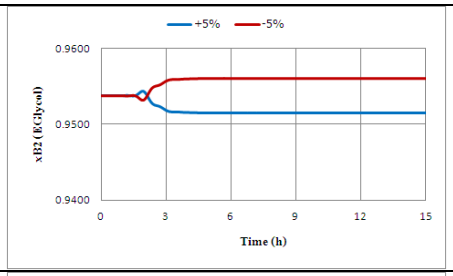
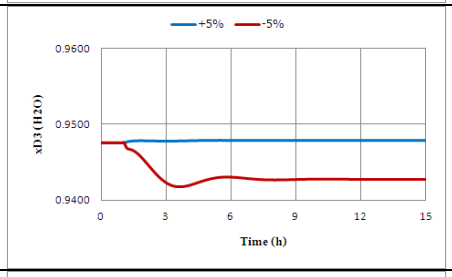
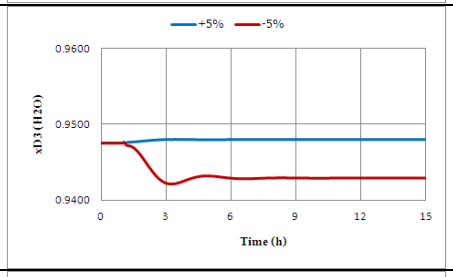
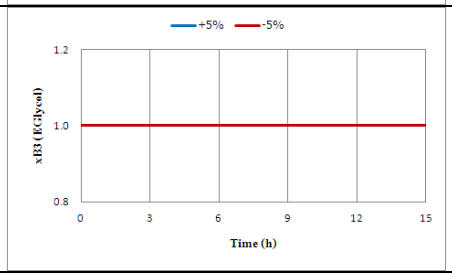
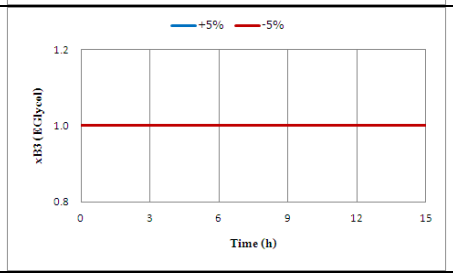
	BC	New design
m.		
n.		
o.		

Table 5.13 Dynamic responses with feed temperature change.

