Advanced Process Control with Applications in the Food Industry

Qingbo Meng

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Candidate Declaration

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2. None of the material contained in the thesis has been used in any other submission for an academic award.

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Name	*Qingbo Meng*
Date	*August 2020*
Award	*PhD*
Faculty	*MERI*
Director(s) of Studies	*Dr. Hongwei Zhang*

Abstract

Due to the requirements for enhanced food safety and different nutrient demand for customers, food process control is becoming an increasingly important issue in the food industry. Many advanced control methods, like adaptive control, predictive control, robust control and fuzzy logic control, have attracted increasing attention and there are many successful applications in the manufacture of dairy products in the last two decades. Applying a multi-effect falling film evaporator to remove the water from the liquid is widely used in the dairy industry. This thesis addresses some key issues concerning dairy evaporation process control which include system modelling, controller development and optimization as well as the results comparison.

Fundamentally, this thesis presents a study of three effects of falling film evaporator for the milk concentration process in the dairy industry as an example. The main aim, however, is to research, develop and demonstrate different advanced control strategies, such as model predictive control (MPC) and fuzzy logic control (FLC), applied to the evaporator system for the process control purposes. They both can deal with the complex non-linear processes. But MPC can maintain the system consistency, FLC has a simple control structure and more flexiable control rules.

A dynamic mathematical model of a three-effect falling film evaporator is developed by using MATLAB based on the mass and energy balance principles in this thesis for analysing the optimization and controllability of the plant. Both conventional and advanced controllers, such as conventional PID, auto-tuning PID,

Model predictive control (MPC), Fuzzy logic, are described to maintain and improve the mathematical model performances.

The product output concentration controlled by different strategies are obtained and compared. The results indicate that all controllers can achieve the desired targets (30%, 38% and 52% for the 1st, 2nd, and 3rd effect) within limits of acceptability, however, MPC is the most competitive advanced control strategy in this milk evaporation process.

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Chapter 1. Introduction

1.1 Background

An old Chinese saying goes like "Food is the paramount necessity of the people", which shows the importance of food to everybody in every country. Here in the UK, ‘the food and drink industry is the UK’s largest manufacturing sector, contributing £28.2bn to the economy annually and employing 400,000 people (The Food and Drink Federation, 2018).’ There is no doubt that control systems are playing an increasingly important part in the food and drink industry. A well-designed controller can make many contributions to the system performances, such as, reducing processing time and cost price, improving system work efficiency, and solving specific problems. Meanwhile, it also can maintain the output performances of the industrial plants, such as hygiene, nutrient, odour, colour, in order to match the required standard.

It is also very important to ensure that the nutrition of the food products is beneficial for the human body. Levine and Labuza (1990) have indicated that food nutrition affects people's behaviours. They found brain damage and poor social and intellectual skills in childhood are due to malnutrition in their early years. Meanwhile, they also pointed out that sleep time increases if people intake high L-tryptophan or carbohydrates, which also results in a decrease in alertness. Inadequate nutrition can also cause heart disease and even cancer.

Russel (1997) indicated that controllers, which are normally used in the dynamic process model, with a part of an intelligent structure are generally referred to as advanced control. In many cases, advanced controllers have better performances than conventional controllers in more complex systems. This study will prove it in the following chapters.

1.2 Motivations

Van der Grinten (1968) indicated three reasons for applying process control: to ensure or enhance process stability; to suppress the influence of disturbances, and to optimise the performance of the process. In food industries, most of the manufacturing plants are highly energy-intensive processes, especially for the large capacity scale processes. Thus, reducing manufacturing costs and energy while improving or maintaining product quality is the top priority from the commercial perspective. With the increasing competition and the changing demands, conventional controllers cannot provide satisfying performances in some fields due to its limitations. More and more advanced control strategies are being studied and implemented in food manufacturing.

Let’s take the dairy industry as an example. Agriculture and Horticulture Development Board (AHDB) Dairy (2016) has reported that the UK is the third-largest milk producer in the EU with 14.6 billion (highest annual figure since 1990) litres of milk being produced in 2014. However, the number of UK milk cows have reduced sharply between 1996 and 2015, from 2.6 million to 1.9 million (i.e. nearly 27% less). Satisfying the needs of the market with limited resources is a priority mission. On the other hand, dairy products are consumed by most people daily.

Tessema and Tibbo (2009) presented that in order to ensure dairy food is safe and healthy, milk products must be controlled to meet the standards, such as nutrition content, chemical composition, and levels of bacteria and other micro-organisms. Both problems can be solved by applying appropriate advanced control technologies to milk processing to increase productivity as well as maintain the quality of the product.

Hence, one of the motivations for this research is that efficient controllers are necessary in the food industry. The work in this thesis has been focusing on the development and applications of advanced controllers (MPC and FLC) for complex dairy manufacturing evaporation process to maintain and improve the system performances.

Many advanced control methods, such as MPC, robust control and fuzzy logic, have attracted increasing attention and have been applied successfully in many different process control areas in the last two decades. However, such studies and applications in the dairy industry are still very limited.

Most of these studies can be generally cataloged into two directions according to the different purposes: the first direction is focused on the reduction of energy consumption and improvement of the system efficiency during the evaporation process (Zhang, et al, 2018; Ramirez, et al, 2006; Schuck, et al, 2015); the second one is studied on a specific control strategy or process parameters to improve the process performances (Winchester, 1999; Bakker, 2004; Foley, 2011; Haasbroek, 2013; Ahmed, et al, 2017; Guichet and Jouhara, 2010; Muhammad et al, 2019). It is useful to develop and compare conventional and advanced control methods in one milk evaporation process to obtain their benefits and drawbacks for the future reference and research studies.

However, since manufacturing processes in the food industry are becoming increasingly complex, and many of their particular characteristics, such as non-linear and dynamic behaviour, multiple inputs and outputs, long time delays, loop interactions make the food manufacturing process difficult to model and control (Suarez, et al 2009).

The control strategies could be advanced, however, the challenges remain the same. The main difficulties involved in controlling the industrial process are the inherent nonlinearities and the multivariable nature of the system. The emphasis for advanced control is on developing strategies that are multivariable and are robust in the face of modelling errors and un-modelled disturbances.

1.3 Research Aims and Objectives:

The Ph.D. project aims to develop advanced process control solutions with applications in a dairy industrial evaporation process. The work will focus on developing such solutions for simulations on some challenging processes and possible implementations at industrial plants.

The objectives of the Ph.D. project can be divided into the following:

1. To develop a mathematical model of three-effect falling film evaporator.

- The first objective involves modifying and aggregating the literature and developing the mathematical model. Additionally, to validate the simulation model by comparing with the similar previously published research data.

2. To design and implement conventional Proportional, Integral and Derivative (PID) and online auto-tuning PID controllers.

- As one of the most classical control strategies, PID is still prevalent throughout the process industry. PID controllers will be developed as a fundamental method to compare the performances with other more advanced control strategies.
- Auto-tuning PID is also implemented as a PID-based controller to complete the control tasks.

3. To investigate the design and implementation of advanced control methods, including Model Predictive Control (MPC) and Fuzzy Logic Control (FLC).

- MPC and FLC will be developed and applied to the simulation model as advanced control strategies to achieve and improve the system performances, such as product concentration setpoint tracking, product temperature and flowrate control, disturbances rejection and so on.

4. To compare the performance of different controllers based on the simulated results.

- The system performances controlled by advanced controllers are going to be compared with the PID controller to investigate the advantages and disadvantages.

- Various aspects of the control performances, such as settling time, overshoot and control errors will be compared to evaluate the controllers' performances.

1.4 Contributions

The three-effect falling film evaporator in this study is extended and developed based on a signle evaporator. It provides an idea of creating a multiple effects evaporator for further research works. In addition, for the real food industries which are currently using single effect evaporator, this study is a potential reference to update their evaporator by adding the same effect to the system. Meanwhile, this study compares the simulation data to the data from previous published work to validate the model. This could be usefully reference for those researches which are model or simulation based works without industrial data. One more contribution of this study is the comparison and discussion of different control strategies, Not like most of other researches that focus on one specific angle, such as a process variable, a control method, or process cost. The four different controllers have their own benefit and drawbakcs that could be supportive for future studies to choise the appropriate controller for the evaporation process. The contributions are summarized below:

Model development and validation:

- A mathematical model of a three-effect falling film evaporator is developed using MATLAB/SIMULINK in this thesis.

- The four sub-systems, which are separators, heat exchangers, steam ejectors and disturbance plants, for the evaporator effects are developed to describe the complete evaporation process.
- To validate the mathematical model against data from from the previous similar studies.

Development and comparison of controllers:

- The conventional PID controller is applied as a fundamental control strategy to maintain and achieve the desired control objectives.
- Another PID based controller auto-tuned by MATLAB is also implemented to the system to find the control gains automatically based on the system frequency response obtained from the input/output data, which is more precise than manually calculation by using Ziegler-Nichols method.
- Other advanced control strategies, such as MPC, fuzzy logic control, have been developed to carry out the control targets. Simulation results are compared to the conventional PID controller to observe the benefits and improvements.

1.5 Outline of this Thesis

This Ph.D. thesis is organised in 8 chapters. In chapter 1 (the present chapter), the background of advanced process control and the difficulties of its applications in the food industry are described. Meanwhile, the motivations and outline of the present thesis are discussed. Some relevant introductory concepts for the developments in the following chapters are also presented.

A comprehensive literature review is conducted in Chapter 2 on the advanced control methods and their applications in the food industry. As an example, literature of multi-effect falling film evaporator modelling, different control strategies implementation and performance comparison, system optimisation in the dairy industry have been discussed. MPC and fuzzy logic are introduced and discussed as the main advanced control strategies in this chapter.

Chapter 3 presents a detailed study on modelling of three effects falling film evaporator in the production of the milk powder process, and model validation by comparing to the previous studies' data and results. The theoretical background of the evaporation process is briefly described. The mathematical model of a three-effect falling film evaporator is developed using MATLAB/Simulink. The conventional PID controller is applied to the Simulink model as a basic control strategy.

Chapters 4 5 and 6 introduce three different controllers respectively including auto-tuning PID, MPC and fuzzy logic control strategies and their development and application in this evaporation process control. Issues and difficulties relating to the development of different control strategies and techniques for applying them to dynamic system problems are discussed. A comprehensive overview of different controllers' performances is presented in this chapter. Meanwhile, at the end of each chapter, the simulated results are compared to discuss the advantages and benefits of each control strategy.

In the last chapter 7, the general conclusions to be taken from the thesis are presented. Some directions for future work have also been pointed out.

Chapter 2 Literature Review

2.1 Overview of Food Processing

The purpose of food processing is to transform the raw food materials from agriculture to the good quality products to satisfy the consumers. Industrial food processing includes many different operations, such as heating, drying, packaging, pasteurizing, mixing, cooling, forming and so on (Greeves, 1991).

Figure 2.1 shows a general industrial milk powder production process. It is a very good example to illustrate the various operations just in one food processing system because it contains product heating, evaporating, pasteurizing and spray drying.

In this section, three types of thermal treatment and a new heat technology called Ohmic Heating will be introduced. Other general processes, such as drying, sterilising, and evaporating process are also being described with their applications in the food industry.

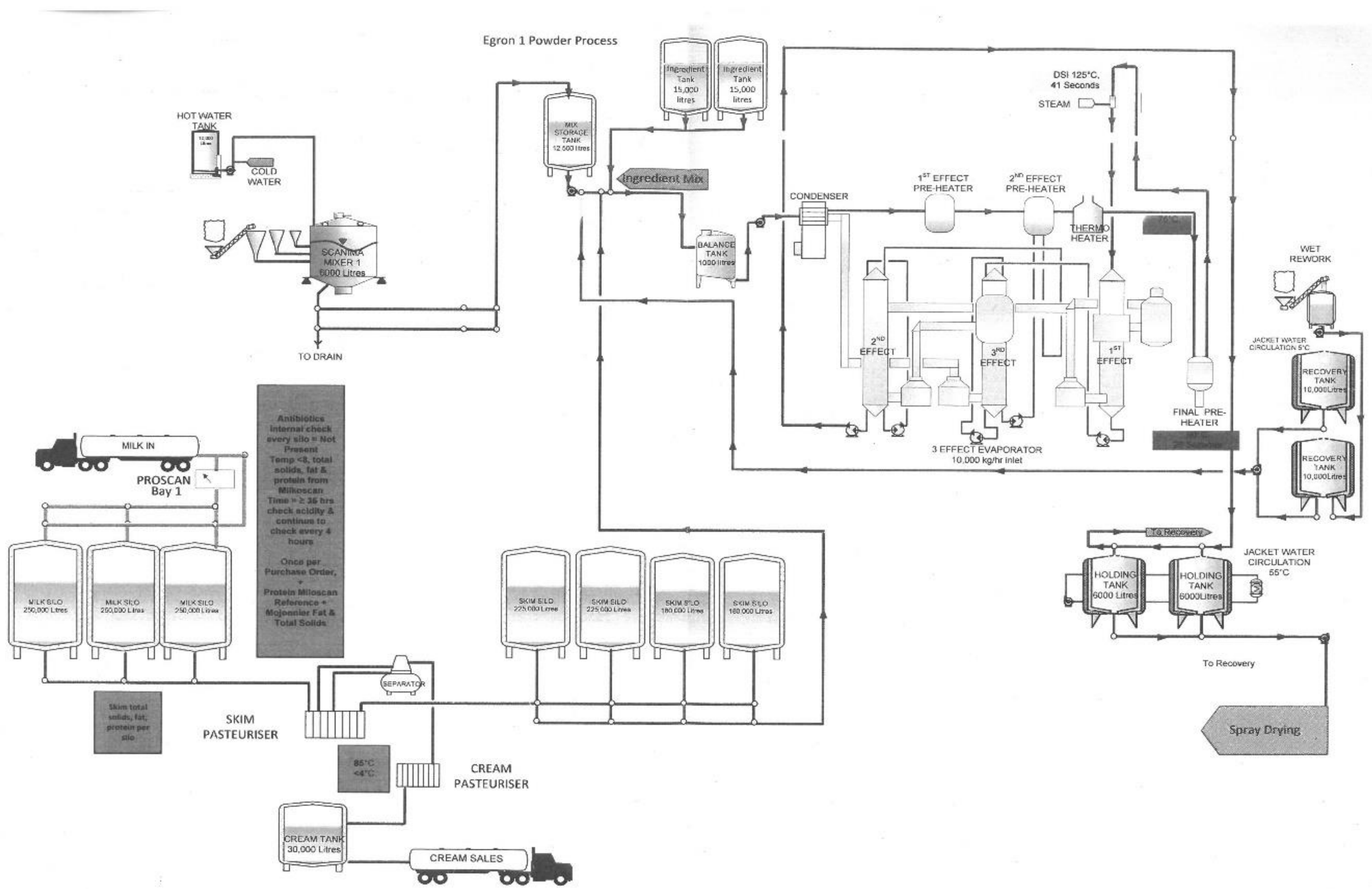


Figure 2.1 Industrial milk evaporation and spray drying process

2.1.1 Food Processing Industry Overview

The food industry today has become highly diversified with manufacturing ranging from small, traditional, family-run activities that are highly labour intensive, to large, capital-intensive and highly mechanized industrial processes. According to the Business Wire’s report (2019), the demands for different kinds of food, such as beverages, dairy products, meat, snacks, fruit, vegetables, and seafood, are still increasing year by year. Now the processed food industries are valued at over $2 trillion globally and consist of over 400.000 businesses. It forecasts from now to 2024, the food processing market is expected to reach an estimated $4.1 trillion.

European Union (EU) play an important role in the food processing industries in the world. European Food and Drink Industry reported that the EU is the first leading region which takes 44% of the turnover, more than double than the USA (20%) and China (19%). In 2019, the turnover number was € 1192 billion for the EU with 2.1% of gross value added (World Trade Organization, 2019).

Due to the increasing demands globally and its huge market size, even a small improvement in food processing technologies can bring huge benefits. To develop and apply advanced control strategies are one of the most preferable methods to achieve energy saving, to reduce the cost price and producing time, and to improve the efficiency of the plants and so on.

2.1.2 Food Processing – Thermal Treatment

The thermal process is ubiquitous and probably the most important process in the food industry that has been used for centuries. In food manufacturing plants, the thermal process heats the product at the desired temperature for a certain time range to achieve a specific purpose. Fryer and Robbins (2005) summarized three main objectives of heating:

- Preservation: this process aims to kill the bacteria or microbial and inactivate enzymes in some cases to ensure food safety and extend its shelf life. For example, the pasteurization of milk and sterilization of canned food.
- Enhance taste and flavor: it is normally applied to meat and vegetable products to lock the flavor and improve the taste at the same time. Additional heat is required when sterilizing the food to carry out physical changes to it.
- To develop the structure of the material: when cooking with flour or starch, such as baking bread or biscuits, heating can act to change the structure and function and also to develop the structure within the materials.

There are also a few other heating processes, such as spray drying, pre-heating, to save energy or reduce the manufacturing time. Therefore, to apply appropriate heating technologies can both improve the system efficiency and maintain product safety and quality.

The traditional heating technologies have been widely used and achieved huge successes. It depends on the contact with the hot sources or indirect contact through an agent, like a hot surface, to finally heat the products. The heat sources could be various, such as heating devices, heated steam, hot air, and water. Heat is generated directly from inside the food is one common characteristic of these conventional thermal technologies, which potentially improves the preservation of vitamins and nutrients in the food (Fellows, 2000). Nowadays, new heating technologies, such as ohmic heating, infrared heating, have been investigated and applied to replace parts or even the whole of the conventional heating methods due to their unique and specific characteristics for the special cases.

In the following sections, three types of heating methods and new thermal (heating) methods (i.e. ohmic heating) will be introduced and discussed. Relevant application examples in the food industry will be given in each section.

2.1.2.1 Conduction food heating

Conduction is one of the most commonly used heating methods in many food industries. It transfers the heat energy between objectives from higher to the lower temperature through direct contact with solid materials. A simple example of conduction in daily life is cooking food on a hot pan. There are several factors to determine the process of heat conduction, temperature gradient, the cross-section area of the material, length of the travel path and material physical properties.

The temperature gradient is the precondition of heat flow, decides the heat travel direction and rate. Heat conduction occurs from hotter to colder then stops when there is thermal equilibrium between two temperature differences. Bigger cross-section area and a longer travel path result in more heat loss. Greater the size and length of an object, the more energy required to heat it. The physical properties can be various in different materials. Specifically, the thermal conductivity coefficient is an important variable in conduction heat transfer. Metal materials are much better conductors than glasses.

One of the weaknesses of this heating method is the relatively low efficiency. Because of the contact and transfer heating pattern, most of the energy lost to the surrounding during the process. But conduction allows the product heated from outside to inside which can preserve the flavour. Heat exchangers are one of the classical applications of conduction thermal process in the food industry.

2.1.2.2 Convection food heating

Convection is the transfer of heat indirectly through fluids, such as liquid, air, between areas of different temperatures. Thermal energy is moved by the mass movement of groups of molecules. There are two types of convection, which are natural convection and mechanical convection, based on the movement of the heated molecules.

Natural convection occurs when a fluid is in contact with a surface hotter or colder than itself. As the fluid is heated or cooled it changes its density. This difference in density causes movement in the fluid that has been heated or cooled and causes the heat transfer to continue. Another type of convection is mechanical, also known as forced convection. It achieves the heat transfer by forcing a fluid past a solid body. The fluid can be hot or cold to heat or cool the products. The faster the fluid is moved, the faster the rate of convection.

The temperature gradient is one of the essential factors to cause heat flow. The fluids relative movement is another necessary factor in a convection heat process. Liquids and gases are usually preferable fluids because their particles can be made to flow and move from one place to another. Because the particles are the unit to contain energy, heat transfer happens during the movement of particles. A simple example of convection is the air conditioner in the room.

Convection is a more effective way of heat transfer than conduction. But additional energy source for operation of a mechanical fan, blower or compressor to facilitate the flow of the fluid is required, which both increase the cost and energy consumption. In the food industry, convection heat transfer is typically used in drying, freezing and steaming processes.

2.1.2.3 Radiation food heating

Thermal radiation is a method of heat transfer that does not rely on any direct or indirect contact between the heat source and the objects. Unlike conduction and convection, which need particles to carry and transfer energy, radiation transfers the heat by using electromagnetic waves with a separated energy source. Radiation can even work in the vacuum environment, such as in the space. The Sun is a huge source of radiation and warms the Earth from 150 million km away.

Two main factors affect the radiation heat transfer: emission and absorption (surface). The radiation emission is a direct result of vibrational and rotational motions of molecules, atoms, and electrons of a substance. The rate of thermal radiation emission increases with increasing temperature. Basically, all materials radiate thermal energy based on their temperature. Hotter objects can produce more radiating energy. But the heat energy can be interrupted by reflecting materials, like windows or aluminium foil. The absorption is mostly determined by the product surfaces. Some surfaces are better than others at reflecting and absorbing infrared radiation. It is generally accepted that dull, rough and matt surfaces are good absorbers of radiation than the shiny surfaces. The same surface material of two objects, thinner and more flat radiates heat energy faster.

Relatively lower energy consumption is the dominating advantage of radiation heat transfer comparing to the other types of thermal heat transfer. In addition, radiation does not need a medium to transfer the heat energy and heats product superfast in 'lightwave' cases, like the microwave. Most applications of radiation heating methods in the food industry are used in drying, evaporating processes, or some processes of heating material rapidly within a short time.

2.2 Ohmic Heating

Ohmic Heating (OH), also known as Joule heating, has attracted the attention of the food industry, as an alternative heating method. Palaniappan and Sastry (1991) claimed that due to the ionic species, such as salts and acids, contain in most food, it is possible to make the electric current pass through and heat the food itself.

Furthermore, Sarang, Sastry and Knipe (2008) pointed out that compared with conventional heating methods, OH can provide more rapid and uniform heating, which barely results in damaging the material. Similarly, Goullieux and Pain (2014) have tested through a very high-temperature within a short-time (HTST), not only can sterilise milk or fruit juice, but also change the taste slightly.

OH is not a new concept. In order to solve the deleterious effects, which is caused by overheating during the milk pasteurisation process, OH uses electricity to pass through the milk itself to heat. However, it was extremely expensive due to the limitation of technology in that era. Stancl and Zitny (2010) illustrate another reason why OH was abandoned, it was because the corrosion effects could not be controlled very well at that time. In the early 20th century, this technology was rediscovered. De Alwis and Fryer (1990) have proposed that solid and liquid food can be offered the same heating rate by using direct resistance heating method, meanwhile, avoiding heating time delay. Castro et al (2003) have proved that the OH system can help to maintain the colour and nutritional value of food. Ruan et al (2001) reported that nearly 18 applications of OH have been applied in the food industry in Europe, Japan, and the United States. Japan created the most successful OH system for strawberries processing control and fruits for yogurt; America OH technique is used on ready to eat meals with low acid.

It is generally believed that the definition of OH is a process, in which the "electric currents are passed through materials (foods) with the primary purpose of heating them by internal energy generation" (De Alwis and Fryer, 1990; Fryer and Li 1993; Knirsch et al, 2010). The schematic diagram is quite simple (see figure 2.2). It contains an AC power supply, which provides the power energy for the whole heating system, and electrodes, which help to adjust the electric field strength by changing the gap (Ruan et al 2001).

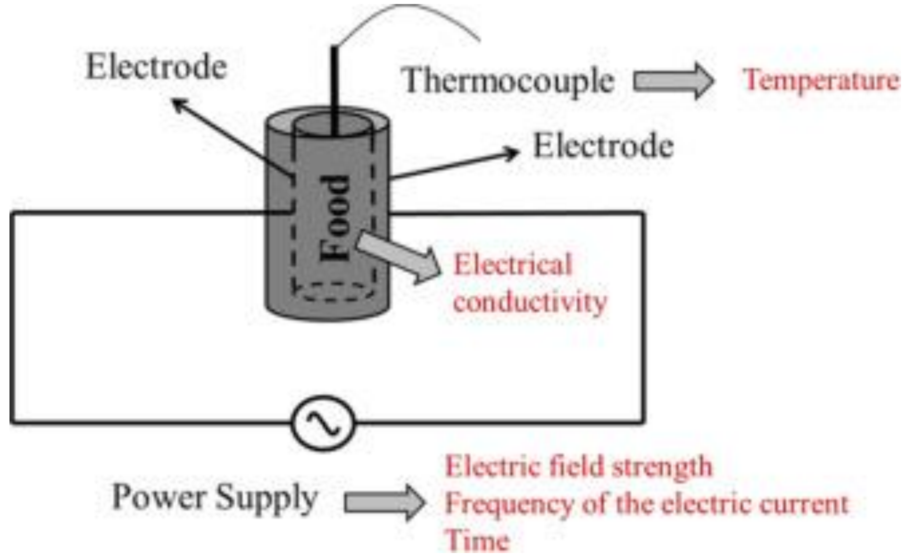


Figure 2.2 Simplified schematic diagram of ohmic heating and relevant parameters (Lopez-Pedrouso, 2019)

The main difference between OH and conventional heating (CH) method, which was discussed by Bozkurt and Icier (2012), is the way of heat transfer. CH transfers heat energy by "touching", which means the closer to the heat source, the more quickly it can be heated. However, there is no need for "touching" to transfer heat by using OH. The energy is dissipated into the food directly, which makes the heating rate very fast. Furthermore, Sakr and Liu (2014) pointed out that OH can avoid fouling of surfaces effectively. Due to these characteristics, a large number of actual and potential applications are available for OH, such as blanching, dehydration, fermentation sterilisation and so on (Sarang, Sastry, and Knipe, 2008).