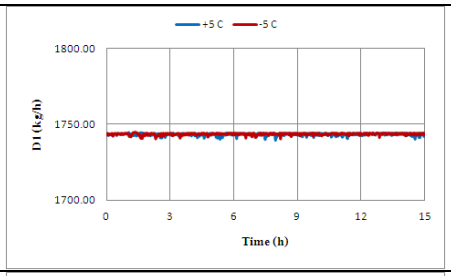
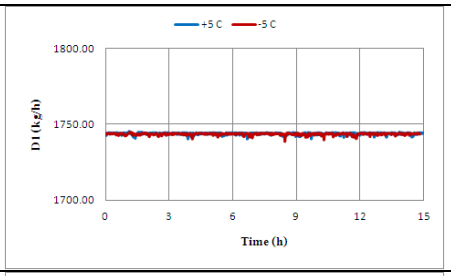
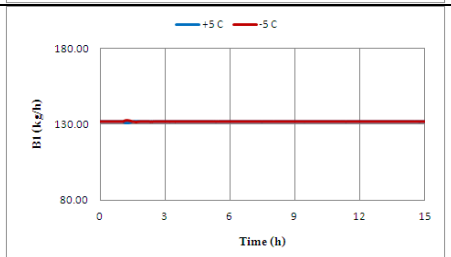
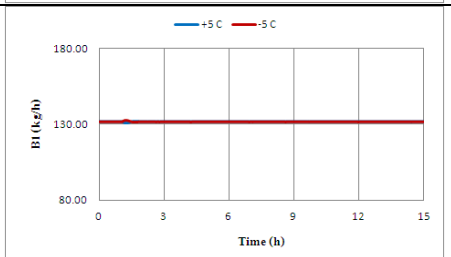
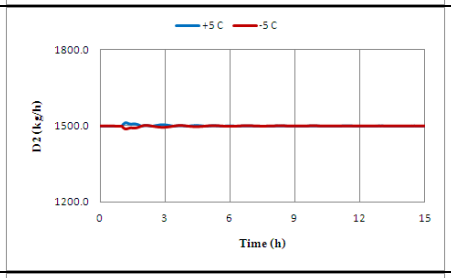
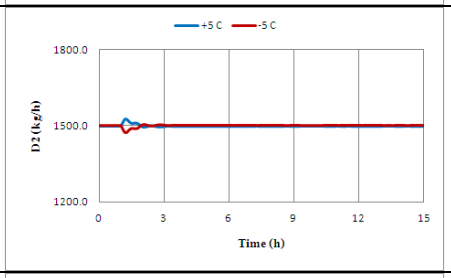
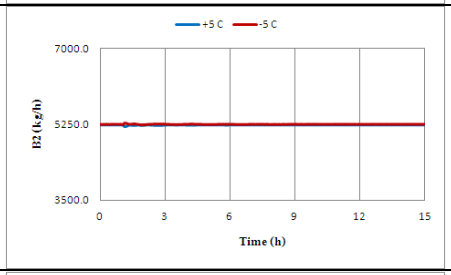
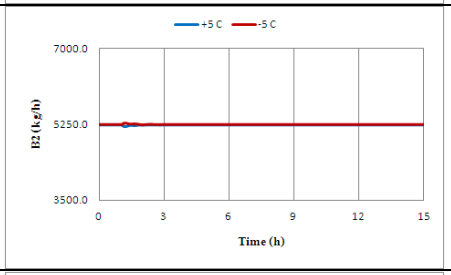
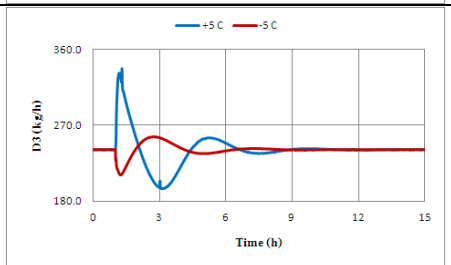
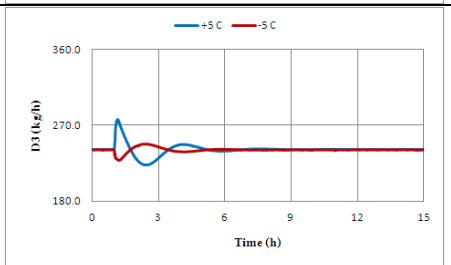
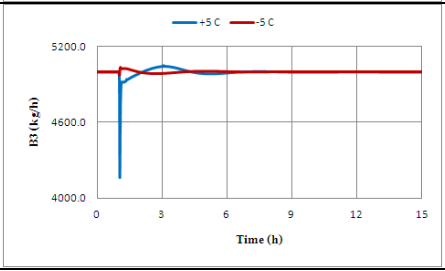
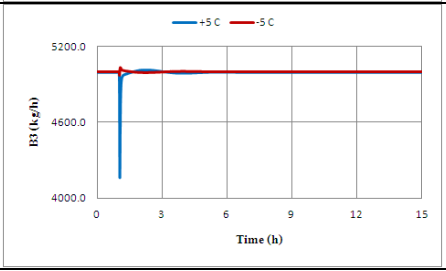
	BC	New design
a.		
b.		
c.		
d.		
e.		
f.		

Table 5.14 Dynamic responses with feed temperature change. (Continue)

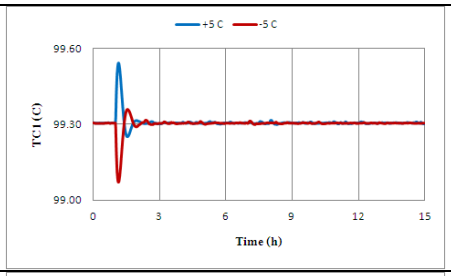
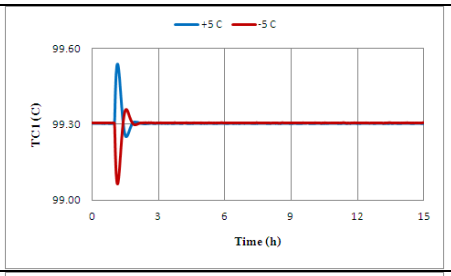
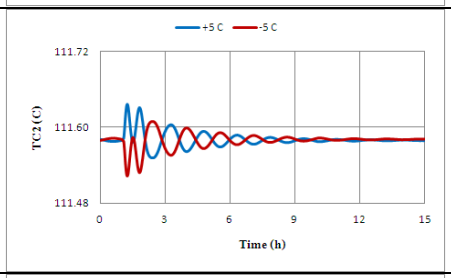
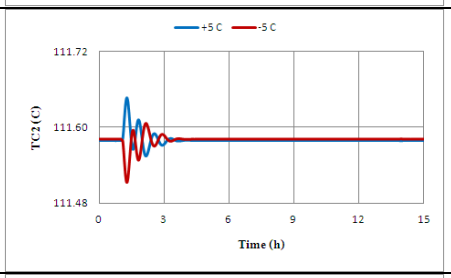
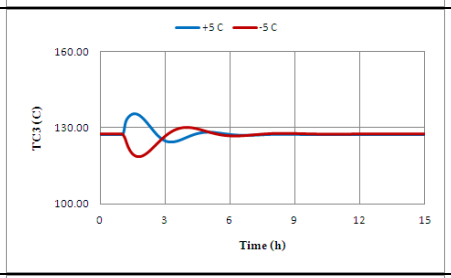
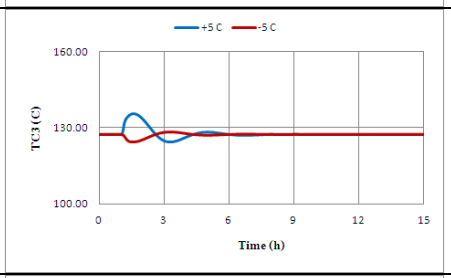
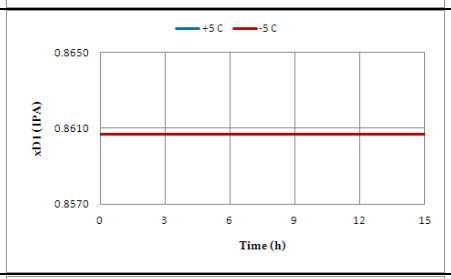
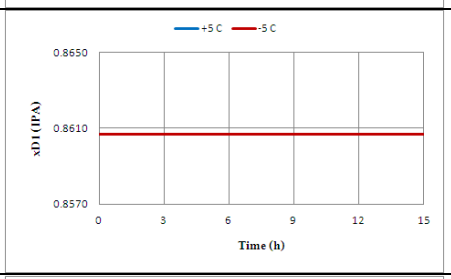
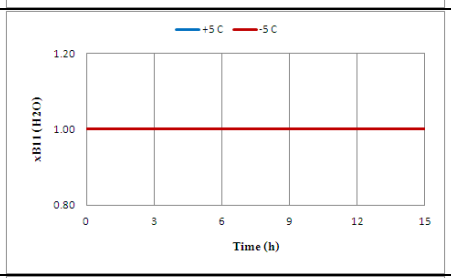
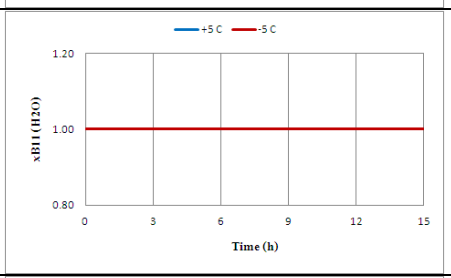
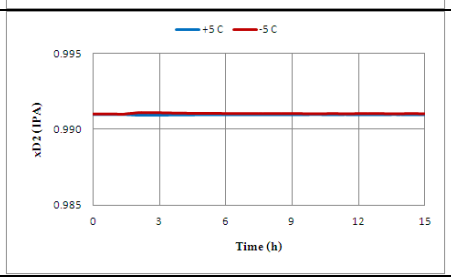
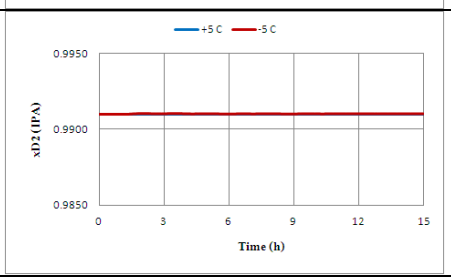
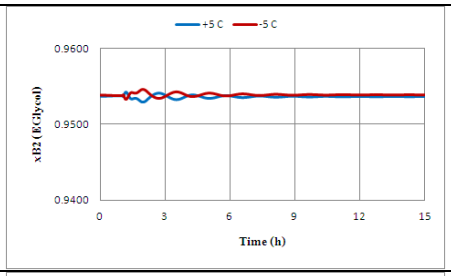
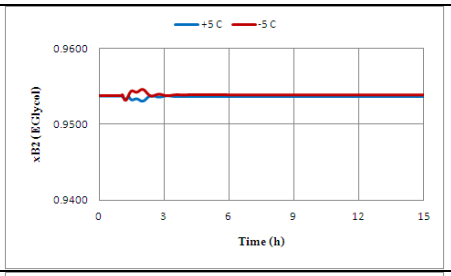
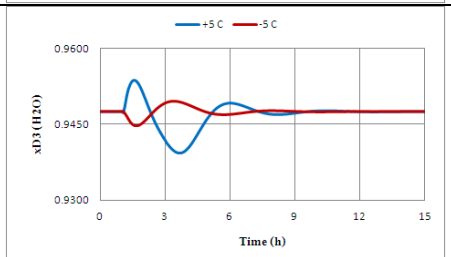
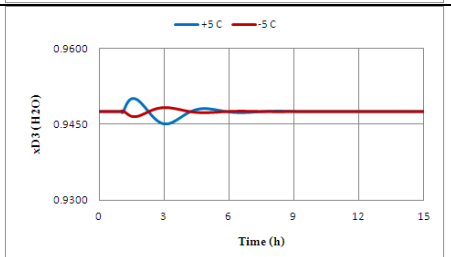
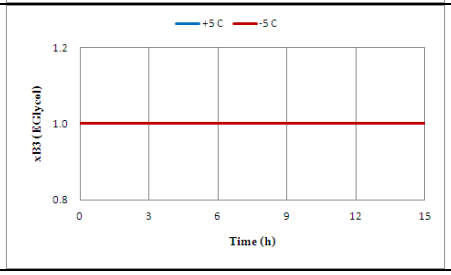
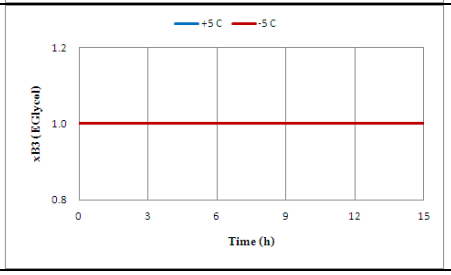
	BC	New design
g.		
h.		
i.		
j.		
k.		
l.		