2.2.1 Main advantages and disadvantages

The reasons why OH is attracting engineers' attention at present are because of its unique characteristics, which are not easy to achieve through other heating methods, such as CH, microwave heating or inductive heating. McKenna et al (2006) have indicated that the rapid heating speed and the relatively uniform heating characteristic are the two main advantages of OH. Heating materials faster avoids thermal abusing and wasting energy, meanwhile, reducing heating time is also a good way to prevent nutrient loss somehow. Skudder (1988) found OH lower the risks of fouling, which is normally caused by heat transfer from a surface to centre in CH and, of materials burning caused by overheat. Also, another attractive characteristic of OH is its high energy efficiency due to the dissipated heating way. Kim et al (1996) reported the electrical energy conversion into heat can reach approximately 90%, which can reduce the cost price of the industrial processing process. Another important advantage is that OH makes the precise temperature control of the heating system realisable (Engchuan et al). Although many positive features show the OH is an excellent alternative heating method, there are still several limitations of applications in industry. Hightech Europe (2012) reported two different limitations and one hazard. The first one is a requirement of the heated materials. The OH technology is not suitable for non-conductive materials because the current cannot pass through it to produce heat energy. Furthermore, OH is not suitable for multi-phases food with completely different conductivities. The second limitation is that OH works more suitable for aqueous systems than non-aqueous systems. Such as meat. Because solid-state material is easier to cause inactivation, overheating at surfaces and electrodes when transfer heat than liquid material.

2.2.2. Important factors of Ohmic Heating

There are plenty of parameters, which may generate from equipment, such as field strength and electrodes, and others from the materials themselves, like electrical conductivity, particle size, and ionic concentration. It has been mentioned by Kaur and Singh (2015) all these factors are found that can influence the ohmic heating rate.

2.2.2.1 Electrical Conductivity

Electrical conductivity is the ability of how well a martial conduct electric current. Among all the factors, electrical conductivity has a supreme influence in ohmic heating modelling (Fryer and Li, 1993). It has been focused on many published papers. Zell et al (2009) designed an experiment to test the relationship between electrical conductivity and OH rate by selecting five different kinds of meat (beef, pork, lamb, chicken, and turkey). They designed the temperature environment from 5 ℃ to 85 ℃. Results showed that the electrical conductivity value of all meats has a linear growth with the increasing of temperature. Chicken, which has the highest electrical conductivity value, can reach the temperature target fastest. Similarly, Shirsat et al (2003) did a similar study but just using pork cuts. They found the difference in conductivity between different parts of pigs due to the structural density or fat content. Furthermore, Sarang et al (2008) proved that the electric conductivity of fruit (red apple, golden apple, peach, pear, pineapple, and strawberry) increases linearly with temperature (25-140 ℃) during the OH process with a constant voltage.

2.2.2.2 Field Strength

Field strength, electric conductivity and OH rate have a positive correlation, which is indicated by Halden et al (1990). Additionally, Castro et al (2004) observed two strawberry pulps and filling have an increasing trend on OH rate and electrical conductivity with the application of stronger field strength. However, with the increasing of field strength, microbes inside have a higher risk to inactivate (Barbosa-Canovas et al, 1998), which results in membrane destruction (Bean et al, 1960).

2.2.2.3 Particle Size

Kim et al (1996) claim that one of the most effective factors that can influence the heating rate is particle size. Zareifard et al (2003) designed a two-phase carrot-starch mixed system, contains the different sizes of carrots (6 and 13 mm) and liquid phase with 4% starch solution and 0.5% salt. The experiment results revealed that the smaller the particle size is, the less time it needs to reach a target temperature, even in a two-phase food system. The same conclusion was also discovered by Palaniappan and Sastry (1991).

2.2.3 Literature of Ohmic Heating

Sinthiya (2015) designed a simplified OH equipment, which consists of a glass tank, several electrodes, DC power supply, and temperature controller. He applied a PID controller in order to improve the accuracy of the temperature control. The experiment showed the main advantages of OH such as rapid and uniform treatment with minimal heat damage and nutrient loss and reducing fouling compared to CH, on one hand, the shelf life of materials can be extended on the other hand. But Sinthiya only tested coconut water as a liquid material, so this result needs further evidence if the material is swapping with solid or mixed. Not like Sinthiya, a large majority of researchers prefer to build mathematical models or software models aim at analysis and compare the results with CH. Marra (2014) has analysed the pasteurisation of solid food by using a mathematical model of OH. He applied Laplace's equation to describe the electrical distribution inside the food and the heat transfer equation. The results showed that electrical conductivity is the key parameter to influence the pasteurisation time. In the 1990s, De Alwis and Fryer (1990) has designed a mathematical model named by their names, called 'De Alwis-Fryer' Model. The heat generation rate was calculated by using Laplace's equation together with transient energy balance without the influence of convection. Furthermore, this model was updated by Zhang and Fryer (1993), which include multiple spheres uniformly distributed on a lattice with a non-convective fluid. Another model was created by Sastry and Palaniappan (1992), called the 'Sastry-Palaniappan' model. Electrical conductivity and heat generation with a particle immersed in a mixed fluid were approximately replaced by circuit analogy. In the same year, Sastry (1992) improved this model by using circuit analogy instead of electrical resistance, which made the high solid concentration of continuous flow OH approached.

More recently, another model, which contains three different solid particles (potato, meat as well as carrot) and 3% NaCl solution, has been conducted to model OH pattern by Shim et al (2010). Their simulation applied computational fluid dynamics (CFD) codes with user-defined functions (UDFs) for electric field equations to model the transient heating patterns of each solid material and medium. The maximum temperature prediction error of this model is only 6 degrees. This study shows that the CFD model can support in designing of continuous flow OH with the pursuit of heating uniformity, furthermore, enduring food safety and quality.

Three-dimensional simulations of OH of highly viscous food, which heat in a chamber with sidewise parallel electrodes were undertaken by Shynkaryk and Sastry (2012). They used commercial finite element modelling (FEM) software to solve the electrostatic and energy transfer differential equations. The result of their work shows the current, fluid flow patterns and uniformity of heating are influenced by the heating chamber geometry.

Due to the limited research on OH, advanced process controllers can be studied and applied to improve both the heating efficiency and accuracy.

2.3 Drying, Sterilising and Evaporating in food processing

2.3.1 Drying

Drying is a very old method of the food treatment process. Ahmed et al (2013) pointed out that in prehistoric times, people knew how to remove water from meat, fruit or herbs through natural sources, such as the sun, wind, or bonfire. There are many benefits to keeping food dehydrated. Firstly, it prohibits the growth of enzymes and bacteria to maintain food quality. Secondly, the dehydration process helps extend the food shelf life. And last, dried foods are easier to store. Nowadays, drying technologies have become more and more mature, and various temperature levels and drying principles have been developed and applied in different industries. In the food industry, most applications of drying techniques cover dairy products, coffee products, powdered drinks, sugar, starch derivatives, fruit, and vegetables.

Burova et al (2017) have classified the drying techniques based on the dried materials' heat supply methods:

- Convective drying: it is a direct contact drying process between the material being dried and the dry agent which is often hot air or gas (usually mixed with air). The hot air or overheated vapor contacts the material continuously to provide the heat energy to evaporate the liquid.
- Contact drying: products are heated from the heating source through a separation wall or surface.

- Infrared radiation drying: infrared rays are the heating sources in this type of drying method. Heat energy used to evaporate the liquid from the material is generated by the infrared radiator.
- Vacuum drying: this drying process is under a low-pressure environment to reduce the requirement for the drying temperatures. The heat transfer could be either convection or radiation methods.
- Dielectric drying: products are heated in the field of high-frequency currents.
- Sublimation drying (freeze-drying): it works by freezing the products and reducing the pressure at the same time to allow the liquid to sublimate directly from solid to the gas phase.

Basically, convective and contact drying are considered to be conventional methods and the others are advanced drying technologies. Conventional methods normally sacrifice the product quality more or less in the drying process. The modern industry offers varieties of advanced drying technologies and equipment which can avoid this problem. There are so many different drying processes and equipment in the food industry, two classical drying processes, which are sun drying and spray drying will be discussed as examples.

2.3.1.1 Sun and solar drying

Sun drying is one of the most traditional methods for reducing the moisture content in food. To dry in the sun, the food is left exposed for a few days to achieve the desired moisture content. In developing countries, this method is popular because it is simple and without extra cost, especially if the countries are in the tropical region with continuous sunlight. Hot, dry breezy days and humidity below 60 percent are the best condition for sun drying. Higher temperature works better for dying food, but 30℃ (86°F) is the minimum temperature requirement.

However, not all food is suitable to be dried in the sun. For example, vegetables contain less sugar and acid which increases the risks for food spoilage; High protein products are not recommended due to the microbial growth under the sun. There are many major difficulties as pointed out by Sontakke and Salve (2015) of sun-drying, such as insect infestation, dust and dirt contamination, quality deterioration and low rate of heat transfer. That is why most applications of sun-drying are in the fruits industry because the high sugar and acid make them safe to dry in the sun directly.

In order to avoid these difficulties, chambers and trays made of wood or screens are used to contain the food to get rid of insects, dust and other kinds of contaminations. Ahmed et al (2013) indicated the best screen materials are stainless steel or plastic. Screens contain cadmium and zinc should be avoided because the oxidization of these materials can leave harmful residues on the food. Locate the trays at the place where has better air movement to allow the wind to take the moist away. When at night, the food must be covered under shelter to prevent moisture added back by the condensation process.

Solar drying is similar to sun drying, but applies solar cells inside the dehydrators, such as foil surface, to improve the efficiency of solar energy usage. Solar drying increases the temperature faster and reduces the drying time which lowers the risks of food spoilage.

2.3.1.2 Spray Drying

Spray drying techniques, also known as spray dryer has been extensively employed by the food industries for the conversion of a suspension or solution into a dry powder product. During the spray drying process, the material is converted into a fog-like mist (particles) and sprayed into the chamber where the liquid is exposed to a flow of hot air. When the particles get contact in the heated gas, the solvent is evaporated and leaving the dry powder product.

Costa et al (2015) illustrated that the main advantages of spray drying are the remarkable versatility and the rapid drying speed (less than a few seconds). Spray dryer has become the main equipment for drying fluid products in many different areas, such as the pharmaceutical process, biochemical industry, and dairy industry and so on. Besides, the fast-drying speed can reduce the exposure time of the particles to the high temperature, which makes it suitable for heat sensitive products, like milk and biological items (Rodrigues-Hemandez et al, 2005). Spray drying can achieve high precision product quality control through controlling the particle size (Gohel et al 2009). However, the equipment is too expensive, and generally have low thermal efficiencies due to the large volume of hot air that circulates in the chamber without contacting the particles.

In the dairy industry, the spray drying process usually follows an evaporation process, which increases the material solid content in order to improve the drying efficiency and save energy.

2.3.2 Sterilising

Sterilisation is actually a heat treatment process of product aiming at the removal of living micro-organisms, stopping enzyme activity and achieve long-term shelf stability. In the beginning, sterilising only used for canned and bottled foods. Now it has been applied to treat all types of food.

There are two different sterilising treatments based on the heating period. The general sterilisation process eliminates all living bacterial in an over 100℃ environment for a period of enough time, and stable the product shelf life at the same time. Another sterilising called Ultra-High Temperature (UHT) treatment, in which the product is heated in the desired high temperature (135-150℃) during a very short time. This method is especially applicable to low viscous liquid products, such as milk, cream, dressing sauces, and normally applied at one step before packaging (Desrosier & Singh, 2018).

According to various heating methods, sterilisation treatment process could be different. With moist heat, the heating temperature is between 110 to 120℃ during the 20 to 40 treatment time. For the dry heat method, Higher temperature (normally range from 160 to 180℃) and longer exposure time (up to 2 hours) are required than moist heat process. There are also other heating methods, such as Ionising radiation (X-ray, gamma radiation), UV irradiation and even chemical substances to complete the sterilisation process (Desrosier & Singh, 2018).

2.3.3 Evaporating

Evaporation is a common operation to remove the water from the liquid or solution in food processing industries in order to concentrate the products. It has been reported that the total dry solids of a liquid can be concentrated from 5% to 70%, or even higher, depending on the material viscosity. As a sort of heat treatment, the microbiological is damaged and shelf life is extended during the evaporation process. Meanwhile, due to the reduction in mass and volume of material, the cost of packaging, storage, and transportation are decreased. In some cases, evaporation is applied as a pre-treatment operation mostly before drying, especially for spray drying, to realize energy saving, efficiency improving and economy increasing of the dryer.

There are vast of evaporators used in food industries. The details will be introduced and discussed in the following sections.

2.4 Process Control Methods and applications

2.4.1 Overview

In the modern manufacturing process, control is an important part of the real plant to maintain or improve the system outputs. There are so many benefits that can be obtained with an appropriate controller in the system. A few of the main benefits have been given as follow:

- Guarantee the product safety and ensure the process running smoothly;
- Achieve the system consistency and reduce variation in the product quality;
- Reduce waste and optimise energy efficiency;
- Other specific tasks, like protect the system or the components.

With the improvement of the new control technologies, more and more conventional controllers have been replaced by advanced control strategies. Although applications with new control methods increase rapidly in a few decades, conventional controllers have their unique advantages.

In the following sections, both conventional and advanced control strategies are going to be introduced and described. Meanwhile, recent applications in different fields are given of each control strategy.

2.4.2 PID Control

PID control is one of the most common methods which has been successfully implemented in the different industrial field. Nowadays, in the process control area, 95% of the control closed loops are PID type. The first theoretical article of PID control was proposed by Nicolas Minorsky in 1922, which was applied to the automatic steering of ships. After that, the first 'three-term' control algorithm described by Taylor Instrument Company was added to their double response controller in order to reset condition actions called 'pre-act' in 1936 (Bennett, 1984). Widely use of the PID controller is due to its several advantages. Ali (2005) has indicated that the PID controller is not only easy to understand and implement in hardware and software but also with a simple structure that can do initialisation and operation without process model. However, there are some limitations pointed out by Khan and Rapal (2006) which PID has difficulty maintaining control performance with unknown nonlinearities, time delay, disturbances, and system variables change. In order to overcome these weaknesses, plenty of researches have tried to improve the conventional PID. Mann et al (2001) have developed a new PID tuning rule by using a separate time response analysis for optimising controller performance. It turned out that compared with other commonly tuning roles, the new one has a better result and the capability to cover time delay ranging from zero to any higher value. Many research studies focus on tuning PID to solve the time delay problem with different theory algorithms. For example, Zhang, Shieh, and Dunn (2004) have designed an updated PID controller to deal with load disturbances based on the linear quadratic regulator methodology for multi-input/multi-output (MIMO) system. There are many improved PID tuning methods being developed, such as fuzzy tuning, robust tuning, self-tuning and online tuning, which can solve the time delay problem effectively and the disturbances of the system (Ali, 1999; Ali 2005; Khan and Rapal, 2006; Sariyildiz et al 2015). Comparing with conventional PID, these new improved PID controllers have shown better performance, such as low overshoot and integral absolute error, less settling time and rise time (Khan and Rapal, 2006; Natsheh and Buragga, 2010; Yayeb and Ali, 2012). PID controllers also have wide applications in food engineering. Kozak (2016) has summarised that the most commonly used PID controllers in the food industry were in order to control flow, temperature, pressure, level, and other industrial process variables. Sato (2009) has developed a PID based predictive controller to a weight feeder, which is mostly used for dispensing materials in processing industries and food industry, to achieve high performance than the conventional one using a step-type reference signal.

2.4.3 Adaptive Control

The adaptive control system can be treated as a kind of self-adjust system, which has enough intelligence to adjust its characteristics according to the changing environment, like a feedback control system. Landau et al (2011) have pointed out that adaptive control has automatic adjustment ability to achieve or to maintain the desired level of control system performance when the parameters of the dynamic model are not known and/or change in time. During the 1950s, the first 'adaptive control' was applied in the aerospace industry to design autopilot (Black et al, 2014). Because of this successful application, adaptive control methods have attracted many engineers' interest and achieved a wide range of successful applications in aerospace. It has been extensively investigated and developed in both theory and application during the past few decades, and it is still a very active research field. Apart from self-adjust, there are some other advantages of adaptive control. Cao et al (2012) have indicated that the adaptive control system can solve complex simulations with unpredictable parameter deviations and uncertainties. However, this method has its limitation, it has major difficulties in dealing with nonlinear and large time-delayed processes. That is the reason it is not implemented successfully in the industry. But there is some published research that overcomes its disadvantages. Rincon and Angulo (2012) have proposed to use polynomials to approximate the highly nonlinear plants behaviour and developed an adaptive control scheme to control the speed of permanent magnet synchronous motor (PMSM). A nonlinear model-based adaptive control proposed by Lee (2012) is applied for a solenoid-valve system. There are also several successful applications of adaptive control in food science. Alonso, Banga, and Perez-Martin (1998) have provided several adaptive control techniques, such as self-tuning regulator, stable adaptive control and adaptive internal model control instead of simple PID-type controller in order to improve the performance of temperature tracking control strategy in batch sterilisation processes.

2.4.4 Robust Control

Robust control is a controller design method that deals with model uncertainty, particularly focusing on the reliability (robustness) of the control algorithm. Robust control is similar to adaptive control; both of them are feedback systems, aiming at solving system uncertain parameters difficulties (Ackermann, 1993). However, there are some differences between these two methods. Qu (2003) has pointed out that if the system unknown parameters, plant or corresponding controller are estimated online and estimates are used to synthesize stabilising control, this control system is adaptive. If the system could have a stable and designed performance controlled by a fixed controller, but there are specific uncertainties, such as unknown functions, parameter variations, unmodeled dynamics or disturbances, the control system is robust. As a new control strategy, robust control has been well developed recently. Halbaoui et al (2011) have indicated the main reason, which makes robust control as an invaluable tool; that is it can generate control laws that satisfy the desired requirement performance and maintain this result performance guaranteed during operation. In other words, once the controller is designed, the system performance is always accepted without changing parameters. Like adaptive control, Williams (2008) has indicated that robust control has great achievement in aircraft and spacecraft field because system stability and reliability are the top priorities in these two fields, also the system uncertain variables can be estimated. Furthermore, robust control is suitable for the process control applications which naturally have large uncertainty and small stability margins. There are many types of researches working on plant uncertainty extension, for example, Horowitz (1993) invented quantitative feedback theory based on the classical control theory; Robust pole placement has been described by Ackermann in the same year; Wang and Stengel (2002) have designed robust controllers with a combination of probabilistic design and evaluation with feedback linearization, and back-stepping techniques. Robust controller systems have a high-level requirement for the designer because the system should run smoothly without many operators' attention once the design is accomplished. However, robust control is a new method, thus there are not many published papers and industrial applications in the food processing control field. Sendin et al (2010) have described an efficient and robust multi-criterion optimisation method that can be successfully applied to a large non-linear dynamic system in food thermal sterilisation processing. Future potential work of robust control, which presented by Hansen and Sargent in 2016, is developing and improving hybrid systems with other different control strategies.

2.4.5 Model Predictive control

Predictive control, known as Model predictive control (MPC), is a multivariable control algorithm used to predict the future output of the system based on the historical information and the predictive model of the process. The basic MPC concept explained by Seborg et al (2011) is that the system future output values can be predicted from a reasonably accurate dynamic model. The objective of MPC calculation is to determine a sequence of control moves, which is manipulated input changes so that the predicted response optimally moves to the setpoint. As one of the most popular and successful model-based control strategies, Model Predictive Control (MPC) has been implemented in many processes control industry because of its four main advantages:

(1): The dynamic and static interactions between inputs, outputs and disturbances can be captured by the process model;

(2): Constraints on inputs and outputs are considered systematically;

(3): The control calculations can be coordinated with the calculation of optimum setpoint, also called the control target.

(4): Accurate model predictions can provide early warnings of potential problems.

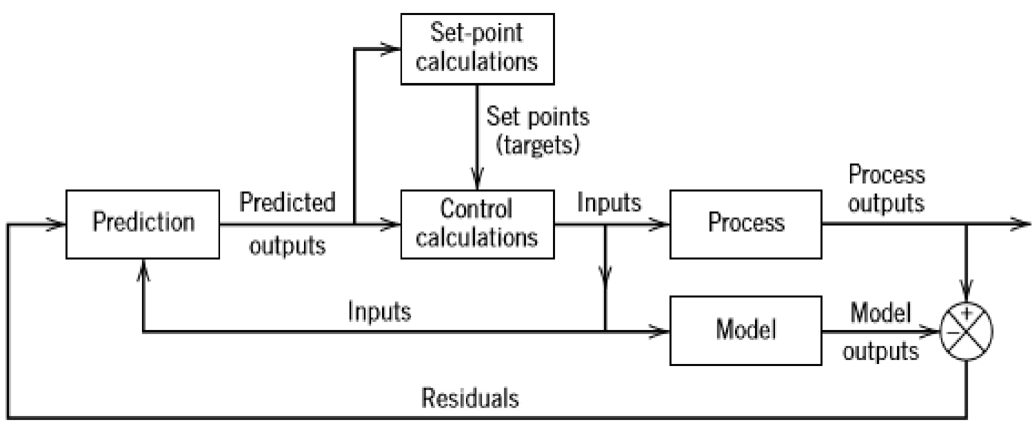


Figure 2.3 Block diagram of Model Predictive Control (Seborg et al, 2011)

Nunes (2001) reported that MPC has become one of only a few advanced control methods that can be applied successfully in many different industrial applications. There were various articles of MPC introduced in the late 1970s. The most famous one called model predictive heuristic control, later known as model algorithmic control (MAC), developed by Richalet and his co-workers in 1978. Since then, MPC technology has progressed and widely applied in a different area, such as chemical, pulp, petroleum, and paper industry over the past three decades (Julien et al, 2004).

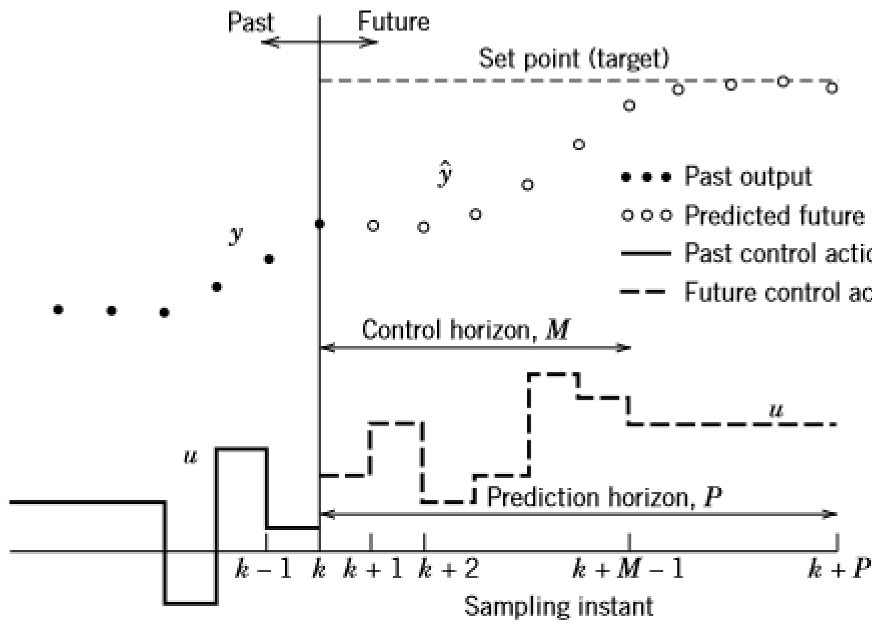


Figure 2.4 Basic concept of Model Predictive Control (Seborg et al, 2011)

A survey data given by Qin and Badgwell (2003) has shown the MPC applications increase approximately double between 1995 and 1999. Linear MPC is rapidly becoming a standard control methodology and is finding successful applications in the manufacturing industry at that period due to the attractive economic benefits (Richalet, 1993; Austin & Bozin, 1996). It is believed that nonlinear MPC will provide further control improvements.

In the 1990s, researches focused on theoretical improvement, behaviour reviews, and industrial applications. For example, Rawlings (2000) has provided an introductory tutorial of MPC aimed to broaden practitioners' perspectives in the area of control technology. Allgower et al (1999) have presented a more comprehensive overview of nonlinear MPC and moving horizon estimation. In 2000, the theoretical result behaviour of MPC algorithms on a closed-loop system has been introduced by Mayne et al. in the following year, Kulhave et al (2001) have described practical MPC in industry and potential improvement in the future from the vendor's angle. Recently, there are several reported applications of MPC focusing on technical capability. Finite Control Set Model Predictive Control (FCS-MPC) is a simple and powerful method developed by Kouro et al in 2009 to control the power converters. This method can improve the system performance, efficiency, and safety by controlling different variables with constraints and some special system requirements, such as reactive power, common-mode voltages and so on. Another MPC application is in the automotive industry. Ripaccioli et al (2009) applied MPC as an energy management controller of Hybrid Electric Vehicles in order to optimise the system performance, such as fuel consumption. According to Gilbert and Kolmanovsky (2002), MPC usually has better performance than other control techniques on the prediction aspect. Although the MPC theory has matured considerably, there are still limitations, such as Dai et al (2012) pointed out that the main disadvantage of MPC is that it has difficulty solving the system model with uncertainties. In order to overcome these limitations, combined with other control strategies is a potential research direction of MPC in the future. For example, Mayne (2014) has indicated adaptive MPC and robust MPC is the development trend for MPC in the future to solve the system uncertainties problems, and inspirable articles have been published by Marafiore et al (2012) and Yu et al (2014) respectively.

In the last two decades, according to Kurtanjek (2008), over 5000 applications of MPC have been applied in industries, but most of them are in chemical and electrical discipline rather than fermentation or food process industries. However, there are still several successful applications of MPC applied in the food industry. For example, Tan and Hofer (1995) have developed a self-tuning predictive controller, which was based on feedforward and internal-model-based predictive control strategies for temperature control in the food extrusion process. In their experimental test, an extended prediction window was used to control and enhance robustness against uncertainties in time delays. Another very recent application is about MPC of an industrial baker's yeast drying process described by Yuzgec et al in 2008. They have investigated a mathematical model consist two differential equations, which are obtained from heat and mass balance inside the dried granules in order to predict two important parameters during the drying process called moisture content and product activity. The result performances are particularly satisfactory for the drying process of a baker's yeast.

2.4.6 Fuzzy Logic Control

Fuzzy control is an intelligent control method based on mathematical foundations with fuzzy set theory and functions to generate control decisions, as well as to describe the system process and dynamic characteristics. Al Gizi et al (2013) have introduced two main kinds of fuzzy logic based controllers. The first one is the "Mamdani" type which is adaptive without explicitly identified system. The second is the “Takagi-Surgeon (T-S)” type which is indirectly adaptive with a system identified by using the T-S fuzzy model.

It is generally believed that fuzzy control has a fast speed development as a potential control strategy for complex problem-solving. Fuzzy set and fuzzy logic theory, including mathematical algorithms, was first introduced by Zadeh in 1965. Since then, this method has gained the interests of many scientists from various research areas because of its unique benefit. Singh et al (2006) have pointed out that due to the capability of approximating any real function of the fuzzy system, a mathematical model is not necessary and all available information can be used in the design of a fuzzy scheme about the process. Zrida et al (2003) have proposed a fuzzy logic based controller to control the feedback data flow rate in order to ensure the system performance stability in one source single bottleneck communication system. The system controller not only solved the uncertainty caused by the communication channel delay but also simplified the control system. Moreover, Lim et al (2001) have applied fuzzy logic prediction successfully on connection admission control and congestion control in high-speed networks to enhance the efficiency of the control system. Applying fuzzy control in food processing is not an easy task. Perrot et al (2006) and Welti et al (2002) have concluded several reasons: too many parameters in the food industry must be taken into consideration; the process control is highly non-linear; biochemical or microbial changes should be considered with the temperature changes. However, there are still many successful applications of fuzzy control in the food science area. Voos et al (1998) have developed a fuzzy control system of the drying process in the sugar industry. Curt et al (2002) have indicated five Takagi-Sugeno models to control the quality of sausage ripening time. Perrot et al (2004) have described a decision help system to control the cheese ripening process based on fuzzy logic strategy. In the meantime, more papers emphasised on the control techniques such as Kupongsak and Tan (2006) have combined fuzzy set and neural network techniques to determine the set point of food process control to produce products of certain desirable sensory quality. Perera et al (2010) have applied high performance nonlinear fuzzy logic controller, which can lower 8.85% of absolute errors of the system performance, in a soft real-time operation drying machine. Podržaj and Jenko (2010) have found fuzzy logic controller can be used for a high degree of precision temperature process control.

2.4.7 Multivariate Statistical Process Control (MSPC)

MSPC is an advanced data analysis approach that can identify the underlying pattern of a data set, also reveal the relationship and impact between variables. Instead of using 'univariate' statistics, such as, mean, median, deviation and so on, in traditional statistical process control, MSPC focuses on 'Multivariate' analysis in order to understand complex process behaviours and control the system better. Hotelling (1947) has applied a multivariate process control technique to solve the bombsights problem. It was the first time that this new control strategy was introduced. With the development of this technology, Jackson (1991) has stated in his book that four conditions should be fulfilled in any multivariate process control procedure: (1) the question 'Is the process in control?' must be available to answer; (2) the correlation of the variables should be taken into consideration; (3) an overall

Type I error (probability of a sample result being outside the control limits when the process level is at the mean or the standard established for that process) should be specified; (4) the problem or the reason can be found if the process is out-of-control. The last condition is much more difficult than the other three, especially with the variables number growth. Recently, multivariate statistical techniques have rapidly developed in the area of statistical process control (Woodall and Montgomery, 1999) due to its several benefits for industry, for example, prevent process failures, improve product quality, reduce process costs, and optimise processes. Two multivariate statistical techniques, known as Principal Component Analysis (PCA) and Partial Least Squares (PLS), are well developed in the process control field.

2.4.7.1 Principal Component Analysis

Principal Component Analysis (PCA) is one of the most popular multivariate mathematical techniques, which can be used in almost all scientific data analysis disciplines. The core concept of PCA is extracting the most correlated information (Principal Component), which could affect the result performances directly, from the data table. Hamilton (2014) has indicated that the main job of PCA is transforming the data into a new, lower-dimensional subspace; into a new coordinate system, where the first axis corresponds to the first principal component. The greatest amounts of variance are explained by this first principal component (PC), which is called 'artificial variables' named by Pareek (2011).

As one of the most wildly used data analysing methods, PCA has abundant applications in many different fields, especially which contain a high dimensional dataset for analysis, such as genetics, business, and social science, chemical and food industry and so on. Shlens (2005) pointed out the reason why PCA is so widespread because it reveals the hidden statistics and extracts important information by providing a mathematical method to reduce a complex dataset to a lower dimension, as a result, it can improve work efficiency. PCA has more than 100 years of development history. It is generally believed that the most famous paper, which mentioned the fundamental ideas of PCA, was published in 1901 by Karl Pearson titled 'On Lines and Planes of Closest Fit to Systems of Points in Space' (Reris and Brooks, 2015). In that article, Pearson did the mathematical calculation in order to find the best-fitting line which can best fit a series of points. In the following year, MacDonell (1902) completed an application, which analyse the relationship between seven physical traits and crime in 3000 criminals. However, after that, few works attempt to continue to do further research according to Pearson's idea until 1933 (Reris and Brooks, 2015). Hotelling's research and publications about PCA are regarded as classic because he applied multivariate data analysis concepts and methods in psychometrics between 1930 and 1940, which is accepted as a revolution of PCA (Brereton, 2000). Nowadays, with the mathematical theory improved, many PCA-based controllers have been introduced, for example, Kernel PCA (Stefatos & Hamza, 2007; Scholkopf et al, 1996) and Non-Linear PCA (Scholz & Vigario, 2002; Friston et al, 2000), meanwhile, PCA has had very broad applications in relevant data analysis area.

As one of the most wildly used data analysis methods, PCA has a unique character: Dimensionality Reduction, which is the most important character and mostly accepted by researchers. Asadi et al (2010) have listed many applications of PCA, reducing the dimension of the data is the main one without much loss of information. Similarly, Ghodsi (2006) has pointed out that it is generally believed high-dimensional data are always multiple, indirect, which typically cannot be measured easily. With the help of dimensionality reduction, less or irrelevant vectors are eliminated, which makes the data analysing more effective. Another benefit of reducing high-dimension is in order to make the data visible. Tran (2011) has indicated a point of view that making data more visual is one of the main reasons to reduce the dimensions. Furthermore, he also illustrated another reason why we prefer low-dimensional because it has fewer distractions. In other words, only significant variation described within the recorded data can simplify data analysis tasks accordingly (Kruger, 2008). Although PCA has many advantages, there are still several limitations. A major challenge that pointed out by Ringner (2008) is deciding how many and which components to use in the subsequent analysis. After then, Karamizadeh et al (2013) have presented another key disadvantage of PCA that is evaluation of the covariance matrix accurately. Additionally, there is one more character we need to consider that is PCA can only be used for analysing data which has correlations. In other words, if the variables are independent, PCA is useless.

There are a lot of applications of PCA in the food industry, especially in the dairy products processing process, because of the different amount of nutrient level, heating treatment temperature or time could have completely different effects to the final products. So to find the principal components, which can mainly affect the result performance, is a necessary step in data analysis. Villar et al (1996) have done microbial populations analysis using PCA in refrigerated raw cow's milk. He collected 402 samples from 18 different dairy farms, which all have more than 25 milking cows, located in the central area of Cantabria (Northern Spain). The findings showed the hygienic status of a farm dairy equipment (include delivery tank) is the main reason for the increase in the number of bacteria. Radzol et al (2014) have applied PCA to remove several highly correlated elements to melamine for detecting the melamine content earlier. Research shows that PCA is suitable for more than 90% of data reduction of high dimension. There are more PCA applications in the milk industry, for example, Vallejo-Cordoba and Nakai (1994) has established an objective and rapid analytical system for milk shelf-life prediction; Amari et al (2009) used PCA to discriminate the raw milk quality with a success rate of 99.94% depending on the storage time. Not only in milk production, but PCA also has many applications in other dairy products, such as cheese and fermented food products. Frau et al (1999) used PCA to evaluate the physical properties of a kind of cheese (Mahon cheese). In this article, PCA was applied to combine measurements for describing the group of physical data, to establish the relationships between the different physical variables and to detect the most important factors of variability. As a result, eight physical variables were reduced to two independent components, which contain 84.4% of the total variance. Another published paper did similar research in milk cheeses by Bromatologia in 1998. This study used ewes' milk cheeses as a sample and PCA was applied in developing and selecting parts to evaluate the main sensory characteristics of ewes milk cheeses. The data analysis results show that about 84% of the sample variability, which already contained enough factors to study the importance of each attribute, was explained by the first and second principal components.

2.4.7.2 Partial Least Squares (PLS)

PLS is an efficient statistical regression technique that is capable of both reducing dimensions of data set and predicting dependent variables from a large set of independent variables. This method originated in social science and introduced by Herman Wold in the 1960s as an econometric technique. Wold also developed a least square algorithm called Nonlinear Iterative Partial Least Squares (NIPALS) in 1973. However, it first became popular in computational chemistry due in part to Herman's son Svante (Geladi & Kowalski, 1986) and in sensory evaluation (Martens & Naes, 1989). Similar to PCA, PLS can extract hidden information in problems with multiple correlated variables effectively. Besides, many research studies have been successfully applied PLS to industrial processes. Boulesteix and Strimmer (2007) have applied PLS to high-dimensional genomic and proteomic data analysis which proved that PLS can offer great flexibility and versatility than other methods. Zhang and Lennox (2003) have developed a predictive model by using PLS in a fed-batch fermentation process to provide an accurate inference of quality variables that are difficult to measure on-line, like biomass concentration. This model also can be used for fault detection and isolation capabilities. Nocita et al (2013) have published that it is possible to predict soil organic carbon content using the PLS regression approach. Song et al (2007) have successfully proposed a PLS-based model predictive control methodology to accommodate the characteristics of a large-scale manufacturing process. The results show that the PLS-based model can minimises both the quality deviation and the labour costs, optimise the algorithm to reduce the computational burden of the large-scale optimisation problem.

2.5 Industrial Evaporators

Evaporation, to explain more direct, is turning the liquid into a vapour by providing heat resources to achieve product concentration. A variety of evaporators and a wide range of industrial applications have been applied in many different fields, however, the majority of them are using in the food industry. Brennan et al. listed four main applications of evaporators in the food industry in 1990:

(1): Pre-concentration of a liquid, which is the most common application, to prepare for further processes, such as spray drying, crystallisation and so on.

(2): Reduce the liquid volume by removing the water to reduce the costs of transportation or storage.

(3): To remove water in certain foods can minimize the ‘water activity’, for example, bacterial growth, to protect the foods.

(4): For the utilisation and reduction of effluent.

2.5.1 Types of evaporators

Nowadays, the applications of the evaporator are intensively found in milk powder, fruit juice, salt, and sugar processing industry. According to different purposes, there are various evaporators have been developed to deal with all kinds of products. In the 19th Century, a vacuum with a steam-heated evaporator, which the vacuum operation environment reduces the boiling temperature, is developed by Kroschwitz and Howe-Grant in the sugar industry. Then the first multi-effect evaporator is built in the late 19th Century to improve energy efficiency in the same industry.

Different application has specific requirements for the evaporator about the design and structures, but several must-have components, which are heat exchangers, separators, a condenser, and the preheater system, are included in a general industrial evaporation process (Brennan et al, 1990).

Heat exchangers to supply the necessary sensible and latent heat to the liquid is usually via saturated steam, sometimes hot water and other thermal fluids can be replaced.

A separator is a place where the vapour and liquid phases are separated.

The vapour is condensed and removed from the system by a condenser. Also, it can create and maintain the partial vacuum environment of the evaporator.

A preheater system is for heating the product to the required boiling temperature before feeding into a heat exchanger system.

Evaporators can be classified according to the size and shape of the heating surfaces and mechanisms used to pass the product over there surfaces. The following types of evaporators are most popular in the food industry.

2.5.2 Vacuum Pans

The first vacuum pan was introduced by Edward Charles Howard in 1813 and was very similar to those which are used in industry today. Vacuum pan is one of the simplest types of vacuum evaporators mostly applied in sugar, jam and sauces manufacture. The main body of a vacuum pan is a large vessel with a heating resource, such as a steam jacket, which connects to a vacuum pump or system (refer to figure 2.5). Because of the relatively simple structure and the favourable heat transfer area to volume ratio, it is more suitable for frequent changes of product and low or variable throughputs. However, the vacuum pans have a small heat transfer area per unit volume, so it takes more running time to reach the desired solids content target.

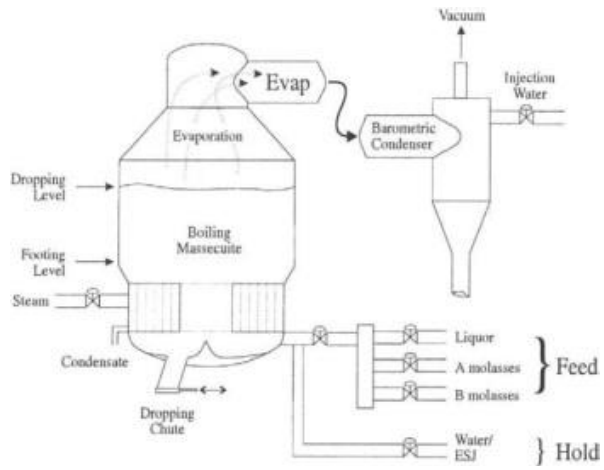


Figure 2.5 Vacuum pan diagram (Vu and Schneider, 2007)

2.5.3 Short Tube Vacuum Evaporators

This one of the oldest types of evaporators consists of a calendrical evaporation chamber surrounded by vertical tubes located near the bottom of the chamber (see Fig. 2.6). The tubes' size could differ from 25-75 mm in diameter and 0.5-2.0 m in length due to its required tasks. The steam flows through those tubes to heat the liquid being feed into the chamber. When the liquid boils, vapour exits the chamber from the top and concentrated product flow down to the bottom through a large, cooler tube in the centre of the chamber.

Short tube evaporators do not have much headroom requirements of the chamber and are relatively cheaper than other evaporators. At the same time, it also can obtain better heat transfer coefficients when the system is in higher operating temperature. But, there is a high risk of fouling because of its relatively long residence time. So this evaporator is suitable for low viscosity liquids and not very heat-sensitive products, such as sugar-related industries. It is not suitable for the crystallisation process unless an impeller is equipped to keep the crystals in suspension.

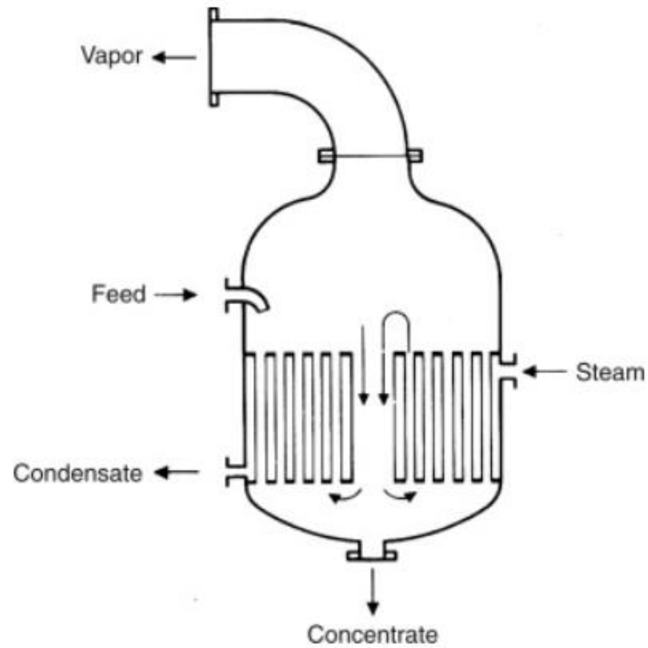


Figure 2.6 Short tube vacuum evaporator diagram (Berk, 2009)

2.5.4 Long Tube Vacuum Evaporators

Long tube evaporators, also known as film-type evaporators, are the most widely used in the food industry. It has bundles of long tubes, which are various from 3-20m in length and 25-50mm in diameter, inside a long vessel. There are two patterns of long tube evaporators, which are climbing (also called rising) film evaporators and falling film evaporators.

In a climbing film evaporator (see Fig. 2.7), the product feeds into the long tubes from the bottom of the large vessel and starts to be heated until reaching the boiling point. The heating medium is circulated inside the vessel but outside the tubes. When the liquid starts to boiling, vapour rises and expands rapidly in the centre of the tube, then a thin film of the concentrated liquid is formed on the tubes' inner wall. With the rising of the mixture of liquid and vapour, it becomes more concentrated and then enters a separator at the top of the tubes. The vapour could be pump into a heating jacket or another vessel if the evaporators have more than one effect, or drawn off to a condenser to remove. The liquid is collected as the product, or fed to another effect in a multiple effects system.

Falling film evaporators (see Fig. 2.8), in an easy way to explain, are an upside-down version of the climbing film evaporators. The feed point is at the top of the evaporator and the separator is at the bottom attached to the main vessel. Some of the falling film evaporators may have a distribution device on the top to distribute the liquid to each tube.

Because of the large heating surfaces, both climbing film and falling film evaporators are able to operate small temperature differences between the heating medium and the liquid, which means these types of the evaporator can cope with viscous materials, such as milk and fruit juice.

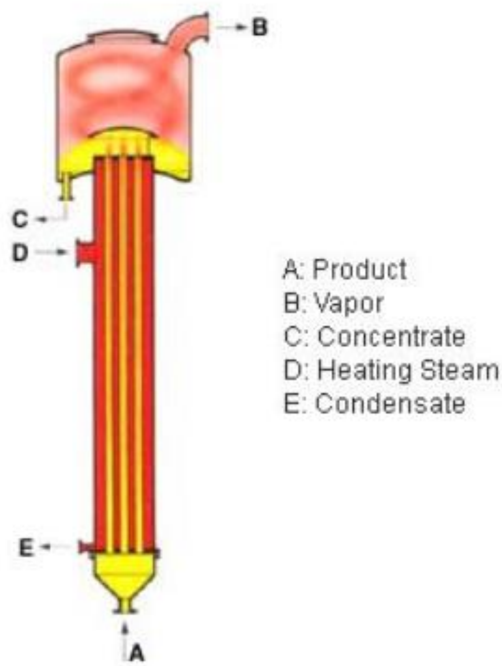


Figure 2.7 Climbing film evaporator diagram (Visual Encyclopedia of Chemical Engineering)

Additionally, their short residence times, especially the falling film evaporator, are ideal for heat-sensitive products like orange juice. As a consequence, a falling film evaporator is an exclusive option in almost all the food industries. However, not like short tube evaporators, a large amount of headroom is required and long tube evaporators are generally not suitable for salting or scaling materials.

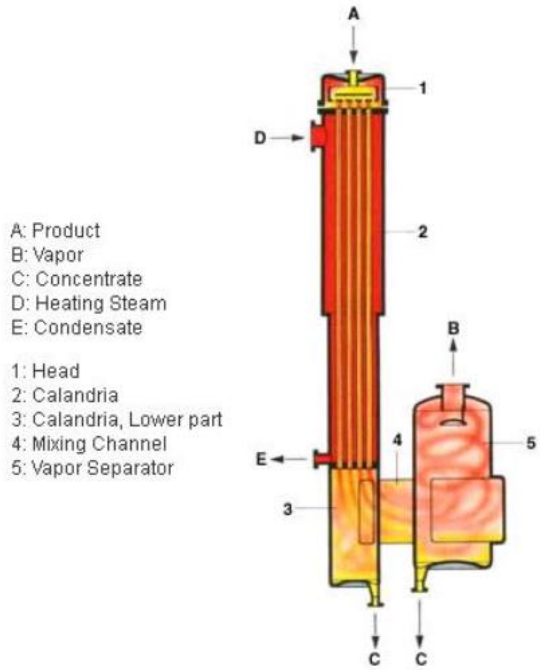


Figure 2.8 Falling film evaporator diagram (Visual Encyclopedia of Chemical Engineering)

2.5.5 Forced circulation evaporators

A forced circulation evaporator includes a heat exchanger and a separate chamber (see Fig. 2.9). A pump or a condenser is used to ensure that the feed product can past the heat exchanger and the separator to complete circulation in the evaporator. This type of evaporator is developed for the materials which are susceptible to scaling and crystallizing because of the high circulation rate. Meanwhile, the hydrostatic head at the top of the heat exchanger prevents boiling. When the liquid is pumped into the separator, it 'flash' into vapour due to the slightly lower pressure environment inside the separator. The vapour exits the system at the top of the separator and concentrated product is collected at the bottom of the separator. For the heat exchanger part, it can either be horizontal or vertical, depending on the available headroom, it also can be multiple effects rather than just single. Like the film-type evaporator mentioned above, forced circulation evaporator obtains high transfer coefficients even with viscous solutions and low residence times so that it can deal with heat-sensitive materials. Also, because of the high circulation rate, the scale formation on the heating surfaces is highly prevented. It is suitable for concentrating sugar solutions, glucose, salt and more heat-sensitive liquid, such as milk, juices, and meat extracts. However, the very high costs for the pumps operating and maintenance makes this evaporator less economical than others.

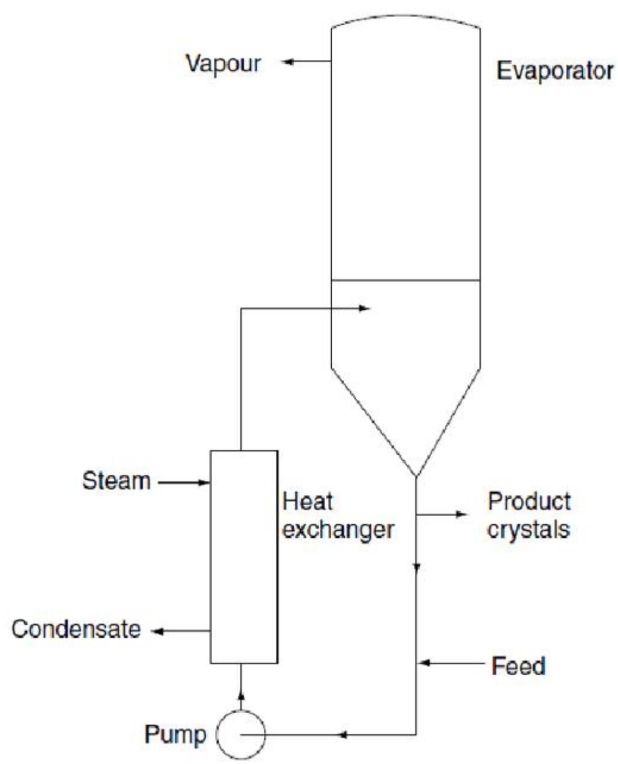


Figure 2.9 Forced circulation evaporator diagram (CHEM Process System)

2.5.6 Horizontal tube evaporators

This kind of evaporators have very simple units and are used for small quantity evaporation requirement. Many horizontal tubes located at the bottom of the evaporation chamber, where the hot steam flows through the inside of the tubes to heat the liquid. The liquid normally is pumped and sprayed above the tubes from the top position in the chamber (see Fig. 2.10). The main advantages of this evaporator are that the headroom requirement is very little, and it is one of the most economical evaporators provided good heat transfer. However, horizontal tube evaporators are not suitable for severely scaling or salting liquid.