Table 5.15 Dynamic responses with feed temperature change. (Continue)

	BC	New design
m.		
n.		
o.		

5.2 Integral Absolute Error analyses (IAE)

The IAE is used for evaluating the control performance of the process when the disturbances are tested. The values of IAE are analyzed from the safety loops, which are temperature, pressure and composition.

The IAE criterion was defined as

$$IAE = \int_0^t |e(t)|\, dt \qquad (7)$$

Note that $e(t) = y_{SP}(t) - y_{PV}(t)$ is the deviation (error) of the response from the desired set point.

Control loops are classified in to three groups: temperature, pressure and composition. Temperature loops consist of stage 11 temperature control of Column C1 (TC1), stage 56 temperature control of Column C2 (TC2), stage 3 temperature control of Column C3 (TC3) and ethylene glycol temperature control (TC4). Pressure loops consist of condenser pressure control of Column C1 (PC1), condenser pressure control of Column C2 (PC2) and condenser pressure control of Column C3 (PC3). Finally, composition loops consist of product composition in

distillate of Column C2 (xIPA), another composition e.g. composition of water in bottom of Column C1 and distillate of Column C2 (xH2O) and composition of ethylene glycol in bottom of Column C3 (xEGlycol).

Total IAE of each loop are divided by span. Span of temperature loop, pressure loop and composition are 50, 0.05 and 0.5 respectively. Then the results are multiplied by weight value. Weight value of temperature loop and pressure is 0.5, weight value of another composition is 0.8 and weight value of product composition is 1.

In the conclusion, the best control structure is the control structure has the minimum value of IAE. Table 5.16, 5.17 and 5.18 show the IAE value of each control structures which has been disturbed by several types of disturbances.

Table 5.16 IAE of control loop with feed rate changed.

Control structure	**Temperature**	**Pressure**	**Composition**	**SUM IAE**
Base case	0.174	**34.213**	0.033	**34.420**
New design	**0.164**	39.617	**0.015**	39.796

Table 5.17 IAE of control loop with feed composition changed.

Control structure	**Temperature**	**Pressure**	**Composition**	**SUM IAE**
Base case	0.120	31.103	0.402	31.625
New design	**0.110**	**29.308**	**0.020**	**29.438**

Table 5.18 IAE of control loop with feed temperature changed.

Control structure	**Temperature**	**Pressure**	**Composition**	**SUM IAE**
Base case	0.287	30.518	0.030	30.835
New design	**0.237**	**29.973**	**0.013**	**30.223**

From the given data can be summarized that the new design control structure has the smallest IAE value of composition loop with feed rate, feed composition and feed temperature change.

5.3 Utilities Cost

There are three types of the utilities in the process, Cooling water, low pressure stream and electrical power. The cost of the both utilities are 0.354 $/GJ $ for cooling water, 4.7 $/GJ for low pressure and 0.04 $ per kW for electrical power.

The electrical power of Base case and new design is fixed (P-101 = 5.27 kW) the cost of the electrical utilities are 1,846.61 $/year.

Table 5.19 Utilities cost when feed rate changed.

Control structure	**Cooling water (GJ/year)**			**Utilities cost ($/year)**	**Low pressure stream (GJ/year)**			**Utilities cost ($/year)**
	+5%	**-5%**	**Average**		**+5%**	**-5%**	**Average**	
Base case	60,619.2	53,874	57,246.6	20,265.296	76,825.2	69,554.4	73,189.8	**343,992.06**
New design	58,954.8	55,188	57071.4	**20,203.276**	76,912.8	69,554.4	73,233.6	344,197.92

Table 5.20 Utilities cost when feed composition rate changed.

Control structure	**Cooling water (GJ/year)**			**Utilities cost ($/year)**	**Low pressure stream (GJ/year)**			**Utilities cost ($/year)**
	+5%	**-5%**	**Average**		**+5%**	**-5%**	**Average**	
Base case	56,589	55,363	55,976.4	19,815.646	72,007.2	73,671.6	72,839.4	**342,345.18**
New design	58,254	53,260	55,757.4	**19,738.120**	72,094.8	73,584	72,839.4	**342,345.18**

Table 5.21 Utilities cost when feed temperature changed.

Control structure	**Cooling water (GJ/year)**			**Utilities cost ($/year)**	**Low pressure stream (GJ/year)**			**Utilities cost ($/year)**
	+5%	**-5%**	**Average**		**+5%**	**-5%**	**Average**	
Base case	55,100	57,202	56,151.6	**19,877.666**	69,554.4	74,635.2	72,094.8	**338,845.56**
New design	56,064	57,290	56,677.2	20,063.729	69,554.4	74,635.2	72,094.8	**338,845.56**

From the utilities cost table shows that for the disturbances tested, which feed rate, feed composition and feed temperature changed. Total utilities cost when feed rate change of base case is 366,103.966 $/year while total utilities cost of new design is 366,247.806 $/year. Total utilities cost when feed composition change of base case is 364,007.436 $/year and total utilities cost of new design is 363,929.910 $/year. Finally, total utilities cost when feed temperature change of base case is 360,569.836 $/year and total utilities cost of new design is 360,755.899 $/year.

Base case has the smallest utilities cost when feed rate and feed temperature change while new design has the smallest utilities cost when feed composition change.

CHAPTER VI

CONCLUTION AND RECOMMENTDATIONS

6.1 Conclusion

The Wongsri's plantwide control design procedure is used to design the control structure of the isopropyl alcohol dehydration process (new design). In the design steps, the procedure is simple and clear cut.

The performance of each structure is evaluated by the IAE value. The new design control structure is the best control structure for the feed rate, feed composition and temperature changed.