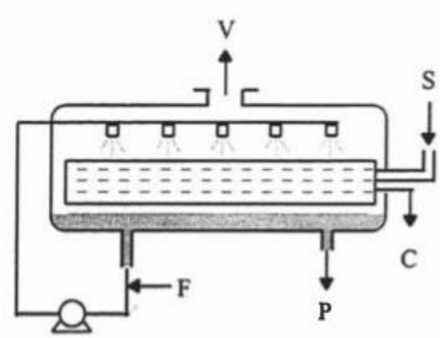


Figure 2.10 Horizontal tube evaporator diagram (Russel, 1997)

2.5.7 Other evaporators

There are several other types of evaporators used in different industries which include:

2.5.7.1 Plate evaporators

These evaporators are more like heat exchangers which consist of plenty of parallel gasketed plates connected one above the others to allow the formation of a series of channels for fluids and vapour to flow between them. The heat medium of this evaporator is the metal plates. When the liquid is pumped to pass the plates usually from bottom to the top on one side, and the steam is circulated from top to bottom on the other side (see Fig. 2.11). Heat exchange happens between the plates where the liquid is evaporated. The advantages of these evaporators are very obvious that is the large heat transfer areas and the high heat transfer coefficients which are higher than the most of the tubular evaporators. The metal plates are easy to remove or add which makes the capacity more flexible. But the fluid leakage between plates is a challenge for industry applications and the installation cost is relatively high.

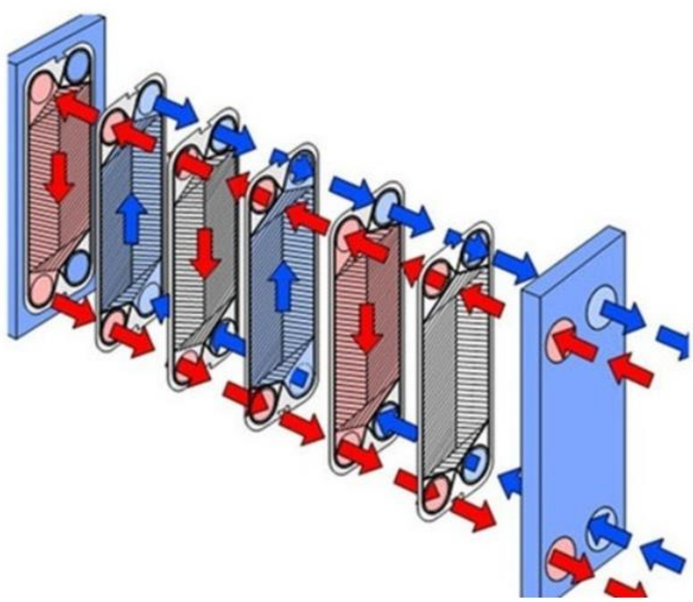


Figure 2.11 Plate evaporator working principle diagram (Joshi, 2016)

2.5.7.2 Centrifugal Evaporators

Centrifugal evaporators, which are also known as ‘centri-therm’ or ‘thin-film’ evaporators, consist of a rotating stack of cones housed in a stationary shell. The thin film is formed and spread quickly over the heating surfaces in each cone due to the centrifugal force when the liquid product is fed. When the heating steam flows into the cones, the evaporation process occurs rapidly. Very high heat transfer rates and short residence times are the main benefits of this evaporator, but high costs and low throughputs are the limitations. Suitable for heat-sensitive and viscous materials, such as fruit and vegetable juices, purees and extracts of coffee and tea.

2.5.7.3 Agitated thin-film evaporators

These evaporators are designed for very viscous materials and the materials which are easy to foul. It is a modified falling film evaporator with a steam jacketed shell contains an internal agitator used for wiping the inner surface in order to form a thin film. The distance between the agitator and the inner surface could be fixed of 0.5-2.0 mm according to the requirement of the film thin level. Similar to falling film evaporators, concentrated liquid flows down to the bottom due to gravity, and the vapour escapes the system through outlets at the top of the tube. Agitated thin film evaporators are the most expensive of all evaporators because of its high operating and maintenance costs. Also, low heat transfer coefficients are another limitation of industrial applications.

2.5.8 Multi-effect evaporation

With increasing competition, only a single effect evaporator cannot be satisfied in the industrial production. The vapour, which contains considerable energy, usually exits the system directly or passes through a condenser and removed from the system in a single effect evaporator. If it could be re-used in another effect to evaporate products, the system efficiency could be increased significantly. The first application of a multi-effect evaporator was invented by Norbert Rillieux in 1846 for a sugar cane plant. Vapour was pumped out from the first pan to heat the sugar cane juice in the next, and into a condenser in the last effect. Pressure in the whole system was reduced from effect to effect in order to lower the boiling temperature of the liquid. The results showed a better quality of sugar with less manpower and lower cost. However, the multi-effect evaporators were not widely used until the 1920s. Between 1920 and 1940, applications of double-effect evaporators have dramatically increased in many industrial manufactures. With the development and improvement of materials and technologies, 6 and 7 effects evaporators were implemented successfully during the 1970s.

The reason for developing a multi-effect evaporator is to achieve a higher energy economy and efficiency, not for the evaporation capacity. Hall et al (1986) introduced a concept named specific steam consumption, which indicates how much steam consumption can evaporate 1kg water. In a single effect evaporator, the number is approximately 1.1-1.3kg, and with more effects added to the evaporation process, this number decreases (see Table 2.1). However, the capital cost of installing extra effects must be considered. An extra same scale effect is normally cost equivalent to a single effect evaporator. Thus, to build a 7 effects evaporator costs approximately 7 times than to build a single effect one. That is why three to five effects are commonly used in industry today.

There are four feed flow patterns of a multi-effect evaporator: forward feed, backward feed, parallel feed, and mixed feed.

Table 2.1 Average steam consumption to evaporate 1 kg of water in an evaporator

Number of effects	1	2	3	4	5	6	7
Specific steam consumption (kg)	1.2	0.65	0.38	0.29	0.23	0.19	0.17

2.5.8.1 Forward Feed

The forward feed of a multiple-effect evaporator is the simplest and most commonly used feeding method in the food industry. Both product and steam are fed into the first effect in the same direction, then to the next effect (see Fig. 2.12).

The liquid is getting more and more concentrated from the first effect to the last. An extraction pumped is needed to take out the concentrated product from the last effect. However, because the system pressure decreases from the first effect (at atmospheric pressure or a little lower), the liquid can move between effects without pumps. This could reduce installation costs. Another benefit of this feed method is that products are collected at a low temperature in the last effect. Thus, heat damage to the products can be minimized. But the overall economy is lower than other multi-effect evaporators if the fed products are cold.

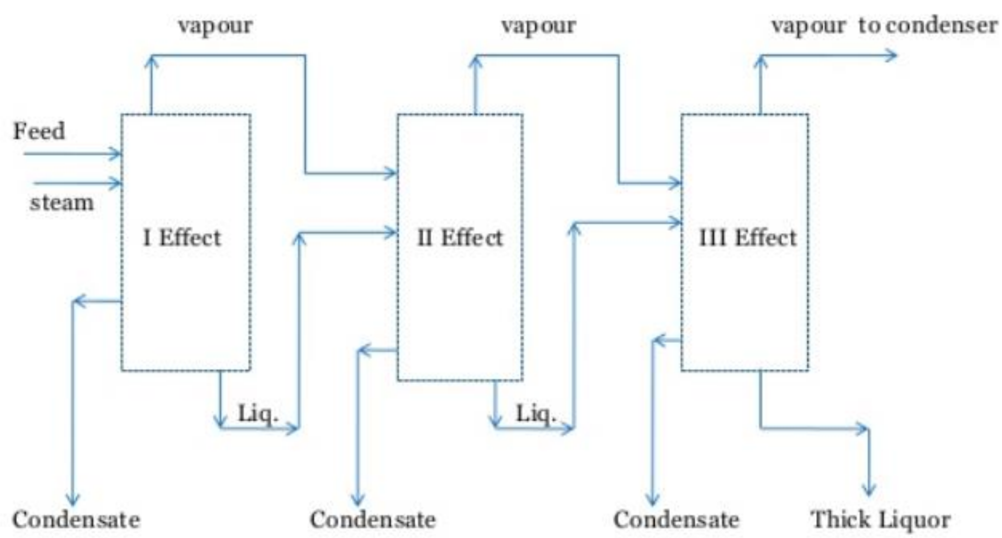


Figure 2.12 Simplified forward feed three effects diagram (Shekhar, 2017)

2.5.8.2 Backward Feed

In a backward feed evaporator, it is obvious that the product is fed from the last effect and passes through the effects in the opposite direction to the steam (see

Fig. 2.13). The hot steam is supplied to the first effect where the product is evaporated and concentrated in the hottest environment which improves the steam economy. It is more suitable for viscous liquid comparing with forward feed method. The only disadvantage is that more pumps are needed between effects due to the flow from low to high pressure.

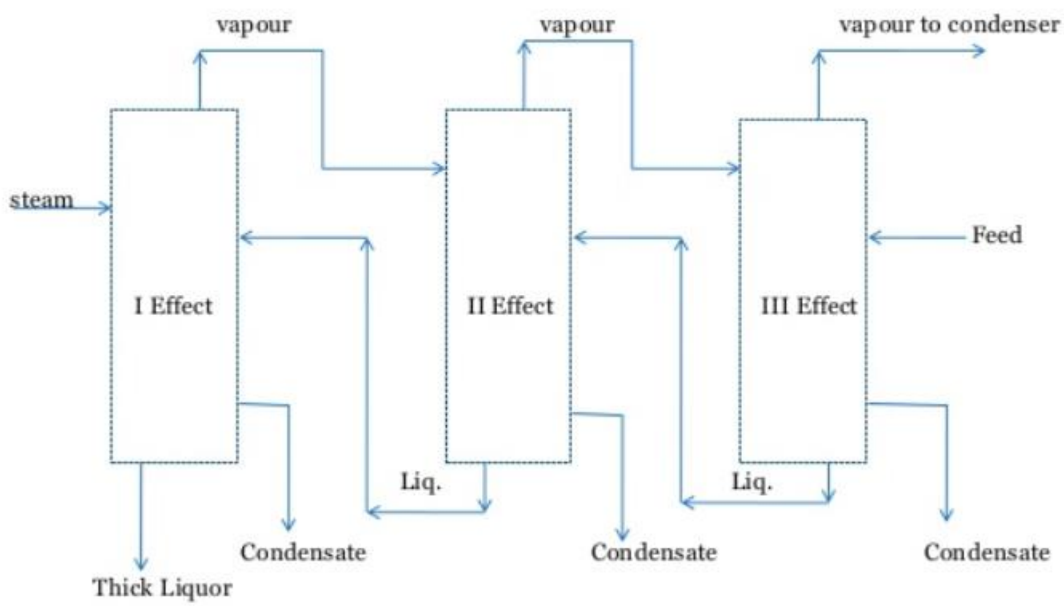


Figure 2.13 Simplified backward feed three effects evaporator diagram (Shekhar, 2017)

2.5.8.3 Parallel Feed

The parallel feed is another popular evaporator arrangement mostly applied in the crystallisation process. As its name, the product is fed into each effect individually as parallel lines. Same as other feed methods, the vapour is pumped into the first effect and used to heat the following effects (see Fig. 2.14).

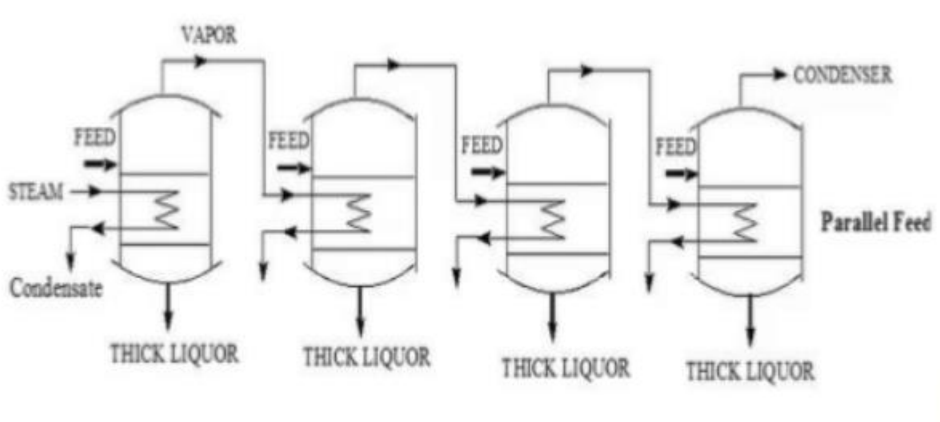


Figure 2.14 Simplified parallel feed three effects evaporator diagram (Arora, 2017)

2.5.8.4 Mixed Feed

The mixed feed is generally a combination of forward and backward feed. It represents the advantages of the two feed methods, which are simplicity of the forward feed and the higher steam economy of backward feed. In a mixed feed evaporation system, the product is fed into a middle effect and flows like a forward feed to the last effect, and then pumped back to the first effect like the backward feed to complete the final concentration step. These feed pattern systems usually contain plenty of effects.

2.6 Vapour Recompression

It is generally known that the external steam supply, also known as the direct heating medium, is used for small-capacity evaporators to support the heating steam. However, if the evaporation system is large, only direct heating is not efficient and economical. Therefore, more applicable techniques of using steam to heat the effect are introduced, named vapour recompression.

Vapour recompression is a procedure that partial or all vapour from the separator or evaporation tubes are compressed and pumped back to the steam jacket or the same effect to use as a recycle heating medium. It can not only reduce the requirement for the direct heating amount but also reduce the specific steam consumption. There are two methods of performing vapour recompression.

2.6.1 Thermal vapour recompression (TVR)

In an evaporation system with the TVR, partial of the vapour from the separator is fed into the next effect. The remainder passes through a steam-jet ejector in order to increase the pressure and then enters the jacket of the evaporation chamber. The main advantages of TVR are relatively simple, inexpensive and long working life.

2.6.2 Mechanical vapour recompression (MVR)

MVR is driven by the mechanical power, such as electricity, motor engine or steam turbine, to compress all the exhaust vapour from the separator. Then the compressed vapour from an effect is returned to the system to re-use. For the whole system working stable, some of the exhaust vapour needs to be condensed by the cooling water. Comparing with TVR, MVR has the lowest operating costs and the highest capital and maintenance costs.

2.7 Dairy industry multi-effect falling film evaporators

Nowadays, the falling-film evaporator, especially the multiple effects one, is the most widely used in the dairy industry due to the following features:

- Short residence time due to the high flow velocities and the thin film formed inner the tube surface suitable for heat-sensitive products;
- The falling film evaporator can be operated at a low-temperature difference because of the large heat transfer surface which can be equipped with many effects;
- The contracture is relatively simple and the capacity is flexible;
- High energy efficiency due to the high operation heat transfer coefficients.

Multiple effects falling film evaporator is generally applied in order to improve the thermal efficiency and reduce steam consumption by allowing the vapour from the first effect to be re-used for heating the following effects. In the recent dairy industry, between two and seven effects are commonly used. A TVR or an MVR is normally included in a multi-effect falling film evaporation system to achieve the vapour recycle.

The major element of a falling film evaporator effect is the heat exchanger chamber, which normally consists of a bundle of tubes with heating jacket (calandria) surrounded, a distribution device at the top of the effect and a separator at the bottom attached the calandria (see Fig. 2.15).

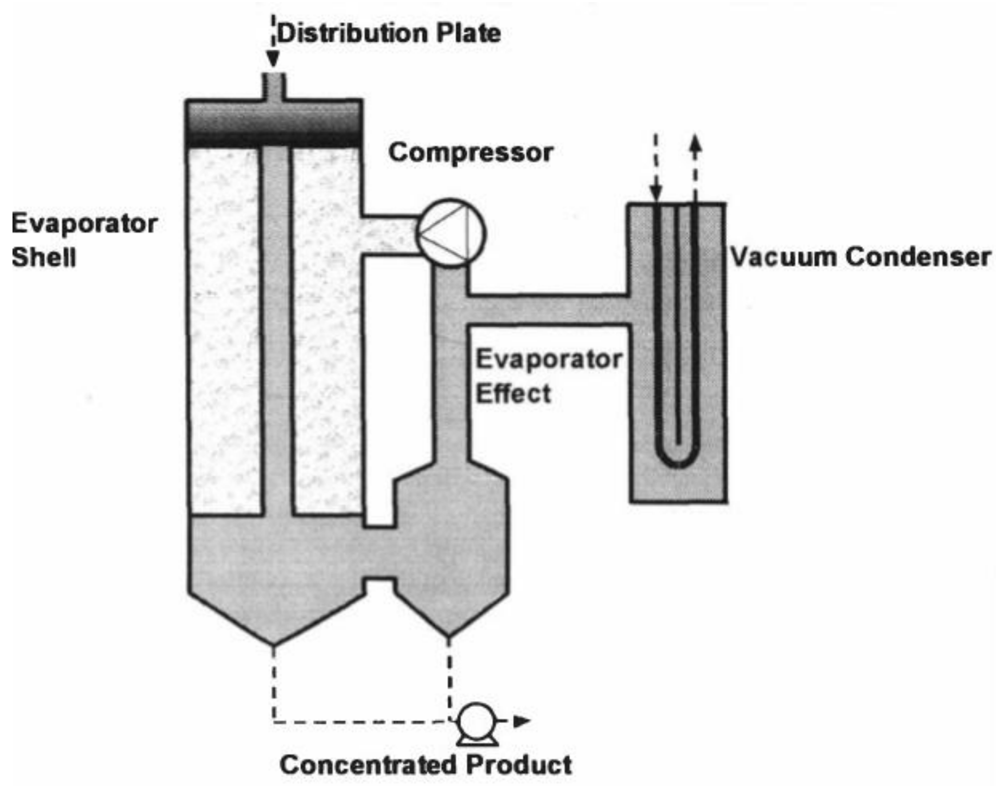


Figure 2.15 A typical single effect falling film evaporator (Winchester, 2000)

2.7.1 Distribution Device

As mentioned above, the distribution plate located at the top of each effect is to allow the liquid to flow into each tube evenly. There are two types of distribution applied, which are dynamic and static, to complete the distribution step.

The following figure 2.16 shows a dynamic distribution. A nozzle spray is used to atomise the liquid product into the evaporation tubes. This type of distributions normally has a simple structure and cheap to install. But the atomised liquid spraying to the tubes cannot be guaranteed as an even coverage. Meanwhile, in order to ensure the uniform spray pattern, relatively steady feed flow and operating pressure are required which increases the difficulties of control.

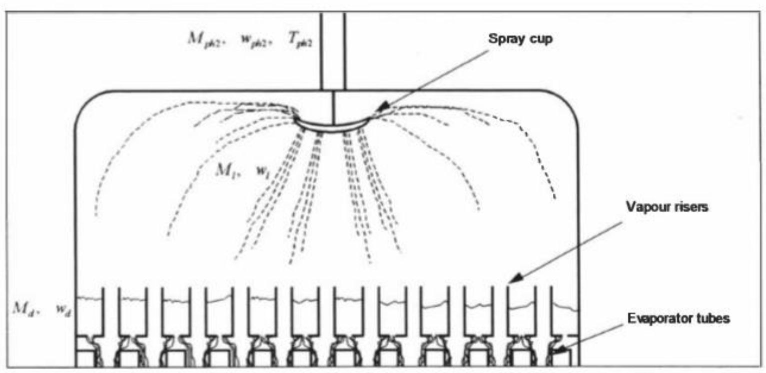


Figure 2.16 Dynamic distribution plate (Winchester, 2000)

In the dairy industry, static distribution is more selected, although it is more expensive and complex compared with a dynamic one. The main advantage of static distribution is that the inlet liquid can be evenly fed into each tube due to the same pressure provided by the level formed on the distribution plate over the tube holes (Figure 2.17). But, the plate holes size is designed according to the specific flowrates to the effect, therefore, the liquid level can be different by changing the feed flowrate which may cause overflow or emptied.

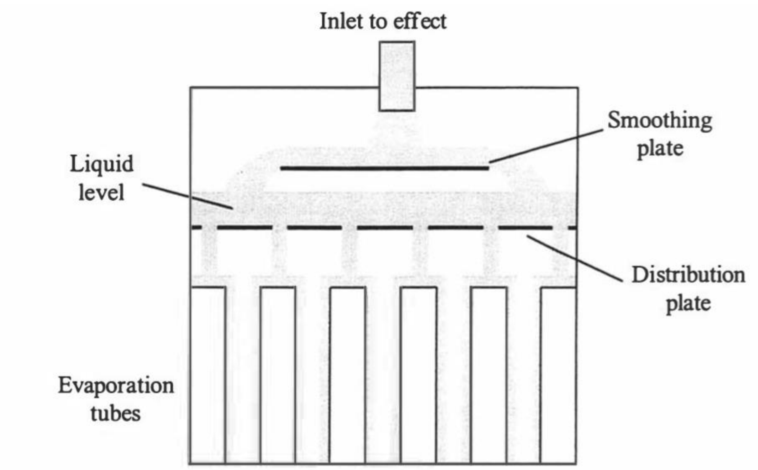


Figure 2.17 Static distribution device (Russel, 1997)

2.7.2 Evaporation Calandria

Once the liquid passes the distribution and flows down to the evaporation tubes, a thin film is formed in the inner surface of the tubes. The product is heated by a heating medium (steam either from the ejector or other effects) on the outside of the tubes. But how to obtain uniform liquid coverage of the product over the inner surface tubes is the biggest obstacle. The thin film can either break near the bottom of the tubes without sufficient flow or even not be formed at the top of the tubes. Furthermore, fouling is another common problem for the evaporation process, in which the tubes are clogged with a deposit because of the increasing downtime.

In order to avoid the problems mentioned above, specific film thickness must be controlled under a certain flowrate according to the length and diameter of the tubes in the real industrial plant. To recirculate the product can control the liquid coverage, but the residence time is increased dramatically at the same time which could higher the chance to damage the product.

The velocity of the product and vapour flow has an important correlation with the operating pressure. The operating pressure drops once the very high-speed liquid flows across a long, narrow tube. It can both increase the boiling temperature and reduce the heat transfer rate. If the tubes are designed shorter with a large diameter to improve the velocity problems, the thin film is more difficult to fully cover the inner surface. As a consequence, it is a compromise between the liquid coverage and pressure loss when designing the Evaporation Calandria. Hahn (1986) has indicated that the length of the tubes between 8 and 12 metres with 30 to 50 mm diameters are appropriate dimensions for an effect.

2.7.3 Separators

After evaporating in the Calandria, the liquid-vapour concentrate mixture flows into a separator where the separation occurs. Liquid drops to the bottom of the separator, collected or pumped to the next effect. The vapour exits the system from the top or compressed and pumped to the effects to re-use. Gravity, baffle and centrifugal are three methods of separation. In the dairy industry, the centrifugal separators are the most commonly used.

2.7.4 Evaporation Process Operation

In the dairy industry, falling film evaporators are applied predominantly as a pre-concentrate step before feeding into a spray dryer to produce milk powder. The spray drying is a huge energy consumption process. By removing the same amount of water, it consumes approximately 5 to 10 times energy more than the evaporation process (Hall et al, 1986). For energy saving purposes, water needs to be removed from the liquid as much as possible in the evaporation process. However, the more water is removed, the higher viscosity the product has. More energy is required for the spray nozzle to overcome the large viscous forces. Therefore, in the real manufacture, 40%-55% of total solid content (concentration) after the evaporation process is generally accepted. The final designed concentration of the milk collected from the third effect in this thesis is 52%.

2.8 Modelling of falling-film evaporators

As mentioned above that falling film evaporators are the first choice for most of the dairy industries to pre-concentrate the liquid for further processes. But the literature of falling film evaporators modelling is very limited. For control purposes, an accurate simulation model of the evaporation process is essential. Food production processes present great challenges to the engineers because it contains many characteristics which make the process particularly difficult to model and control.

Basically, there are two groups of models, which are physical or mathematical models and black box (empirical) models.

Physical models are very commonly used because physical phenomena can be obtained and studied during the evaporation process. Additionally, variables can be calculated and controlled based on physical principles. However, due to the complex process, good knowledge of physics is required, such as heat transfer coefficients, physical properties of the liquid, even surface tension and spray angle, etc. Because a few physical variables are not easy to measure during the manufacturing process, such as the thickness of the thin film, and some of the physical principles are very complicated. Thus, assumptions and simplifications are always applied when physical models are developed.