The total utilities cost of both control structures is slightly difference when has been disturbed by several types of disturbances.

6.2 Recommendation

The sensitivity test is not appropriate for process which changes component flow, composition and temperature make vapor fraction of stream change (e.g. vapor fraction change from liquid to vapor)

REFERENCES

Buckley, P. s. Technique of Process Control; Wiley : University of Michigan, 1984.

Douglas, J. M. Conceptual Design of Chemical Process, New York: McGraw-Hill (1988).

Drowns, J. J. Distillation control in Plantwide control Environment, Chap.20 in Practical Distillation Control, W. L. Luyben (ed.), New York: Van Nostrand Reinhold (1992).

Detjareansri, S. Plantwide Control Structures Design for Alkylation Process. Master's Thesis. Department of Chemical Engineering, Faculty of Engineering, Chulalongkorn University, 2009.

James J. Downs, Skogestad. An industrial and academic perspective on plantwide control. Annual Reviews in Control 35 (2011) : 99-110.

Konda, N. V. S. N. M., Rangaiah G. P., Krishnaswamy, P. R. Plantwide Control of Industrial Process, An Integrated Framework of Simulation and Heuristics. Industrial & Engineering Chemistry Research 44 (2005) : 8300-8313.

Kookos, I.K., Perkins, J. D. Heuristic-Based Mathematical Programming Framework for Control Structure Selection. Industrial & Engineering Chemistry Research 40 (2001) : 2079-2088.

Luyben, W. L.. Design and Control Degree of Freedom. Industrial & Engineering Chemistry Research 35 (1996) : 2204-2214.

Luyben, M. L., Tyreus B.D. and Luyben, W. L.. Plantwide Control Design Procedure. American Institute of Chemical Engineers 43 (1997) : 3161-3174.

Luyben, W. L., Tyreus, B. D., Luyben, W. L. Plantwide Process Control. McGraw-Hill : New York, 1998.

Luyben, W. L. Plantwide Control of an Isopropyl Alcohol Dehydration Process. American Institute of Chemical Engineers. 2006, 2290-2296.

M. Moran, Y. Arkun, G. Stephanopolous, Studies in the synthesis of control structures for chemical processes, Part I. Formulation of the problem. Process decomposition and classification of control tasks. Analysis of the optimizing control structures. American Institute of Chemical Engineers 26 (1980) : 220–231.

M. Moran, Y. Arkun, G. Stephanopolous, Studies in the synthesis of control structures for chemical processes, Part II. Structural aspect and the synthesis of alternative feasible control schemes. American Institute of Chemical Engineers 26 (1980) : 232–246.

M. Moran, Y. Arkun, G. Stephanopolous, Studies in the synthesis of control structures for chemical processes, Part III. Optimal selection of secondary measurements within the framework of state estimation in the presence of persistent unknown disturbances. American Institute of Chemical Engineers 26 (1980) : 247–260.

N.V.S.N. Murthy Konda, G.P. Rangaiah, P.R. Krishnaswamy. A simple and effective procedure for control degrees of freedom. Chemical Engineering Science 61 (2006) : 1184-1194.

Shinskey, F. G. Process Control Systems, 3d ed., New York: McGraw-Hill (1988).

Skogestad, S. Plantwide control. The search for the self-optimizing control structure. Journal of Process Control 10 (2000) : 487-507.

Skogestad, S. Conrol Structure Design for Complete Chemical Plants. Computers and Chemical Engineering 28 (2004) : 219-234.

Sommer, S. Melin, T. Design and Optimization of Hybrid Separation Processes for the Dehydration of 2-Propanol and Other Organics. Ind. Eng. Chem. Res. 43 (2004) : 5248-5259.

Suntrisrikomol, S. Plantwide Control Structures Design Procedure Applied to the Hydrodealkylation Process using Fixture Point Theorem. Master's Thesis. Department of Chemical Engineering, Faculty of Engineering, Chulalongkorn University, (2008).