The black-box model, also known as the empirical model, is developed by using identification techniques to investigate the relations between inputs and outputs parameters. Comparing with physical models, black-box models are more relay on the data accuracy rather than the physical principles. The performances are often particularly accurate in a limited range around a specific operating point. However, this type of models are very sensitive to the product properties and the process conditions.

In the manufacturing industry, a falling film evaporator with two to seven effects is mostly used. Several researches have described the modelling of falling film evaporators from single-effect to multi-effect. A single effect with four-stage calandria and recompression unit falling film evaporator is developed by Choudhary et al in 2018 for the milk concentration process. 9% and 13% of total solids were being concentrated at 60℃ and 70℃ receptively. The final product concentration was controlled between 25-35%. A double-effect was developed even earlier by Runyon et al (1991) for commercial tomato paste concentration manufacture.

Three to five effects falling film evaporators are very popular for concentrating the liquid material in the food industry and many studies have been published (Zain & Kumar, 1996; Agarwal, 1992; Agarwal et al, 2004; Munir et al, 2014; Zhang et al, 2018; Medhat et al, 2015). Quaak et al (1994) developed a four effect falling film evaporator and an empirical model that are both used to develop a multivariable control system for the evaporation process.

The purposes of developing more than two effects falling film evaporator of these kinds of literature are mostly for improving the system performances, introducing a new control strategy, comparing the different controllers, etc.

Not only in the dairy industry, but falling film evaporators have also been modelled for non-dairy products, such as juice, seawater, and paper industry (Angeletti & Moresi, 1983; Chun & Seban, 1972; Murthy & Sarma, 1977; Kumar et al, 2012; Schwaer et al, 2020; Nasr and Rahaei, 2019;). Kumar et al (2012) have developed an unsteady-state multi-effect evaporator model with mixed feed in the paper industry to study the dynamic behaviours.

Winchester (2000) developed a multi-effect falling film evaporator for the milk powder production process with a direct steam injection unit.

The evaporation process is difficult to model in many aspects. Being a complex system, a large number of physical, chemical phenomena occur inside and outside the evaporator devices introduced above. One of the most challenges of evaporator modelling is the description of the distribution device of the falling film. There are a few studies have been working on the characters of the distribution process in the falling film evaporator, but not specifically for milk products (Bui & Dhir, 1985; Van der Mast et al, 1976).

Another important variable during the evaporation process is the heat transfer coefficients. A large number of studies are focusing on these variables to improve system efficiency (Benjamin 1957; Dukler et al, 1952; Fulford, 1964; ; Hu et al, 2019; Fang et al, 2019;). More recently, the empirical model has been developed to investigate the heat transfer coefficients (Chun and Seban, 1971; Alhusseini et al, 1998). However, very few of these works were for milk production. Some studies have focused on milk solutions, but the work is purely empirical and suffers from a lack of ‘good’ theoretical groundwork (Bouman et al, 1993; Jebson & Lyer, 1991).

For the dynamic studies of evaporators, Burdett (1971) developed a dynamic model for a falling film evaporator for the seawater evaporation process. But seawater evaporators are quite different from the typical dairy evaporators. Until 1990, Quaak and Gerritsen have developed the dynamic model of falling film evaporators for the milk solution concentration. The distributed parameter is not neglected in their model. Constant falling film velocity and constant evaporating heat transfer coefficients are assumed when modelling the system, but they transformed the partial differential equations with delays.

2.9 Control of Industrial Evaporators

Evaporation is the second most energy-intensive process in milk powder production after drying. Therefore, energy consumption during evaporation has a substantial impact on the cost of milk powder production (Munir et al, 2014). In milk powder production, the water needs to be removed from the evaporation and the drying process is 80-90% and 10-20% respectively. A steady input to the drying process is very important, which means to achieve good control in the evaporation stage is necessary.

Stapper (1979) presented two control objectives in order to ensure the downstream process, which has a huge influence on the following drying process, in the calandria.

The concentrated out-flowing product needs to be regulated accurately.

The product flowrate through the process needs to be controlled at the desired level.

Both objectives are aiming to maintain the evaporation output within a steady range. On the other hand, energy consumption and product quality must be considered while achieving the control targets. It is discussed above that improving the energy efficiency of the evaporation process is a key requirement in the commercial manufactory. This could be achieved by re-using the heat steam. Furthermore, the vacuum operation environment can reduce the amount of steam for the evaporation process.

Another factor which has to be guaranteed is product quality. In falling film evaporators, most of the product damaging is because of the over-heating. Generally, to minimize the fluctuations of product flowrate and hot steam injection are two effective methods to avoid product overheating. Additionally, a well-designed system can provide a steady operation condition, which also can maintain product quality.

Russell (1997) presented two manipulated input variables, which are steam flowrate and the cooling water flowrate, to regulate the product concentration. But the long-time delay response between the final product concentration and the input steam flowrate must be considered.

In conclusion, in order to achieve the process control, Russell listed three main manipulated variables:

- Steam flow
- Feed flow
- Cooling water flow

However, many interactions can happen in the evaporation process. Russell has summarized as follows:

1. Feed flow or feed temperature change: the temperature difference between steam and product in an effect is usually below 10℃, even only 0.1℃ changes could affect 1% evaporation rate.
2. Pressure change: the operation pressure has strong influences on the product and vapor feed into or exit the system.

3. Cooling water flow in the condenser change: this can cause the vacuum change and affect the operating temperature
4. Feed concentration change: this can be treated as a disturbance, different feed concentration requires a different amount of steam to heat
5. Steam pressure and quality change: another disturbance to the system which can influence the evaporation rate

2.9.1 Falling film Evaporator Control Development

In real dairy manufactory, conventional PID (Proportional-Integral-Derivative) is a basic control strategy implemented to the falling film evaporator. Several studies focussing on PID have been performed. Winchester & Marsh (2000) have developed three core loops based on PID to control the effect temperature, product dry mass fraction and product flowrate of a falling film evaporator with mechanical vapour recompression. Cunningham et al (2006) presented dual PID to control both the water level in the steam generator and the external steam amount supply in two effects falling film evaporator for dairy products. But the limitation of their work is the model is linear. More recently, Dwi Argo et al (2015) applied Particle Swarm Optimization (PSO) by adding a weighting factor of inertia to the optimisation of the PID controller for a multi-effect falling film evaporator.

Many PID-type controllers have been studied in the last two decades. Cascade control is a typical one. Karimi and Jahanmiri (2006) described a two loops cascade control algorithm based on the PID control strategy and applied it in a three effect falling film evaporator for the milk powder production process. Results showed that cascade control has a good performance on the disturbances rejection. In the same year, Bakker et al applied cascade control in a two effect falling film evaporator and found the same conclusion. A few years later, Farsi and Jahanmiri improved the cascade control with triple loops for the same three effect falling film evaporators in 2009. Comparing with two loops cascade, the triple one can reduce the overshoot from 4.8% to 1.04%, and setting time from 505 to 287 seconds.

Nowadays, only PID-type controllers cannot be satisfactory due to the increasing competition for commercial purposes. Especially for the large capacities industrial plant. Therefore, more and more researchers are focusing on advanced control strategies, such as Model Predictive Control (MPC), Fuzzy Logic Control (FLC) and Neural Network Control and so on.

MPC has been extensively applied in many industrial fields especially in powder, chemical, and refining plants because it can provide a better solution to improve the control performance for optimum process operation (Fahmy, et al, 2015). There are many successful applications of MPC in the non-dairy evaporation process.

A very early predictive control study on the multi-effect evaporation system was in 1986 by Ricker et al. They presented a multivariable predictive control strategy in a six-effect falling film evaporator for the pulp mill production process. A linear ‘internal model’ in the impulse-response form was used to find the relationship between inputs and outputs in order to achieve prediction targets. Suarez et al (2009) described an MPC based on a neural network model, which is obtained from a phenomenological simulation model validated with industrial data, for a 4-

effect evaporator in the grape juice industry. In this system, the controller has adequately responded when there are modelling errors, such as different feed properties of grape juice. Similarly, Atuonwu et al (2009) developed an industrial-scale five effects evaporator with nonlinear model predictive control based on neural network. Levenberg-Marquardt algorithm with automatic differentiation was used training input and output data obtained from the system identification experiments.   Ahmed et al (2012) proposed an MPC based on a Takagi-Sugeno (TS) fuzzy model and applied successfully in a single evaporator. The fuzzy algorithm can provide an adequate and accurate model even using only the measurable input and output information. Huusom and Jorgensen (2014) developed an offset-free MPC based on auto-regressive models. However, the MPC application of the evaporation process in the dairy industry is very limited. A nonlinear neural network model predictive control (NNMPC) is investigated to control a three-effect falling film evaporator in the dairy industry (Ahmed et al, 2017). The whole evaporation process was developed by using MATLAB/Simulink. The results indicated that NNMPC can achieve the control targets of the evaporation process, especially when there are disturbances in the system.

Apart from MPC, fuzzy logic is another advanced control strategy that has many successful applications for the evaporation process. Raghul et al (2012) have implemented a fuzzy logic controller to a three-stage evaporation process in the sugar industry. Results show a significant improvement with the fuzzy logic control strategy comparing with PID control. Another application of fuzzy logic in the sugar industry was done by Pitteea et al in 2004. They applied genetic algorithms to tune the scaling factors and membership functions of the fuzzy logic controller automatically to achieve the optimization of this control strategy. Recently, George and Kyatanavar (2016) optimized the fuzzy logic control with integrated Taguchi technique to reduce energy consumption and enhance the product quality in the sugar evaporation industry. However, the applications of fuzzy logic in the dairy industry are very rare. A fuzzy logic approach has been indicated by Perrot et al (2004) to help the operators to evaluate the cheese ripening degree during the manufacturing process.

In summary, multiple effects of falling film evaporators have been used in food manufacturing extensively for the product evaporation process. Meanwhile, there are many control strategies have implemented successfully to achieve specific control targets. For example, the PID loops (cascade) control can reject the system disturbances effectively. MPC is capable of handling the process with multiple input/output variables. However, the advanced control methods applications in the falling film evaporation process are only a few. It is a potential process that needs efficient and reliable advanced controllers for system control and condition monitoring. The following chapters are aiming to address several problems in modelling, control, and results comparison of a three-effect falling film evaporation process in the milk powder production plant.

2.10 Summary

Due to the strong incentive from booming industries, advanced process control methods for the evaporation process has been receiving significant attention in recent years. Nevertheless, this is also an area in need of research to deliver an efficient and reliable method for control. The following chapters are aiming to develop the mathematical model, different control strategies of a three effect falling film evaporator, and performances with different controllers are going to be compared and discussed.

Chapter 3 Mathematical Modelling and Simulation of Three Effects Falling Film Evaporators

3.1 Introduction

Simulation is the imitating operation of the real process or system over time. It is a very helpful and valuable step before applying any control systems and optimisations to the real manufacturing plant in the food industry. Many benefits can be obtained by doing this, such as reducing chances of failure to meet specifications, eliminating potential system faults, lowering the costs, helping predict possible results and testing any new methodologies (Maria, 1997). From a control engineer's point of view, the main purpose of developing the system model is for designing the control system, which is commonly referred to as model-based control system design, in order to achieve control targets or optimize the system performances.

Mathematical model of multi-effect falling film evaporators which describe the interaction and correlations between variables, have been developed and studied throughout the years in the dairy industry. The basic requirement for the mathematical model is to copy the actual process behaviours as accurately as possible in order to measure the system outputs in response to the various input and disturbance variables. The model is based on energy and mass balances through each part of the system.

In this chapter, a three-effect falling film evaporator is used as an example to show how the mathematical model is developed in MATLAB/Simulink, and how the model can be used for different control strategies development. Also, model validation is presented against results from similar previous published studies.

3.2 Basic Mass and Energy Balance of evaporators

There are many different types of evaporators as mentioned in Chapter 2. However, the basic working principles are the same. The whole evaporation process complies with the law of conservation of mass and energy no matter if the evaporator is a single-effect or multi-effect.

3.2.1 Mass Balance

For a continuous evaporation process, the mass balance is represented by the following equations (Winchester, 2000):

$$M_p = M_{in} - M_e \tag{3.1}$$

Where, M_p = mass flow of product exiting the evaporator,

M_{in} = mass flow of product entering the evaporator,

M_e = mass flow of water evaporated (or removed) from the evaporator, This is a linear equation and suitable for all types of the evaporation process. Meanwhile, Eq. 3.1 is also a basic formula, in which most of the mass correlation equations of the evaporation process are based on it.

It is common knowledge that the main purpose of the evaporation process is to increase the total solid content from a lower level to a designed higher level. Therefore, the concentration of the final product leaving the evaporator is an important variable that needs to be controlled appropriately. Equation 3.2 (Winchester, 2000) indicates how the product concentration is calculated.

$$\omega_p = \frac{M_{in} * \omega_f}{(M_{in} - M_e)} \tag{3.2}$$

Where, ω_p = final product concentration,

ω_f = feed production concentration,

3.2.2 Energy Balance

According to the general energy balance equations given by Earle (1983):

Energy In = Energy Out + Energy Stored. As an evaporator, the energy transformation is between the hot steam and the product via a tube surface. So the heat energy balance is the priority principle that should be considered for the evaporation process.

The following formula is the heat balance for one effect:

$$Q_1 = U_1 * A_1(T_s - T_1) = U_1 * A_1 * \Delta T_1 \tag{3.3}$$

Where, Q_1 = Heat Transfer Rate of the first effect,

U_1 = The overall heat transfer coefficient of the first effect,

A_1 = Total Heat Transfer Surface Area,

T_s = Feed Steam Temperature to the first effect,

T_1 = Boiling Temperature of the liquid in the first effect,

ΔT_1 = Temperature Difference between the heating medium and the boiling liquid,

Similarly, for a multi-effect evaporator, the following effects are developed base on Eq. 3.3, but the steam fed is the vapour from its last effect. A better understanding of the above equation 3.3 is not only important for the whole evaporation plant design but also necessary to consider for the selection and operation of the evaporators. Base on the Eq. 3.3, it is obvious that the heat transfer coefficient can be improved by increasing the heat transfer surface, or the temperature differences with a certain heat flow feed.

However, larger heat transfer surface area evaporation tubes incur more capital costs of the system. On the contrary, a smaller heat transfer area requires a larger temperature difference, which could both increase the energy input and enhance the risk of heat-sensitive products being damaged. Thus, designers have to continually revisit this equation in order to optimise the system performance according to the plant size, physical properties of the product and the temperatures used.

3.3 Mathematical model development

3.3.1 Thermodynamics of the model

The thermodynamic behaviours of the process modelling are based on the fundamental First Law of Thermodynamics and the Law of Conservation of Mass (Energy and Mass balances) given by Sandler (1989, p30). The equation below shows the basic principle:

$$\sum Mass_{exit} = \sum Mass_{feed} - \sum Mass_{remain} - \sum Mass_{removed} \tag{3.4}$$

Where, $\sum Mass_{exit}$ = Total mass exiting the evaporation system.

$\sum Mass_{feed}$ = Total mass feeding into the evaporation system.

$\sum Mass_{remain}$ = Total mass remaining in the evaporation system.

$\sum Mass_{removed}$ = Total mass removed during the evaporation process.

As we know that Joule developed the First Law of Thermodynamics originally by experimenting in the 19th century. The following original equation (3.5) from the experiment shows the details (Winchester, 2000):

$$\frac{d}{dt}\left[M\left(U+\frac{1}{2}v^2+\varphi\right)\right] = \sum_{k=1}^{k} M_k\left(U_k+\frac{1}{2}v_k^2+\varphi_k\right) + q + W_s - P\frac{dV_T}{d_t} + \sum_{k=1}^{k} M_k(P_kV_k) \tag{3.5}$$

Where,

M = Total mass of material in the control volume. (kg)

U = Total internal energy of the control volume. (J.kg)

v = Velocity of the control volume. (J/kg)

φ = Potential energy of the control volume. (J/kg)

U_k = Internal energy of an input/output flow. (J/kg)

M_k = Mass flow of an input/output flow. (kg/s)

φ_k = Potential energy of an input/output flow. (J/kg)

v_k = Velocity of an input/output flow. (m/s)

q = Total net heat flow into the control volume. (W)

W_s = Total net work applied to the control volume. (W)

P_k = Pressure of an input/output flow. (Pa)

V_k = Specific volume of an input/output flow. (m^3/kg)

V_T = Total volume of the energy balance control volume. (m^3)

But in practice, there are many assumptions applied to simplify the equation. Firstly, enthalpy is normally used instead of internal energy. Secondly, some of the terms in the above equation can be assumed as constant or neglected, such as potential energy. According to these two assumptions, Eq. 3.5 can be simplified as follows (Winchester, 2000):

$$\frac{d[M*U]}{dt} = \sum_{k=1}^{k} M_k * H_k + q + W_s \quad (3.6)$$

If the density of the material is large, the variable enthalpy can be assumed equal to the internal energies (U=H). The First Law can be rewritten as:

$$\frac{d[M*H]}{dt} = \sum_{k=1}^{k} M_k * H_k + q + W_s \quad (3.7)$$

From the equation above, the enthalpy is a very important parameter. Sandler (1989) gave the differential equation of enthalpy as follows:

$$dH = C_p * dT + [V - T(\frac{\partial V}{\partial T})_p]dP \quad (3.8)$$

Where, C_p is the constant pressure heat capacity of material (J/kg.℃), *V* and *T* are the specific volumes (m^3/kg) and temperature (℃) of the material respectively, *P* is the pressure (Pa). Due to the relatively small value of the variables V and $\frac{\partial V}{\partial T}$ in Eq. 3.8, it can be simplified to the following:

$$\begin{cases} dH = C_p * dT \\ \Delta H = \int_{T1}^{T2} C_p * dT \end{cases} \quad (3.9)$$

The heat capacity of material always changes with different temperature conditions. However, if there is a very small temperature difference, to assume the heat capacity as a constant is acceptable. For milk, heat capacity depends on dry matter and composition. Sandler (1989) described the enthalpy equation for milk solution with the ideal mixture condition as follows:

$$\begin{cases} C_{p\_milk} = C_{p\_water} - C_{pTS} * w \\ H_{milk} = \left(C_{p\_water} - C_{pTS} * w\right) * T \end{cases} \quad (3.10)$$

Where, H_{milk} = Enthalpy of the liquid milk. (J/kg)

C_{p_milk} = Heat capacity of the milk solution (J/kg.℃)

C_{p_water} = Heat capacity of water (J/kg.℃)

C_{pTS} = Coefficient relating milk heat capacity and dry mass fraction (J/kg.℃)

w = Dry mass fraction of the milk. (kg/kg)

3.3.2 Model Structure and derivation

The three-effect falling film evaporator system approach of this thesis is referring to the previous work by Newell & Lee (1989) and Winchester (2000). The whole evaporation system was divided into two main sub-systems, which were contained in each effect of the evaporator. They are distribution plates and the evaporation sub-systems. The evaporation sub-system includes a heat exchanger, a separator, a condenser, and a steam ejector. Pre-heater is not introduced in this thesis and it is replaced by an input with disturbances generator.

3.3.2.1 Distribution plate

In a falling film evaporator, maintaining a uniform distribution of the product over the heating surface is very important during the evaporation process. That is why a liquid height above the distribution plate is required in order to buffer and reduce the feed flowrate disturbances. Besides, the distribution plate sub-system is normally assumed to be a perfectly mixed tank when developing the model. A standard orifice equation (Eq. 3.11) was given by Nevers to calculate the flow of liquid passing through the distribution plate by in 1991.

$$Q_d = \frac{C_d * A_h}{[1-\beta^2]}\sqrt{\frac{2*\Delta P}{\rho_d}} \quad (\Delta P = \rho_d * g * h_d) \tag{3.11}$$

Where, Q_d = Volumetric flow of liquid passing through the distribution plate. (m^3/s)

A_h = Surface area of holes in the distribution plate. (m^2)

ΔP = Pressure difference across the distribution plate. (Pa)

β = Ratio of areas, between a distribution plate hole and calandria.

C_d = Discharge coefficient for the distribution plate holes.

h_d = Height of liquid above the distribution plate. (m)

ρ_d = Density of liquid above the distribution plate. (Kg/m^3)

g = Acceleration due to gravity (m/s^2)

Because of the ratio of areas ($\beta = \frac{A_h}{A_c}$, where A_h is the distribution plate hole area and A_c is the calandria area) is a very small value, therefore, it can be neglected. So the equation 3.11 above can be simplified to the following:

$$Q_d = C_d A_h \sqrt{2 * g * h_d} \tag{3.12}$$

The liquid height above the distribution plate is another important variable to maintain the material fed into the effect tube more stable. It is given by the following differential equation (Newell and Lee, 1989).

$$\frac{dh_d(t)}{dt} = \frac{Q_f(t) - Q_d(t)}{A_d} \tag{3.13}$$

Where, Q_f is the product feed/flow into the distribution plate system and A_d is the liquid surface area.

The product concentration (total solid/dry matter) has a slight difference when it exits the distribution plate because the feed temperature is normally a little higher than the effect temperature which causes part of the water evaporated (Paramalingam, 2004). It can be calculated by the equation below:

$$\frac{dw_d}{dt} = \frac{\rho_{in}Q_f(t)w_{in}(t) - \rho_{out}Q_d(t)w_d(t)}{\rho_{out}(t)A_d h_d(t)} \tag{3.14}$$

Where, ρ_{in} and ρ_{out} are the product densities entering and exiting the distribution plate respectively, w_{in} is the input product concentration. However, due to the tiny concentration difference before and after the liquid passing through the distribution plate, in our study, product concentration in each effect is assumed as a constant.

3.3.2.2 Evaporation tube effect sub-system

It is generally believed that this is the key sub-system for an evaporation process. The liquid is heated by contacting the external hot steam over the tubes inside the effect; falling-film forms inside the surface of the tubes and the water starts to be evaporated from the product. Before modelling the tube effect sub-systems, two assumptions must be made to simplify the system due to the system's complexity and modelling difficulties. The first assumption is the velocity of the falling down liquid in the tube effect is treated as constant and uniform. Because it has been proved that the variety of feed flowrates have a small influence on the residence time through the effect tubes. The second is the heat flow across the tube assumed as a constant. It means a small heating temperature difference in the tube should not affect much of the results.

The residence time τ_e can be calculated by using the following equation (Russel, 1997):

$$\rho_e(t)Q_e(t) = \rho_d(t-\tau_e) * Q_d(t-\tau_e) - \frac{1}{\tau_e}\int_{t-\tau_e}^{t} \frac{q_{shell}(k)}{\gamma_e(k)} dk \qquad (3.15)$$

Where, Q_e is the volumetric product flow from the evaporation tube, q_{shell} is the heat supplied to the liquid and γ_e is the vaporisation latent heat of the liquid. The integral part in equation 3.15 is actually calculating the mass of the evaporated liquid within the residence time. It can be replaced by the total mass of vapoured M_v which is given as follows:

$$Q_e(t) = \frac{\rho_d(t-\tau_e)}{\rho_e(t)} Q_d(t-\tau_e) - \frac{1}{\rho_e(t)\tau_e} M_v(t) \qquad (3.16)$$

So, according to Eq. 3.15 and Eq. 3.16, M_v could be estimated by:

$$\frac{dM_v(t)}{dt} = \frac{q_{shell}(t)}{\gamma_e(t)} - \frac{q_{shell}(t-\tau_e)}{\gamma_e(t-\tau_e)} \qquad (3.17)$$

Base on the mass balance, the concentration of the product exiting the evaporation effect calculation is below:

$$w_e(t) = \frac{\rho_{out}(t-\tau_e)Q_d(t-\tau_e)w_d(t-\tau_e)}{\rho_e(t)Q_e(t)} \qquad (3.18)$$

As mentioned above, an evaporation effect sub-system includes a heat exchanger, a separator, a condenser, and a steam ejector. There are a few equations below (from Eq. 3.19 to Eq. 3.33) derived by Newell and Lee basically according to the mass and energy balance in 1989.

The mass balance on the solute in the heat exchanger can be referred to as the following equation:

$$M\frac{d_{w_e}}{dt} = F_{feed}w_{feed} - F_p w_p \tag{3.19}$$

Where, M is the total amount of liquid in the effect and is normally assumed to be a constant, F_{feed} and F_p are the feed and product flowrate in effect respectively, w_{feed} and w_p are feed and product concentration respectively.

The liquid and vapour mass balance can be described by the following two equations:

$$\rho A\frac{dL}{dt} = F_{feed} - F_v - F_p \tag{3.20}$$

$$C\frac{d_P}{d_t} = F_v - F_c \tag{3.21}$$

Where, ρ	= Product density	(kg/m^3)
L	= Liquid level in separator	(m)
A	= Cross-sectional area of the separator	(m^2)
F_v	= Vapour flowrate	(kg/min)
F_c	= Condensate flowrate	(kg/min)
P	= Operating pressure	(kPa)

C is a coefficient from the ideal gas law, which converts the mass of vapour into an equivalent pressure.