APPENDICES

APPENDIX A

PROCESS CONDITION

Table A.1 Stream table for the isopropyl alcohol dehydration process

Name	FEED	1	2	3	WATER	4	Makeup	5	6
Vapor Fraction	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00
Temperature (C)	76.85	76.85	82.50	104.75	102.26	79.18	99.85	99.85	99.85
Pressure (kPa)	152.00	117.00	111.46	119.77	109.77	97.66	130.00	120.00	120.00
Mollar Flow (kgmol/h)	45.78	45.78	38.38	7.40	7.40	38.38	80.56	80.56	80.56
Mass Flow (kg/h)	1875.00	1875.00	1741.75	133.25	133.25	1741.75	5000.00	5000.00	5000.00
Liquid Volumn Flow (m3/h)	2.29	2.29	2.15	0.13	0.13	2.15	4.50	4.50	4.50
Heat Flow (kJ/h)	-1.35E+07	-1.35E+07	-1.14E+07	-2.06E+06	-2.06E+06	-1.14E+07	-3.56E+07	-3.56E+07	-3.56E+07
Name	**7**	**8**	**9**	**10**	**IPA**	**11**	**12**	**13**	**RAFFINATE**
Vapor Fraction	0.00	0.00	0.00	0.00	0.02	0.08	0.00	0.00	0.01
Temperature (C)	99.85	99.85	76.33	159.52	73.19	141.29	70.20	177.95	64.29
Pressure (kPa)	110.00	84.70	81.06	113.16	71.06	46.76	40.53	54.22	30.53
Mollar Flow (kgmol/h)	80.56	80.56	25.50	93.43	25.50	93.43	12.88	80.55	12.88
Mass Flow (kg/h)	5000.00	5000.00	1501.00	5240.75	1501.00	5240.75	240.99	4999.75	240.99
Liquid Volumn Flow (m3/h)	4.50	4.50	1.91	4.75	1.91	4.75	0.24	4.50	0.24
Heat Flow (kJ/h)	-3.56E+07	-3.56E+07	-7.84E+06	-3.82E+07	-7.84E+06	-3.82E+07	-3.63E+06	-3.44E+07	-3.63E+06

Table A.2 Stream table for the isopropyl alcohol dehydration process (Continue)

Name	14	15	16	RECYCLE					
Vapor Fraction	0.02	0.00	0.00	0.00					
Temperature (C)	171.98	179.05	179.03	179.03					
Pressure (kPa)	44.22	130.00	130.00	120.00					
Mollar Flow (kgmol/h)	80.55	80.55	80.56	80.56					
Mass Flow (kg/h)	4999.75	4999.75	5000.03	5000.03					
Liquid Volumn Flow (m3/h)	4.50	4.50	4.50	4.50					
Heat Flow (kJ/h)	-3.44E+07	-3.43E+07	-3.43E+07	-3.43E+07					

APPENDIX B

TUNING PARAMETERS

Table B.1 Type of controllers and tuning parameters of base case

Equipment	Controller	Controlled variable(CV)	Manipulated Variable (MV)	Type	Action	Nominal value	PV range	Tuning Parameters		
								K_C	τ_I	τ_D
Feed	FC1	Mass flow rate	Feed flow rate	PI	Reverse	1875(kg/h)	937.5-2812.5	0.087	0.018	
EGlycol feed	FC2	Total EGlycol flow rate	EGlycol flow rate	PI	Reverse	5000(kg/h)	2500-7500	0.065	0.088	
	RATO1	Mass flow rate	Set point of FC2	PI	Reverse	2.667	0-100%	0.653	0.887	
Column C1	TC1	Stage 11 Temperature	Reboiler heat input	PI	Reverse	367.7(K)	300-500	0.298	5.36	
	PC1	Condenser pressure	Condenser heat removal	PI	Direct	1.1(atm)	0.55-1.65	3.05	4.35	
	LC1	Reflux drum level	Distillate flow	P	Direct	50%	25-75%	2		
	LC2	Base level	Bottom flow	P	Direct	50%	25-75%	2		
	RATO11	Reflux flow rate	Set point of reflux flow control	PI	Reverse	0.637	0-100%	0.653	0.887	
Column C2	TC2	Stage 56 temperature	Reboiler heat input	PI	Reverse	384(K)	300-500	0.503	5.25	

Table B.2 Type of controllers and tuning parameters of base case. (Continue)