The liquid and vapour temperature can be calculated according to the liquid energy balance equations 3.22 and 3.23 below:

$$T_p = 0.5616P + 0.3126w_p + 48.43 \tag{3.22}$$

and

$$T_v = 0.507P + 55.0 \tag{3.23}$$

Where, T_p and T_v are the product and vapour temperature respectively, and both are linearisation of the saturated liquid line for water about the standard steady-state values (liquid is assumed to be ideally mixed).

Because the dynamics of energy balance are very fast so that:

$$F_v = (Q_H - F_{feed}C_p(T_p - T_f))/\lambda \tag{3.24}$$

Where, C_p is the heat capacity of the liquor and λ is the latent heat of vaporization of the liquor. The temperature difference between T_p and T_v is considered as a small value compared to the latent heat. In addition, heat losses to the process environment are neglected.

For the heat exchange evaporation process, the steam pressure is a variable that determines the steam temperature under saturated conditions. The correlation between steam temperature and steam pressure can be obtained by the following equation:

$$T_{steam} = 0.1538P_{steam} + 90.0 \tag{3.25}$$

Where, T_{steam} and P_{steam} stands for the steam temperature and pressure.

Meanwhile, the heating rate transfer to the boiling process liquid is given by:

$$Q_{HD} = UA_1(T_{steam} - T_p) \tag{3.26}$$

Where, Q_{HD} = Steam heat duty (kW)

U = Overall transfer coefficient

A_1 = Heat transfer area (m^2)

UA_1 can be described as a function of the total flowrate through the tubes in the effect:

$$UA_1 = 0.16(F_{feed} + F_{cir}) \tag{3.27}$$

F_{cir} is the circulating flowrate.

The steam flowrate is obtained from:

$$F_s = Q_{HD}/\lambda_s \tag{3.28}$$

λ_s is the latent heat of steam at the saturated conditions.

Cooling water is used as a coolant in the condenser. Its flowrate is a manipulated variable and temperature is a disturbance variable. The energy balance for cooling water can be described as:

$$Q_c = F_{cw}C_p'(T_{c\_out} - T_{c\_in}) \tag{3.29}$$

Where, Q_c = Condenser duty (kW)

F_{cw} = Cooling water flowrate (kg/min)

C_p' = Heat capacity of the cooling water (kg/min)

T_{c_out} = Cooling water outlet condenser temperature (℃)

T_{c_in} = Cooling water inlet condenser temperature (℃)

The heat transfer rate equation is:

$$Q_c = UA_2(T_v - 0.5\left(T_{c\_out} + T_{c\_in}\right)) \quad (3.30)$$

The above two equations 3.29 and 3.30 can be combined and eliminate T_{c_out} to get:

$$Q_c = \frac{UA_2(T_v - T_{c\_in})}{1 + \frac{UA_2}{2C_p' F_{cw}}} \quad (3.31)$$

And according to equation 3.29,

$$T_{c\_out} = T_{c\_in} + \frac{Q_c}{F_{cw} C_p'} \quad (3.32)$$

So that, the condensate flowrate is:

$$F_c = Q_c/\lambda_w \quad (3.33)$$

Where, λ_w is the latent heat of vaporization of water.

3.4 Summary of model variables and Initial Conditions

3.4.1 Evaporator variables

Table 3.1 lists the initial conditions of the four main input variables, which include the names, descriptions, and units as used in developing the mathematical model and simulation in the following sections. Disturbances are generated and added to the four main input variables to replicate the real industrial process. The values are maximum ±50% of the input initial values.

Table 3. 1 Main Input Variables

Variable	***Description***	***Units***
F_{feed}	Product feed flowrate	Kg/min
w_{feed}	Product feed concentration (total solid)	%
T_p	Product feed temperature	℃
T_{c_in}	Cooling water inlet condenser temperature	℃

There are also some additional inputs variables which are considered to be constants or setpoints for the evaporation process model as shown in Table 3.2.

Table 3. 2 Additional Inputs variables

Variable	Description	Units
P_s	Steam pressure	kPa
F_c	Circulating flowrate	Kg/min
F_{cw}	Cooling water flowrate	kPa
P	System operation pressure	kPa

For each effect, there are 7 state variables as shown in Table 3.3:

Table 3. 3 Variables in the i (=1, 2, 3) effect

Variable	Description	Units
F_{p_i}	Product flowrate	Kg/min
F_{v_i}	Vapour flowrate	Kg/min
F_{c_i}	Condensate flowrate	Kg/min
T_{s_i}	The steam temperature in effect	℃
T_{p_i}	Product temperature in effect	℃
ω_{p_i}	Product concentration	%
L_{s_i}	Separator level	m

The model has two main outputs in each effect which are used to compare with different control strategies. Meanwhile, these two outputs (as shown in Table 3.4) and two process variables (condensate flowrate and product temperature) are also treated as input variables for the second and third effects.

Table 3. 4 Variables in the i (=1, 2, 3) effect

Variable	Description	Units
F_{pl_i}	Product flowrate leaving the effect	Kg/min
ω_{p_i}	Product concentration	%

3.4.2 Initial conditions and constants

Table 3.5 lists the simulation initial conditions and values of all the coefficients, which are obtained from Newell and Lee (1989) and are used in developing the mathematical model and running the simulations in the following sections.

Table 3. 5 Initial conditions and values

Variable	Initial value	Unit	Minimum	Maximum
Product feed flowrate ($\mathrm{F_{feed}}$)	10	Kg/min	8	12
Product feed concentration ($\mathrm{w_{feed}}$)	5	%	4	6
Product feed temperature ($\mathrm{T_p}$)	40	℃	32	48

Cooling water inlet condenser temperature (T_{c_in})	25	℃	20	30

3.5 Development of a Simulink model in MATLAB

The simulation model of a three-effect falling film evaporator was developed using software MATLAB and SIMULINK based on the mathematical differential equations above. The simulation input variables are written in a script using MATLAB language and have an .m filename extension. In a typical dairy evaporation process, the product concentration increases from about 5-6% to 50% before further treatment. In this thesis, the designed product concentration increases effect by effect from 5% to 30%, 38%, and 52% finally. The initial conditions and constants of the simulation have been given in Table 3.5.

The SIMULINK block diagrams in the following figures show the details of how the effects, controller and inputs linked together in the MATLAB dynamic environment. There are three levels to the evaporator SIMULINK model.

i. The top level (Figure 3.1) shows the main inputs and outputs of the simulation model.
ii. The middle level (Figure 3.2) is where the main structure of the model can be observed. It shows how the three effects, controller blocks, and inputs with disturbances linked together and the information paths between them.
iii. The lower level (Figure 3.3) is an example of the first effect system which shows how the heat exchanger, steam ejector, separator and condenser linked together.

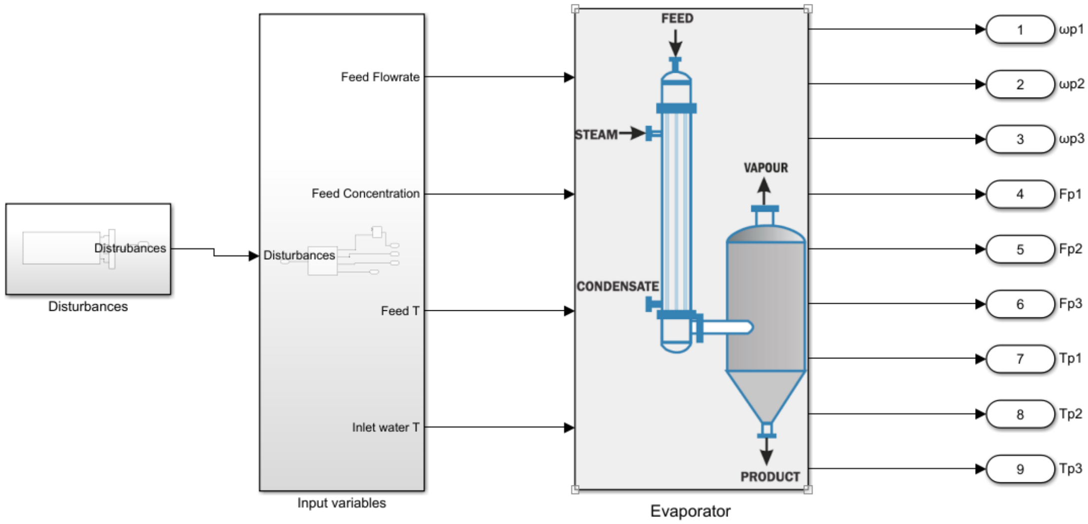


Figure 3.1 Top level of evaporator SIMULINK block diagram

The four main inputs variables with disturbances have been shown at the top level of the simulation block. The three effects and controller blocks are developed as sub-systems under the evaporator block. The right-hand side in figure 3.1 is the mainly observed outputs, which are the product concentration, flowrate, and temperature of the three different effects.

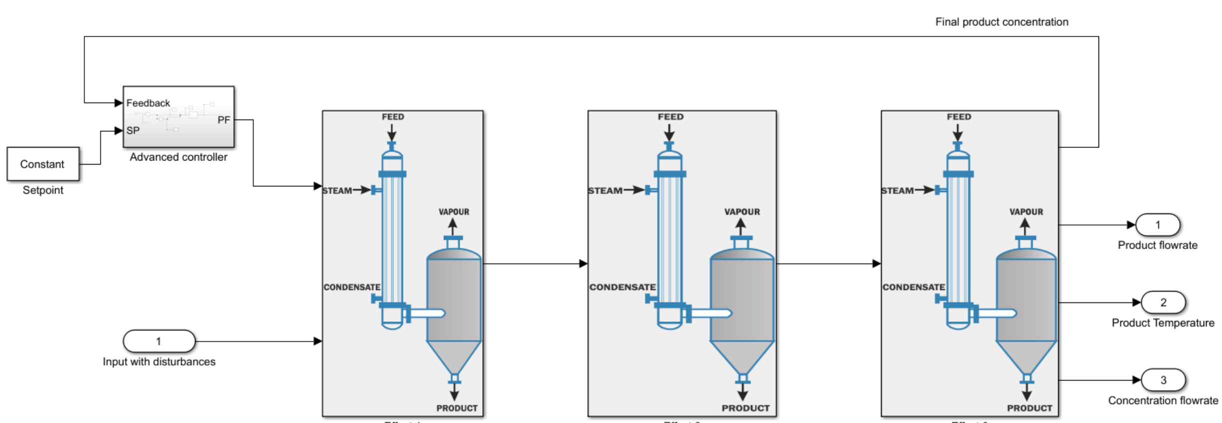


Figure 3.2 Middle level SIMULINK diagram of three effect falling-film evaporator

Figure 3.2 includes three effects blocks and a controller block. The advanced controller block is an outer-loop controller. There is one basic PID controller under each effect block. Meanwhile, there are many ‘From’ and ‘GoTo’ SIMULINK blocks under each effect to link them together.

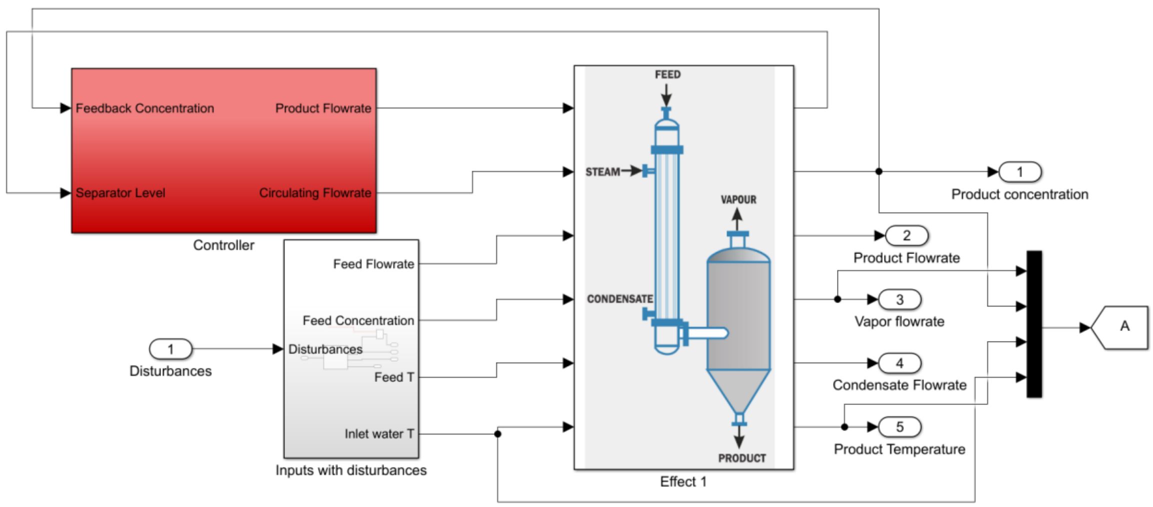


Figure 3.3 An example of lower level effect SIMULINK diagram

Figure 3.3 has indicated the arrangement of inputs and outputs to the first effect. Four output variables from the first effect are treated as inputs to the second effect. The SIMULINK block ‘GoTo’ is developed to link effects together. Product flowrate is a manipulated variable to control the product concentration under the basic PID controller block. Further details of the sub-systems can be referred to Appendix A.

3.6 Model validation

The purpose of model validation is to determine whether the model is an adequate representation of the physical process. In many practical situations, how accurate the simulation model can imitate the physical process is the main consideration to evaluate if the model is good enough. From a control engineers' perspective, the feasibility of the controllers and its implementation in the real industry also need to be considered.

However, it is impossible to obtain a perfect simulation model that can reproduce the entire industrial process without any errors. As discussed in the first Chapter, the simulation model developed in MATLAB is essentially required to be used for control systems design. Considering this, the dynamic behaviour of the evaporation process should be represented correctly by the model. The simulation error could be acceptable as long as the system model estimates follow the real industrial evaporation process and remain with the vicinity of the actual values.

The most common model validation method is by comparing the simulation results with the real manufacturing plant data. However, because the model developed in this thesis is an extension and improved from the model based on Newell and Lee's study, it is not mimicking a specific real industrial evaporation process. So to obtain real plant data as a reference is impossible. Therefore, the validation process involves applying the inputs to the simulation model and comparing the results trend with the previously published research work.

3.6.1 Comparison of the simulation results and previously studies

In order to evaluate the quality of the simulation model, two previous published research studies were used to obtain the necessary data. Both of the studies (Farsi and Jahanmiri, 2009; Russel, 1997) were focusing on the modelling and control of a three-effect falling film evaporator. The mainly compared variables are product concentration, flowrate, and temperature. As mentioned above, the simulation results data, trend, and rationality of the system are the preferential information which we are going to observe, analyse and compare.

A basic PID controller is applied to control the simulation performance in this model. The initial input variables are generated by adding a disturbance of 5% on each of them. An example of the random input sequences applied to the evaporator can be seen in Figure 3.4.

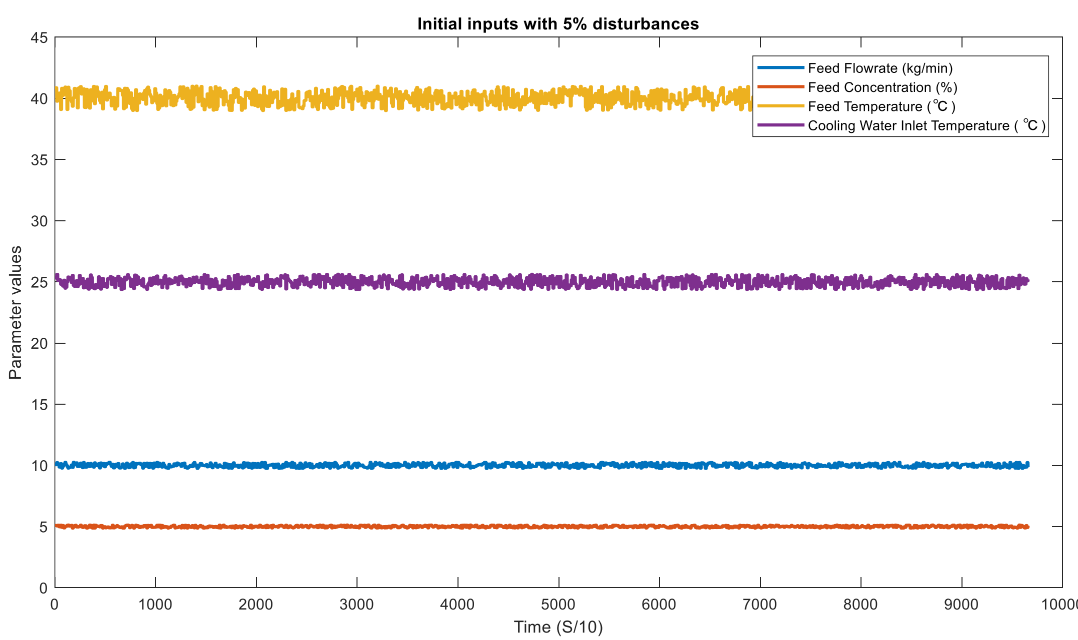


Figure 3.4 Initial input variables

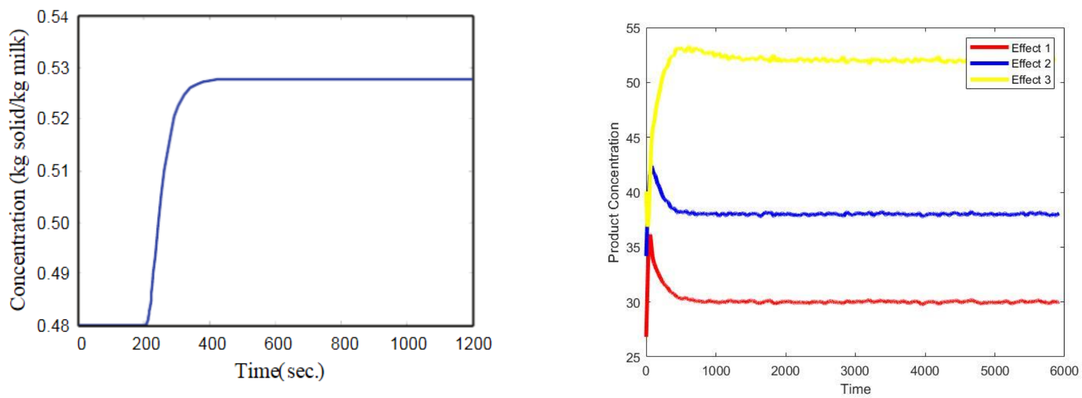
The designed simulation value of product concentration in each effect is 30%, 38%, and 52%. The study of a similar three effect falling film evaporator is published by Farsi and Jahanmiri in 2009. They indicated that the designed product concentration increases from 8.5% to 48%. The details are listed in Table 3.6.

Table 3.6 Designed product concentration in each effect of two studies

	Product Concentration	
	Farsi 2009	**This thesis**
Feed	8.5%	5%
Effect-1	13.35%	30%
Effect-2	29.56%	38%
Effect-3	48%	52%

From the table above, both studies are aiming to increase the product concentration from raw milk. And they all realised the target effect by effect. The difference between them is the concentration gaps between effects. The biggest gap of concentration increasing in Farsi’s study is from effect 2 to effect 3, which is approximately 19%. However, the value from this thesis is exact 25% from the first effect to the second one. Another difference is the total concentration increased range which is 39.5% and 47% respectively. Because the targets of both studies are similar, the process parameters trend should be relatively similar as well.

The following two figures show the trend of the product concentration for both studies. Farsi's study just published the data of the third effect with 10% step input disturbances (Fig 3.5a). It is indicated that the two figures have a very similar trend of product concentration from a lower value to a target value.



(a) (b)

Figure 3.5 Product concentration simulation results for both studies

However, due to the different scales of the two studies and the concentration gaps between effects, the process parameters are likely to be varied. The following table shows the product feed flowrate and the flowrate in effects. The units of flowrate are not the same for both, which Farsi's is in kg/hr and this thesis is in kg/min. In order to make the comparison easier to distinguish, the kg/min is transformed into kg/hr.

It has been generally accepted that the higher concentration the material is, the slower it flows in the evaporator. And the simulated product concentration increased from effect to effect are different. It is reasonable to assume that the flowrates in the evaporation process have a proportional correlation. The table above can be re-calculated as:

Table 3.7 Flowrate variables in each effect for both studies

	Flowrate steady-state values of each effect (kg/hr)	
	Farsi 2009	**This thesis**
Feed	10000	600
Effect-1	6367.0	400
Effect-2	2875.5	219
Effect-3	1770.0	61.9

Table 3.8 Re-calculated flowrate variables values

	Flowrate steady-state values of each effect (kg/hr)	
	Farsi 2009	**This thesis**
Feed	1	1
Effect-1	0.6367	0.667
Effect-2	0.2875	0.365
Effect-3	0.177	0.103

It is obvious that with the same feed flowrate, the product flowrate in each effect has a similar trend of the two studies. The following figure 3.6 shows the trend.

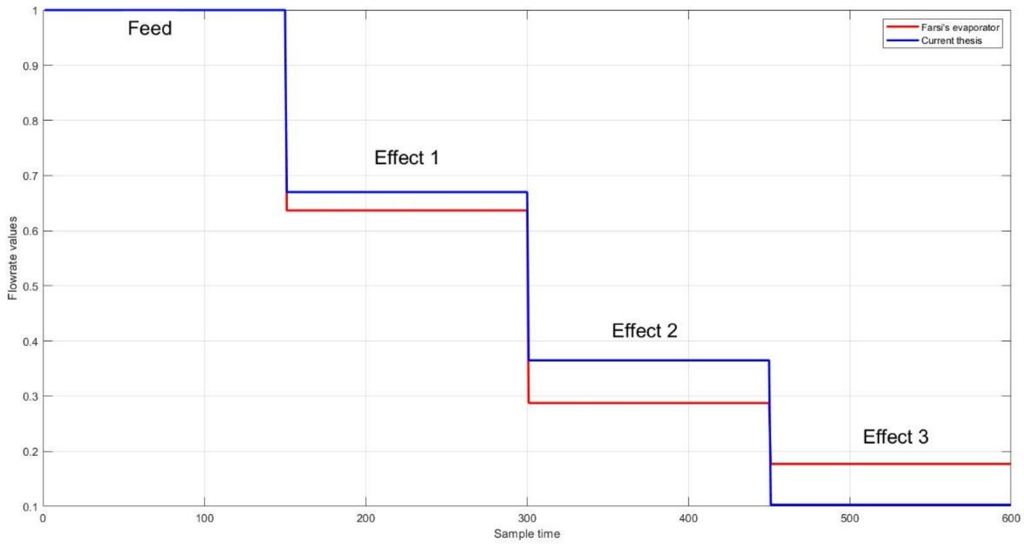


Figure 3.6 Flowrate trend of both evaporation process

Another important variable is the product temperature in the effect. In the industrial manufacturing evaporation process, the product temperature normally decreases progressively from effect to effect. In Farsi's paper, the three effects product temperature values are 72℃, 58℃ and 45℃ from the first to the third respectively. The values of current model simulation results are 88.86℃, 53.27℃ and 44.38℃ respectively as shown in Table 3.9.

Farsi's model contains 4 pre-heater system to heat the raw milk until 72℃ before feeding into the first effect. Meanwhile, there is an assumption that the heat loss between the pre-heater and the first effect system is neglected. That results in the product temperature of the feed and the first effect are both 72℃. With the water removed from the milk, the product temperature drops from 72℃ to 45℃ finally when leaving the evaporator.

Table 3.9 Product temperature of each effect of Faris's study and this thesis

	Product temperature steady-state values of each effect (℃)	
	Farsi 2009	**This thesis**
Feed	72	40
Effect-1	72	88.86
Effect-2	58	53.27
Effect-3	45	44.38

Unlike Farsi's model, the pre-heater system was not developed and applied in the simulation model in this thesis. The feed temperature is designed to be 40℃ with disturbances. The concentration increasing gap between first and second effect is larger than Farsi's model, which means more water need to be evaporated in the first effect.

So the higher product temperature is a reasonable value. Comparing the second and third effects, both studies have similar values.

According to the main variables (flowrate, temperature, concentration) comparison with the previous study, most of the simulation data is within an acceptable range, and the figure trends are very similar and reasonable. It is believed that this model is accurate enough to achieve process control comparison and the controller improvement for the evaporation process.

Chapter 4 Conventional PID and Auto-tuning PID Control of Three-effect Falling Film Evaporators

4.1 Introduction

Process control is very important for large industrial manufacturing plants to provide uniform plant operation and high-quality products. Both of these often lead to greater profits through lower operating costs and higher revenue. In this chapter, the conventional PID control strategy will be applied to the simulation model. It is well known that the traditional PID is a widely used controller in many industrial applications to regulate temperature, flow, pressure and other process variables. Meanwhile, an auto-tuning PID controller will be investigated as an optimised PID to the system model to achieve the control targets.

Both PID and auto-tuning PID works by calculating an error value continuously as the difference between the desired set-point and a measured process variable, then applies a correction based on the proportional, integral, and derivative terms (P, I and D).

4.2 Process Control and disturbance variables

The three-effect falling film evaporator contains many variables that have been listed in chapter 3. For control purposes, the following 3 manipulated variables, 3 disturbance variables, and 3 process variables were discussed in this chapter.

Figure 4.1 describes the diagram of the evaporation process with all these 9 variables as inputs/outputs and the details are shown in Table 4.1 as follows:

Table 4.1 Control variables of the evaporation process

Manipulated Variables	**Disturbance Variables**	**Process Variables**
Steam Pressure (P_s)	Cooling Water Inlet Temperature (T_{c_in})	Product Flowrate (F_{p})
Cooling Water Flowrate (F_{cw})	Product Feed Concentration (w_{feed})	Product concentration (ω_p)
Product Feed Flowrate (F_{feed})	Product Feed Temperature (T_p)	Product temperature in effect (T_{p})

Product concentration is the main controlled variable in each effect. It is also the feedback signal subtracted by the set-point to generate the system error. However, controlling the production concentration directly is difficult to achieve. According to mass balance Eq. 3.17, the product concentration depends upon its flowrate in the effect. As a result, it is easier to control the product flowrate rather than the concentration directly.

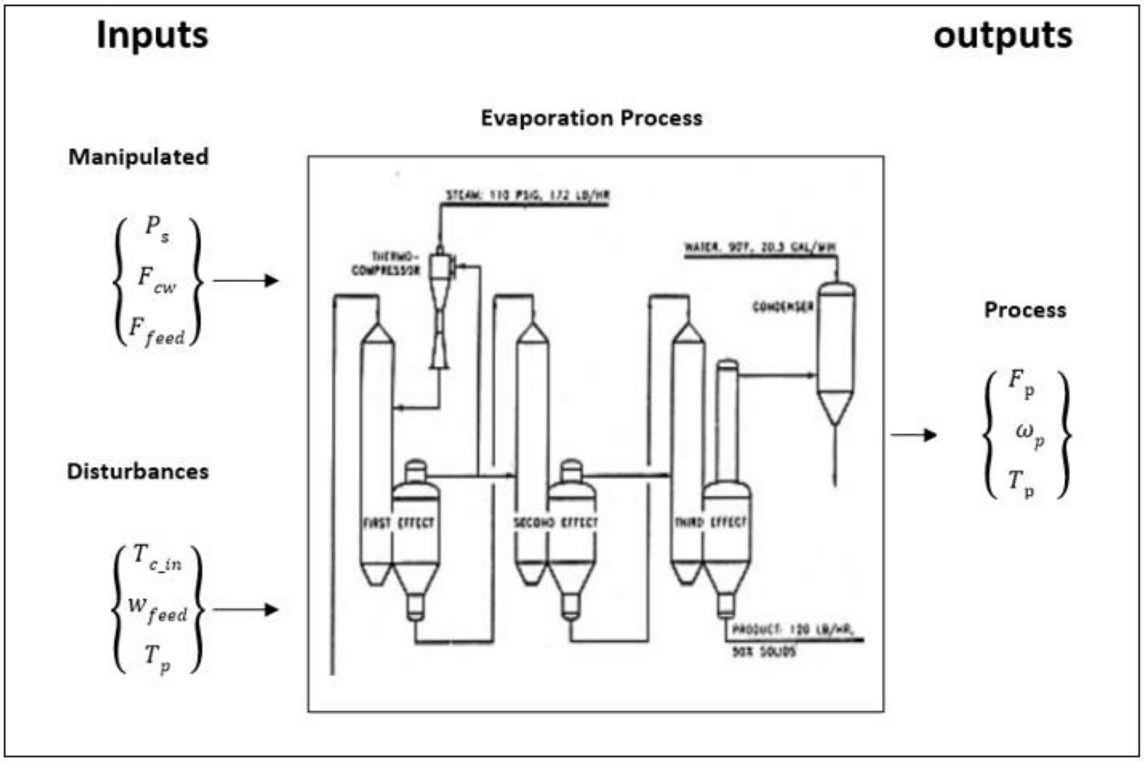


Figure 4. 1 Control variables in the falling film evaporator

4.3 Proportional Integral Derivative (PID) control

The first formal PID control law was developed by Minorsky (1922) to analyse and apply in the field of automatic steering systems for ships. Since then, this control method is becoming more and more popular and implemented successfully in many industrial processes. The ability to use three control terms, which are proportional, integral and derivative, is the unique feature of a PID controller. Figure 4.2 shows the structure diagram of a traditional PID controller.

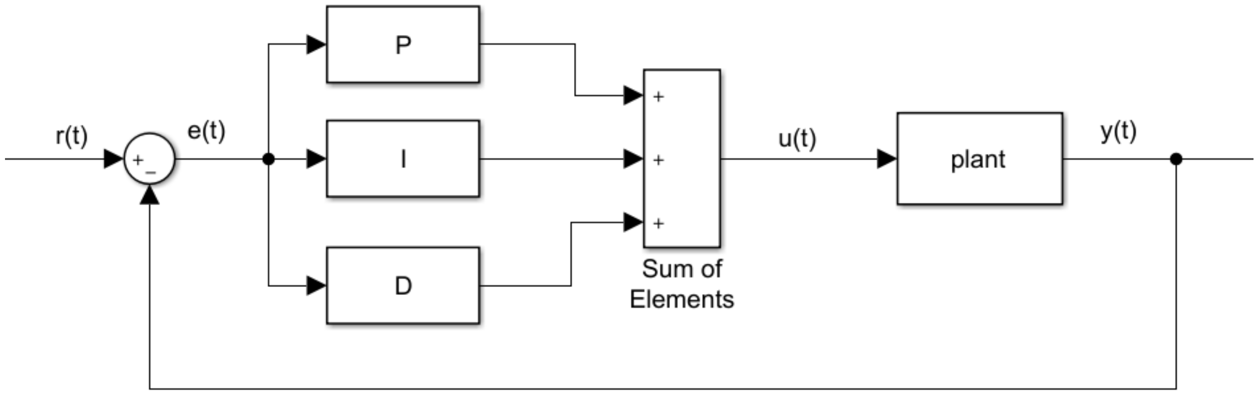


Figure 4.2 PID controller structure diagram

Where *r*(*t*) is the desired process value or set-point, *u*(*t*) is the controlled input signal to the process plant, and e(t) is the system error generated by *r*(*t*)-*y*(*t*).

The following equation shows the mathematical function of the PID controller:

$$u(t) = K_p e(t) + K_i \int_0^t e(t')dt' + K_d \frac{de(t)}{dt} \quad (4.1)$$

where K_p, K_i and K_d are the coefficients for the proportional, integral and derivative terms respectively. (P, I and D in the figure).

As a classic feedback control, the PID controller works by measuring the control variable, comparing it to a specific or desired value, known as set point, and applying the control law to determine the controller action that causes the control variable to track the set point. With a given constant set point, PID controller normally provides disturbance compensation or regulation. If the setpoint is a trajectory signal, the PID offers the servo control.

The three terms P, I and D have a different way to control the system. Term P is proportional to the current system error e(t). Considering the gain factor 'K', the control output is proportionately large and positive with a large and positive error. By only using the term P can provide an error between the setpoint and the actual process value. But in order to generate the proportional response, a system error is required. This means it can never reach a steady-state condition even under a stable operating environment. Due to the limitation of the proportion only controller where there always exists an offset between the process value and setpoint, the term I, which provides necessary action to eliminate the steady-state error, is needed. The error has been integrated over some time until its value reaches zero. If the error is negative, the integral control decreases its output. Meanwhile, it limits the speed of response and affects the system stable. For most of the cases, the PI controller is used particularly if a high-speed response is not required. The term D has the capability to predict the future behaviour of error based on the current rate of error change.

In this thesis, only PI controllers are needed to control the simulation model. Because the term D can result in too much time to run the simulation but tiny improvement of the controller's performances. The mathematical form of the PI controller can be re-organized according to Eq. 4.1:

$$u(t) = K_p \left(e(t) + \frac{1}{T_i} \int_0^t e(t')dt' \right) \tag{4.2}$$

where T_i is the integration time.

There are many methods to determine the value of the PI controller tuning parameters. Common ones include the method of Cohen-Coon, Integral of Time-weighted Absolute Error (ITAE), relay method, auto-tune, and the most popular Ziegler-Nichols (ZN) 1st and 2nd tuning methods.

The first ZN method is using the S-shaped step input response curve to locate the inflection point and draw a tangent line to find the delay time L and time constant T to calculate the control terms value. The two intersectional points of the tangent line with the time axis and the steady-state level line decide the values of L and T (Fig. 4.3).

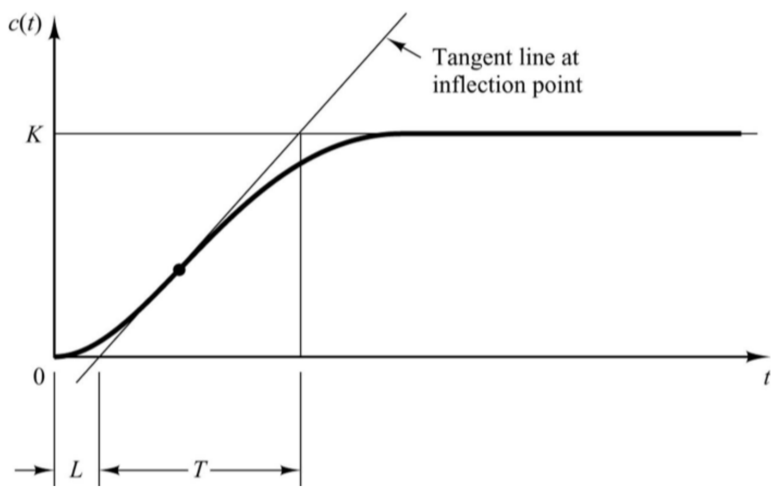


Figure 4.3 S-shaped step input response curve

After finding the L and T value from the curve, the final control terms value is calculated based on table 4.2.

Table 4.2 Ziegler-Nichols 1st tuning rule based on step response of plant

Type of Controller	K_p	T_i	T_d
P	T/L	∞	0

PI	0.9T/L	L/0.3	0
PID	1.2T/L	2L	0.5L

The second NZ method needs to find two values as well. A gain proportional K which makes the system in a steady-state oscillation condition and the system period P (see Fig. 4.4). To find K, it normally starts with a closed-loop system and a lower or zero value of gain K. Increase the K value until a steady-state oscillation occurs.

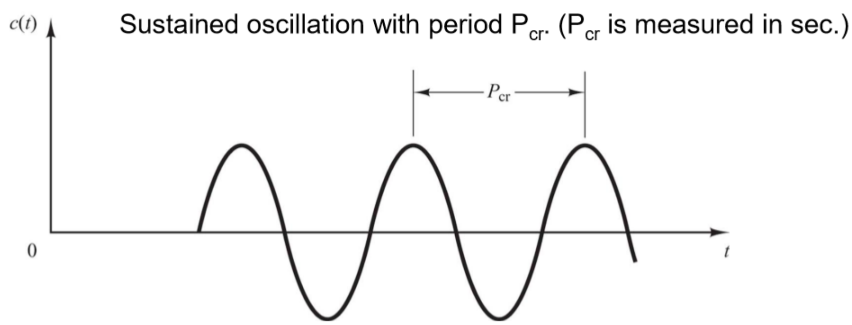


Figure 4.4 An example of steady-state oscillation

Once the K and P have been found form the curve, Table 4.3 is needed to calculate the PID control terms:

Table 4.3 ZN second tuning method gain estimator rules

Type of Controller	K_p	T_i	T_d
P	0.5K	∞	0

PI	0.45K	P/1.2	0
PID	0.6K	0.5P	0.125P

The second NZ tuning method is applied to this nonlinear model to determine the PI control gain for the production concentration (ω_p) and production flowrate (F_{p}) loop. In this tuning procedure, a proportional-only controller is implemented with a set-point change of the concentration input aiming to find the gain at which the process first begins to oscillate with a constant amplitude. The result was:

K=3.68， P=0.05

So the real PI values recommended follow ZN's second method was:

K_p = 3.68*0.45=1.656,

T_i = 0.05/1.2 = 0.0417

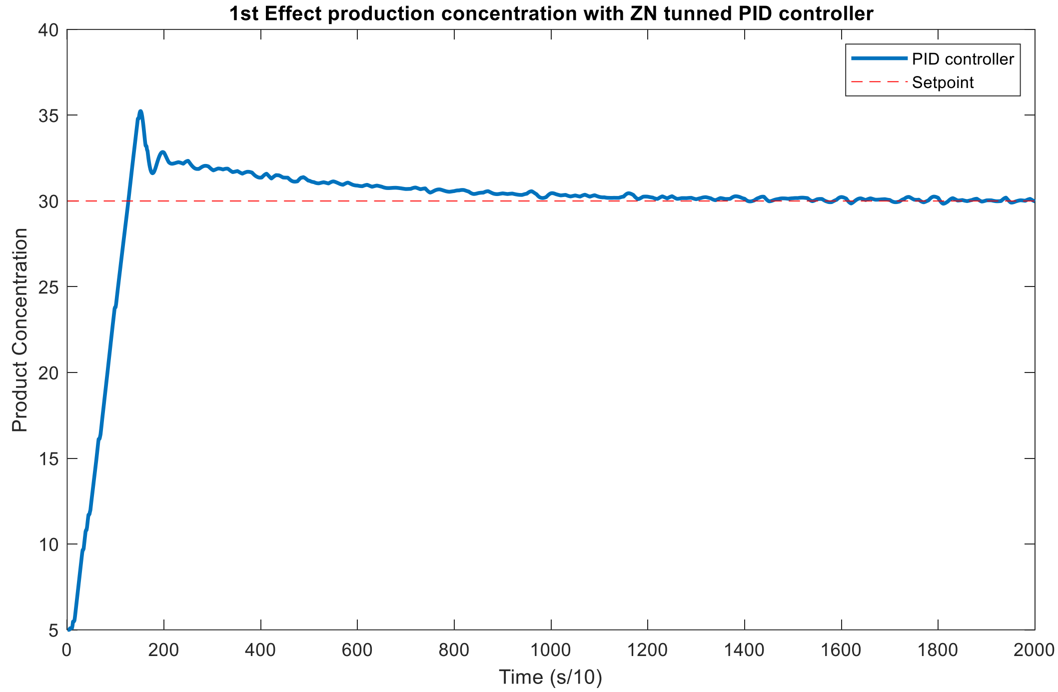


Figure 4.5 Concentration of 1st effect with ZN tuned PI controller

Figure 4.5 shows the first effect evaporated the milk from the feed 5% to the designed 30% under the conventional PI controller. The max amplitude value is 35.22 and the overshoot is 19.28%. Rise time is about 20 seconds which means that the system responding speed is very fast. However, the setting time shows in the figure is nearly 100 second which needs to be improved to ensure the output reach and remain within the acceptable error as quickly as possible.

Repeat the same procedure above, the 2nd and 3rd effects control gains are found and listed as follow:

Table 4.4 Control gains of the second and third effect

2nd effect		3rd effect	
K = 5.47	P = 0.9	K=3.82	P= 0.65
K_p= 2.46	T_i = 0.75	K_p= 1.72	T_i = 0.54

Figure 4.6 below shows the controller performance on the production concentration of the second and third effects.

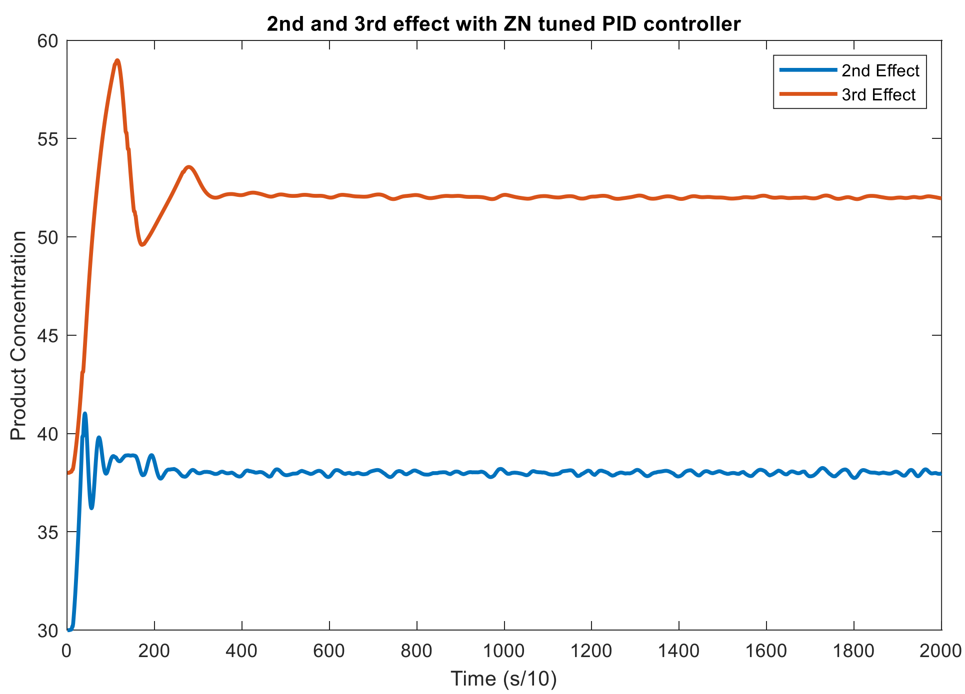


Figure 4.6 Production concentration of 2nd and 3rd effect with ZN tuned PI controller

The control performances are not easy to describe and compare from the figures of the three effects. But the following statistics table is more straightforward.

Table 4.5 Signal statistics of the 2nd and 3rd effect production concentration

	2^{nd} effect	**3^{rd} effect**
Max Amplitude	43.16 (38)	66.22 (52)
Rise Time (s)	6.47	11.04
Overshoot	64.5%	101%
Setting Time	200	220

As we discussed above, all the PID gains are calculated by following the ZN second method and applied to control the simulation process. However, both the figures and the table data indicated that the conventional PID has a better performance on the first effect than it's on second and third. The main reason is that with the increasing of the effects, the evaporation process is getting more and more complex. Conventional PID cannot deal with a very complex system. This is one of its disadvantages. As a result, only the conventional PID control strategy is not enough to maintain the complex system performances in most cases.

4.4 Auto-tuning PID Control

An auto-tuning PID controller is introduced in this section and applied to control the falling film evaporation simulation process. The results are going to be compared with the conventional PID control strategy. A MATLAB auto-tuner SIMULINK block is used to do the plant frequency estimation in real-time according to the input-output data to obtain the best P, I value for the process.

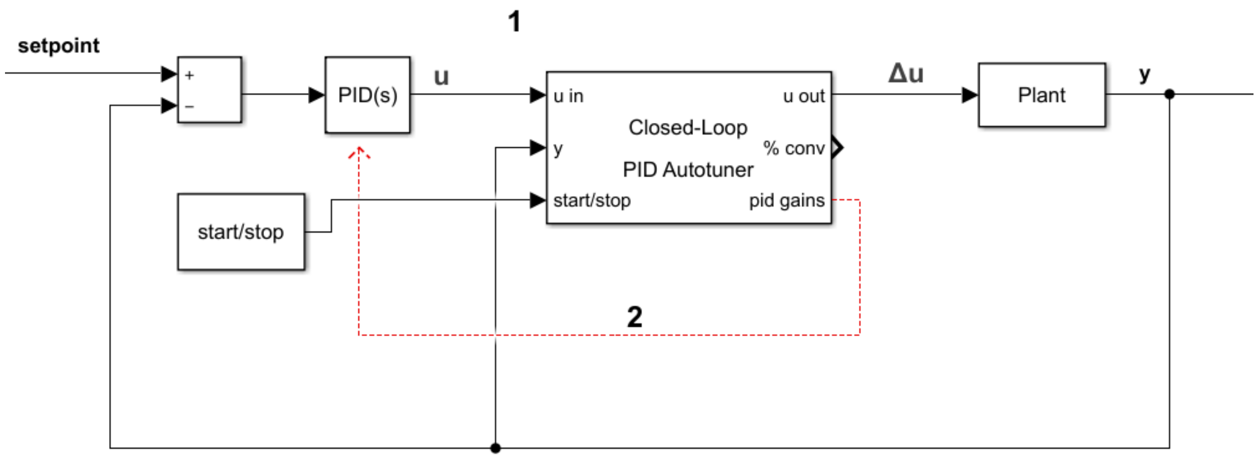


Figure 4.7 Auto-tuning PID MATLAB structure

The auto-tuning process works in two steps:

1. When the auto-tuning process starts, a test signal is injected into the auto-tuner block and as an input to the plant system. After that, the plant output data is collected and stored back to the auto-tuner. These series of input-output data are used to estimate the system frequency response in real-time. Once the tuning ends, the PID gains are computed based on the estimated system frequency.

2. The new PID gain values are transferred to the PID controller (see red dotted line in the figure above) and re-write the old PID values. Now the system is under control of an updated PID controller. Meanwhile, in this step, the plant output signal feeds back to the set-point, not the auto-tuner block.

Because this is an online tuning process, and the tuning algorithm aims to balance system performance and robustness while eliminate the errors between the system feedback and the setpoint. It can find more appropriate PID gains for the plant system comparing to the manual calculating methods. Besides, to find a better gain also depends on the start-stop duration.

The auto-tuning PID developed and used in this simulation model is shown in Figure 4.8 below. Two inputs in the figure are the control target (set-point) and production concentration from the plant (feedback) respectively. There is a PID controller to provide an initial control gain to generate the input signal (u) to auto-tuner. It is also used to receive and re-write the updated PID gains, after auto-tuning, to control the whole system plant. When it starts to tune, the output signal (u+Δu) from the auto-tuner fed to the plant to obtain the plant output. During the tuning process, the input-output data is collected and applied to estimate the plant frequency response.

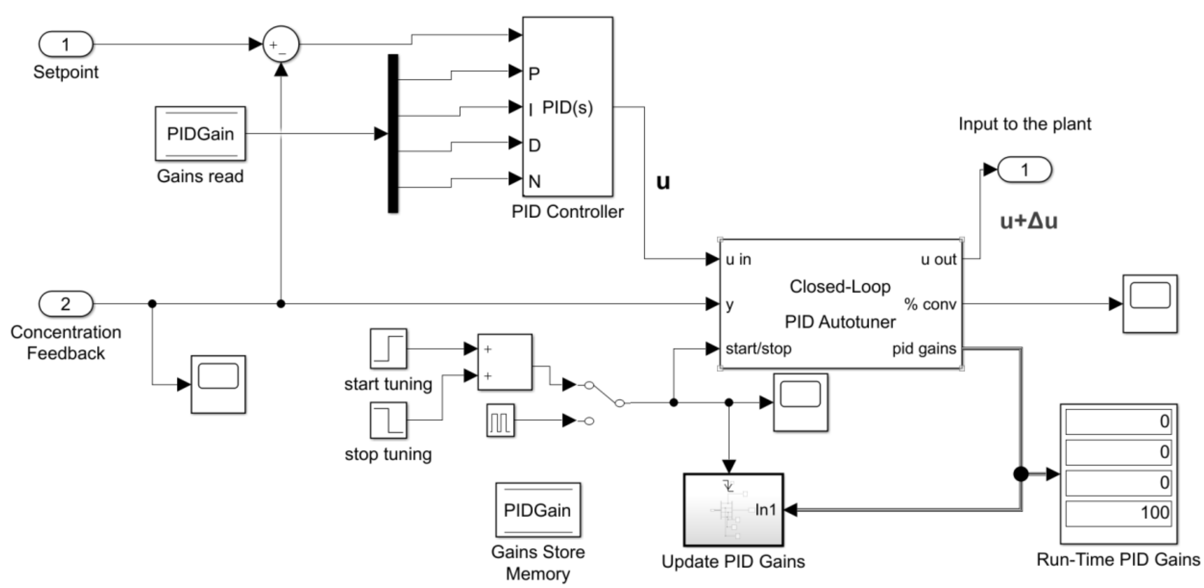


Figure 4.8 Auto-tuning PID MATLAB structure

Once the auto-tuner receives the stop signal, the new PID gains are computed according to the frequency response. These values are shown on the ‘Run-Time PID Gains block’, saved in the Gains Store Memory and finally transferred to the Gains read to update the PID controller. In thesis, the running time for the testing auto-tuning process is 960 seconds. The auto-tuning duration is in 5 groups which are 30, 60, 120, 240 and 480 seconds. The results with different tuning period are compared to indicate the most suitable tuning time for the model.