Equipment	Controller	Controlled variable(CV)	Manipulated Variable (MV)	Type	Action	Nominal value	PV range	Tuning Parameters		
								K_C	τ_i	τ_D
Column C2	PC2	Condenser pressure	Condenser heat removal	PI	Direct	0.8(atm)	0.4-1.2	11.3	2.73	
	LC3	Reflux drum level	Distillate flow	P	Direct	50%	25-75%	2		
	LC4	Base level	Bottom flow	P	Direct	50%	25-75%	2		
	RATO21	Reflux flow rate	Set point of reflux flow control	PI	Reverse	2.04	0-100%	0.653	0.887	
Column C3	TC3	Stage 3 temperature	Reboiler heat input	PI	Reverse	402(K)	300-500	0.237	15.2	
	PC3	Condenser pressure	Condenser heat removal	PI	Direct	0.4(atm)	0.2-0.6	0.39	0.182	
	LC5	Reflux drum level	Distillate flow	P	Direct	50%	25-75%	2		
	LC6	Base level	Makeup flow rate	P	Reverse	50%	25-75%	2		
	RATO31	Reflux flow rate	Set point of reflux flow control	PI	Reverse	1.5	0-100%	0.653	0.887	

Table B.3 Type of controllers and tuning parameters of new design.

Equipment	Controller	Controlled variable(CV)	Manipulated Variable (MV)	Type	Action	Nominal value	PV range	Tuning Parameters		
								K_C	τ_I	τ_D
Feed	FC1	Mass flow rate	Feed flow rate	PI	Reverse	1875(kg/h)	937.5-2812.5	0.087	0.018	
EGlycol feed	FC2	Total EGlycol flow rate	Makeup flow rate	PI	Reverse	5000(kg/h)	2500-7500	0.065	0.088	
	RATO1	Mass flow rate	Set point of FC2	PI	Reverse	2.667	0-100%	0.653	0.887	
Column C1	TC1	Stage 11temperature	Reboiler heat input	PI	Reverse	367.7(K)	300-500	0.277	6.73	
	PC1	Condenser pressure	Condenser heat removal	PI	Direct	1.1(atm)	0.55-1.65	3.05	4.35	
	LC1	Reflux drum level	Distillate flow	P	Direct	50%	25-75%	2		
	LC2	Base level	Bottom flow	P	Direct	50%	25-75%	2		
	RATO11	Reflux flow rate	Set point of reflux flow control	PI	Reverse	0.637	0-100%	0.653	0.887	
Column C2	TC2	Stage 56 temperature	Reboiler heat input	PI	Reverse	384(K)	300-500	0.614	5.03	

Table B.4 Type of controllers and tuning parameters of new design. (Continue)

Equipment	Controller	Controlled variable(CV)	Manipulated Variable (MV)	Type	Action	Nominal value	PV range	Tuning Parameters		
								K_C	τ_I	τ_D
Column C2	PC2	Condenser pressure	Condenser heat removal	PI	Direct	0.8(atm)	0.4-1.2	11.3	2.73	
	LC3	Reflux drum level	Distillate flow	P	Direct	50%	25-75%	2		
	LC4	Base level	Bottom flow	P	Direct	50%	25-75%	2		
	RATO21	Reflux flow rate	Set point of reflux flow control	PI	Reverse	2.04	0-100%	0.653	0.887	
Column C3	TC3	Stage 3 temperature	Reboiler heat input	PI	Reverse	402(K)	300-500	0.214	15.2	
	PC3	Condenser pressure	Condenser heat removal	PI	Direct	0.4(atm)	0.2-0.6	0.39	0.182	
	LC5	Reflux drum level	Distillate flow	P	Direct	50%	25-75%	2		
	LC6	Base level	Bottom flow	P	Direct	50%	25-75%	2		
	RATO31	Reflux flow rate	Set point of reflux flow control	PI	Reverse	1.5	0-100%	0.653	0.887	

VITAE

Mister Siwapat Tochan was born in Kanchanaburi on November 2, 1987. After graduating high school from Assumption College Ubonratchathani, He entered Ubonratchathani University in May 2006 and received his Bachelor of Engineering degree in Chemical Engineering in April 2010. He began his graduate studies in May 2010 when he entered the Graduate School of Chulalongkorn University and joined the Control and System Engineering Group at Department of Chemical Engineering.