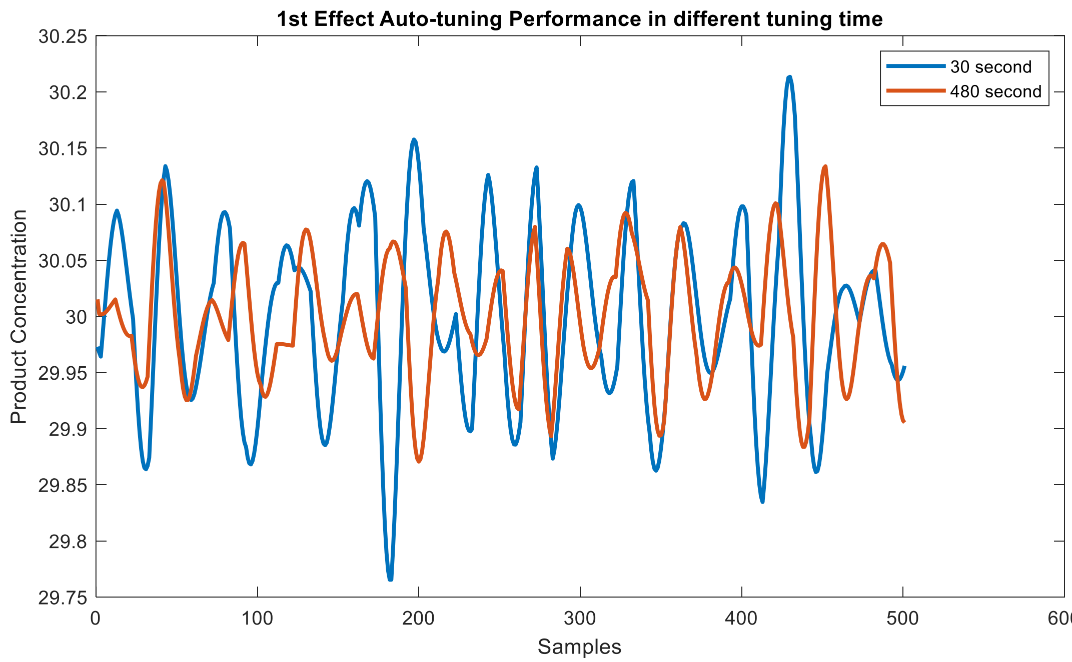
Table 4.6 listed the PID controller gain values found by the auto-tuner within different tuning period.

Table 4.6 PID gain values in the different auto-tuning period

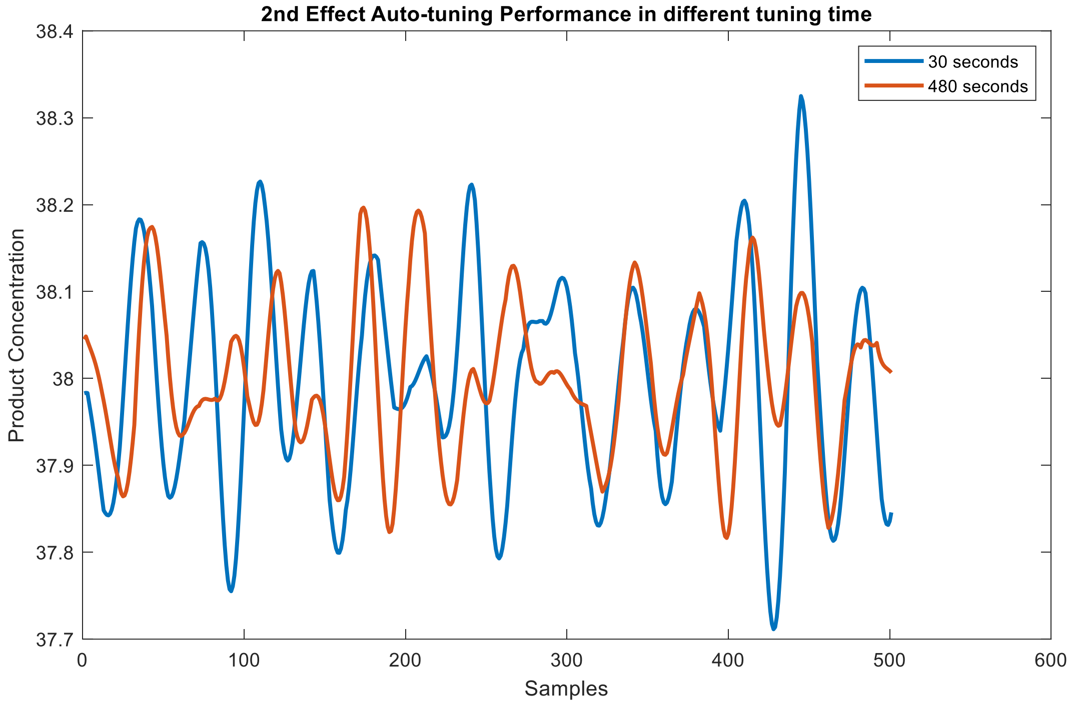
Period	**Gains**	**1st Effect**	**2nd Effect**	**3rd Effect**
30 (s)	K_p	3.39	3.486	2.068
	T_i	0.684	1.124	0
60 (s)	K_p	3.657	3.736	1.83
	T_i	0.6776	1.217	0
120 (s)	K_p	3.632	3.576	1.835
	T_i	0.5025	0.9554	0
240 (s)	K_p	3.631	3.706	1.779
	T_i	0.3799	0.8672	0

480 (s)	K_p	3.551	3.54	1.773
	T_i	0.3618	0.879	0

The simulation results indicated that in the 1st and 2nd effect, only 30 seconds auto-tuning period is enough for finding the appropriate controller gains. According to figure 4.9 (a) and (b) below, the product concentration trend and amplitude of the 1st and 2nd effects are very similar in 30 seconds and 480 seconds.



(a) 1st effect



(b) 2nd effect

Figure 4.9 1st and 2nd effect auto tuning performance in different tuning period

However, without enough tuning time, the controllers' performances of the third effect are different. The red and blue curves in the following figure (4.10) are controlled in 30 and 60 second tuning time. The product concentration is maintained at 53% and 52.7% respectively, which are relatively far from the control target 52%. The other three lines at the bottom controlled by the tuning period are 120 (yellow), 240 (cyan) and 480 (black) second. The cyan and black curves stay closer to around 52% than the yellow one.

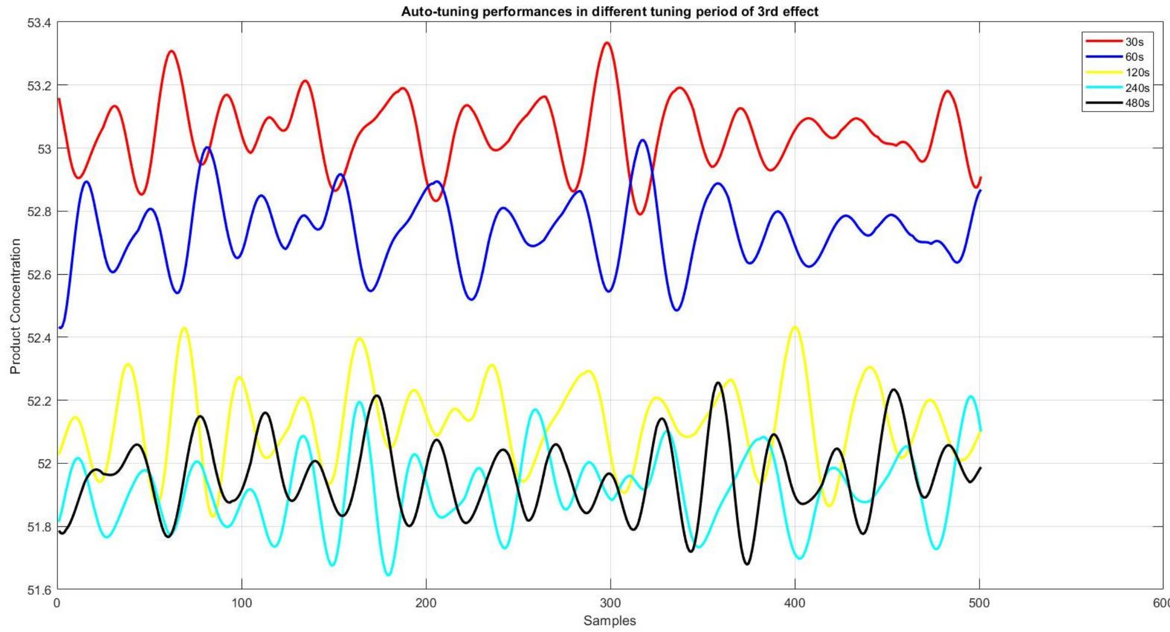


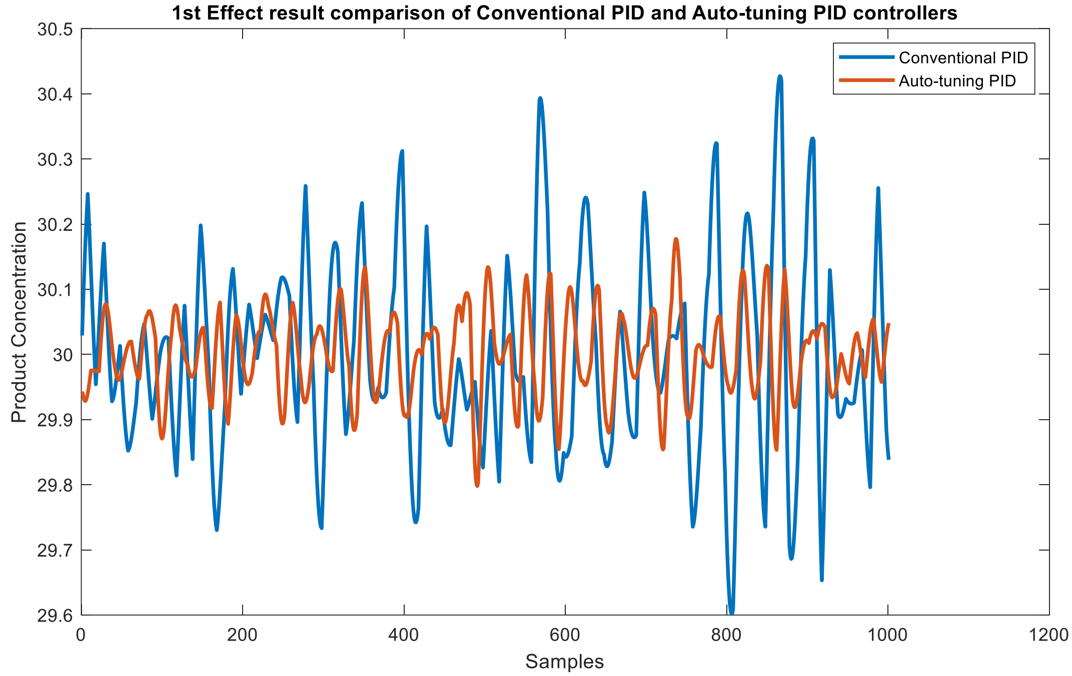
Figure 4.10 3rd effect auto-tuning performances in different tuning period

4.5 Comparison of conventional and auto-tuning PID with step input

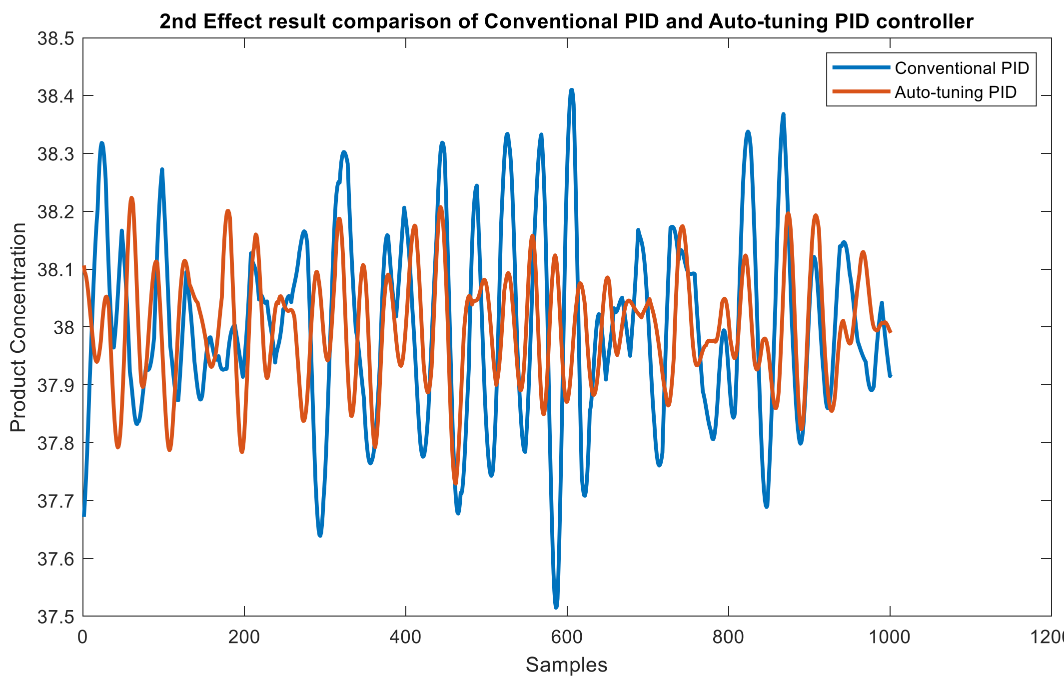
It is assumed that the auto-tuning PID should have a better performance than the conventional PID controller. Because the conventional PID control gain values are decided by the manual calculation based on the system mathematical equations theoretically. However, almost all of the industrial manufacturing processes are highly complex non-linear systems. To obtain these transfer functions are difficult and not precisely enough to describe the nonlinear systems. For better understanding and control purposes, linearization is normally applied to develop a linear approximation of a nonlinear system that is only valid in a small region around an operating point. This process cannot replicate all the characteristics of a nonlinear system. However, the auto-tuning method used in this model depends on the real input-output data collected from the model plant to estimate the frequency response to compute the PID gains.

A step signal was generated as the specific control objectives. Auto-tuner PID employs 480 seconds tuning period to control the model and to compare to the conventional PID controller performances. 20% of disturbances are added to the four initial input variables. 1000 data samples of product concentration were extracted from the simulation results in each effect controlled by both conventional PID and auto-tuning PID controllers to obtain the performance. The comparison results are demonstrated in Figure 4.11.

Figures 4.11 (a), (b) and 4.12 have indicated the comparison results that the auto-tuning PID controller has a better performance than the conventional PID in this simulation model. The red curves in these three effects are covered by the blue curves. It is not that obvious in the first effect, but very clear in the second and third effect. This means in the same condition, product concentration in each effect has a smaller amplitude and closer to the target value controlled by the auto-tuning PID (Meng et al, 2019).



(a) 1st effect



(b) 2nd effects

Figure 4.11 Results comparison of 1st and 2nd effects

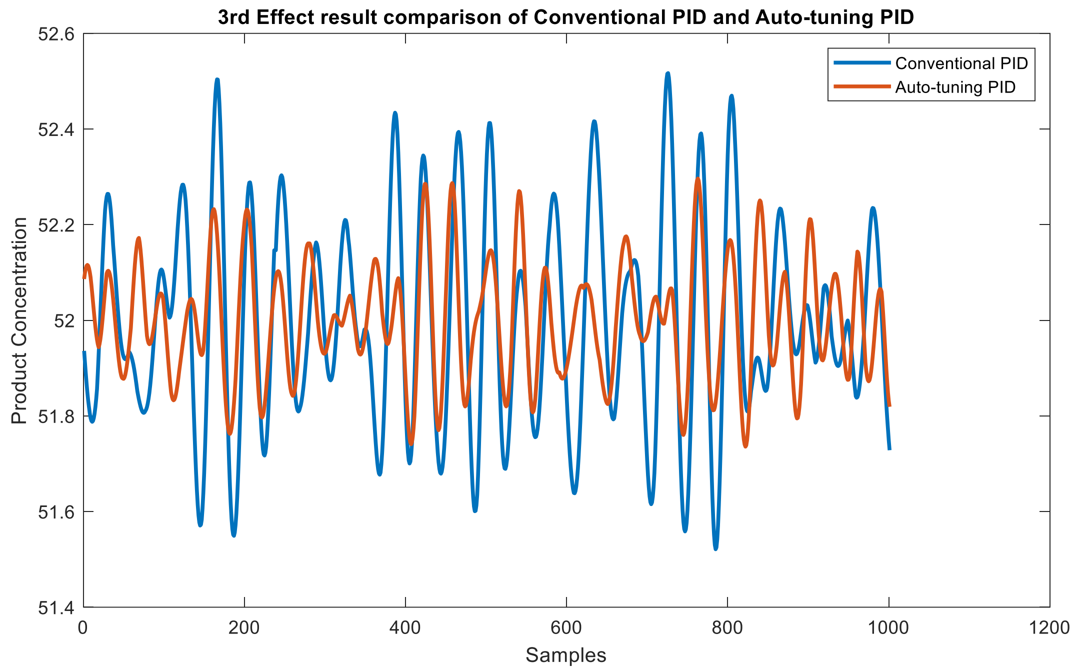


Figure 4.11 Results comparison of 3rd effect

4.6 Conclusions

The product concentration controlled by auto-tuning PID has smaller errors than controlled by conventional PID of this evaporation process in this thesis. This means the proportional and integral gains computed by the online auto-tuner based on the system input/output data can achieve more precious control results than the gains calculated by the Ziegler–Nichols method.

One of the disadvantages of the auto-tuning method is the tuning time requirement to get the input/output data. For the evaporation process in this thesis, at least 240 seconds is needed tuning automatically to find the appropriate PID gains. This period could vary in different simulation processes.

However, although the auto-tuning PID performs a little bit better, it is still a PID based control strategy essentially. As it is shown in the figures above, the model results improved, but not very much. Therefore, more advanced control strategies are required to develop and improve system performances.

Chapter 5 Model Predictive Control (MPC) of Three-effect Falling Film Evaporators

5.1 Introduction

Model predictive control (MPC) is a multivariable control strategy applied to predict the future output of the system based on the historical information and the predictive model of the process. The objective of MPC calculation is to determine a sequence of control moves, which is manipulated input changes so that the predicted response optimally moves to the setpoint. A general model predictive controller consists of 5 basic elements (Russell, 1997):

1. A dynamic model from the predictions of future plant output, which can be obtained based upon the historical information and a future set of control input moves.
2. The future reference signal, which is predefined or determined through the use of a separate process optimisation procedure, for the plant outputs.
3. A function of the future output errors (**e**), and the set of proposed control inputs (**u**), named cost function $\emptyset(\boldsymbol{e},\boldsymbol{u})$**.**
4. An optimisation routine which minimises $\emptyset(\boldsymbol{e},\boldsymbol{u})$ either unconditionally or subject to various constraints placed upon control and process operation.
5. Various filters which help to broaden the range of the performance and provide robustness against unmeasured disturbances or plant-model mismatch.

The concepts, principles, and applications of MPC have already been introduced in Chapter 2 (literature review).

As discussed in the first chapter that the main difficulty in controlling food manufacturing plant is the inherent nonlinear and multivariable nature of the processes. While basic and simple control schemes, such as the PID controller, are in common use throughout industry indeed and can be satisfied in some cases. However, there are more efforts required to develop new control techniques to improve the quality and consistency of industrial processes. For example, in chapter 4, it has been proved that the PID controller is very weak to provide adequate control of the multi-input and multi-output systems. Meanwhile, the performances for controlling a complex system fail to meet our expectations. That is why advanced control approaches are required.

In this chapter, two particular and popular MPC methods, which are Dynamic Matrix Control (DMC) and Model Algorithmic Control (MAC), will be introduced and implemented as the control methodologies in the simulation model.

5.2 Two Types of MPC controller

5.2.1 Dynamic Matrix Control (DMC)

Dynamic Matrix Control (DMC) is one of the most classic MPC strategies, introduced by Cutler and Ramaker in 1979, which applied initially to a fluid catalytic cracker industry. A linear step response was used to represent the process to model the system. The following equation is the general form of the models:

$$\hat{y}(t+1) = y(t) + A\Delta u \tag{5.1}$$

Where, $\hat{y}(t+1)$ is the future model output predictions,

$y(t)$ is the measured plant outputs,

A is the dynamic matrix of step response coefficients,

Δu is the matrix of past and current control moves.

According to this form, a necessary step is to find the N_c set of future control moves ($\Delta u(t+i), i = 1, 2, \ldots, N_c$), also known as control horizon at each present time *t*.

$$\emptyset = \sum_{i=1}^{N_p} e(t+i)Qe(t+i)^T + \sum_{i=1}^{N_c} \Delta u(t+i)R\Delta u(t+i)^T \tag{5.2}$$

Where $$e(t) = r(t) - \hat{y}(t),$$

$$\Delta u(t) = u(t) - u(t - q1),$$

$Q\ and\ R$ are positive weighting matrices, N_p and N_c represent the specific prediction horizon and a set of future control moves respectively. The function acts on the vector of output errors from the process $\boldsymbol{e}$, and penalises large changes in the control inputs, $\boldsymbol{u}$.

The DMC optimisation strategy is carried out subject to defined process constraints and offers flexible constraint handling within the control strategy. The control problem can be formulated to minimise the objective function subject to the following constraints:

- Constraints on the control inputs:

$$u_{min} \leq u(i) \leq u_{max} \tag{5.3}$$

$$i = t, \ldots t + N_c$$

- Constraints on the deviation from the previous control inputs:

$$u(i-1) - \Delta u_{max} \leq u(i) \leq u(i-1) + \Delta u_{max} \tag{5.4}$$

$$i = t, \ldots t + N_c$$

- Constraints on the model outputs:

$$y_{min} \leq \hat{y}(i) \leq y_{max} \tag{5.5}$$

$$i = t, \ldots t + N_c$$

- Constant control action assumed for time steps beyond the control horizon, N_c:

$$u(i) = u(t + N_c - 1) \tag{5.6}$$

$$\forall i > t + N_c - 1$$

The optimisation outlined above constitutes a standard quadratic program which can be solved on-line in a finite number of steps. Meanwhile, the algorithm provides a wide choice of control response through the use of the various tuning parameters available. The adjustable parameters are N_p, N_c, $Q\ and\ R$.

Austin and Bozin (1996) found that the N_p is usually equal to the step response setting time of the system because of the following reasons:

1. The control action is more aggressive,
2. The system response is faster
3. The closed-loop system is less robust to plant-model mismatch

To reject the disturbances by adding robustness to the control scheme, a correction is applied in DMC to obtain a better prediction. This correction is based on the difference between the actual value of the controlled output at the current sampling instant and that predicted by the model:

$$d(y) = y(t) - \hat{y}(t) \tag{5.7}$$

It is assumed that the value remains constant over the prediction horizon from time t to $t + N_p$. The error signal can be re-organized as:

$$e(t+1) = r(t+1) - \hat{y}(t+1) - d(t), \quad i = 1,2,\dots,N_p \tag{5.8}$$

Where the r(t) is the reference signal, which can be a constant or a smooth trajectory curve reach the desired setpoint. A block diagram representation of the MPC system is given in the following Figure 5.1.

5.2.2 Model Algorithmic Control (MAC)

MAC was initially introduced by Richalet to develop a model predictive heuristic control in 1978 and then applied it successfully to the control software system in a steam generator and an oil refinery distillation column. These three distinct aspects which a general MAC is different from DMC:

1. An impulse response model involving u is used in MAC instead of the step response model with Δu in DMC.

2. Only one control horizon in MAC, therefore, $N_c = N_p$.

3. The disturbance estimation equation (5.7) is filtered and re-wrote as follow:

$$d(t+i) = \alpha d(t+i-1) + (1-\alpha)[y(t) - \hat{y}(t)] \tag{5.9}$$

$$i = 1,2, \ldots, N_p,\ d(t) = 0\ and\ 0 \le \alpha \le 1.$$

A first-order exponential filter is added in Eq.5.7 in the feedback path. α is the filter coefficient which is a much more direct and convenient tuning parameter than the cost function weights or horizon length in the general MPC formulation.

5.3 Model Predictive Control strategy and Tuning

5.3.1 MPC objctive function and diagram structure

As we discussed in the previous chapter, the main control objectives for the evaporation process are to maintain consistent product concentration by controlling the product flowrate in the effects. For a general evaporator system, the manipulated variables are usually the steam flow, feed flow, and condenser cooling water flow. Both the flowrate and steam input influence the rate of evaporation and temperature in the effects.

The product temperature in each effect is an important controlled variable. Firstly, by regulating the temperature the energy input into the system can be controlled. Secondly, the temperature has a critical influence on the product quality, especially for heat sensitive products. Therefore, it is essential to control the temperature in the system, especially the first effect which normally has the highest temperature in real industrial evaporators. Because the mass of product in the system is constant, the product flowrate through the plant is determined by the feedback concentration level through the MPC. As a consequence, the MPC strategy applied to the simulation model aims to control the product flowrate in the first effect and temperature.

Both variables can be controlled via an MPC controller with a concentration setpoint due to the product concentration correlates with both variables. The controlled and manipulated variables of the MPC are summarised in the following table:

Table 5.1 Controlled and manipulated variables for MPC

Controlled variables	Manipulated variables
Product Flowrate	Product Concentration
The temperature in the first effect	

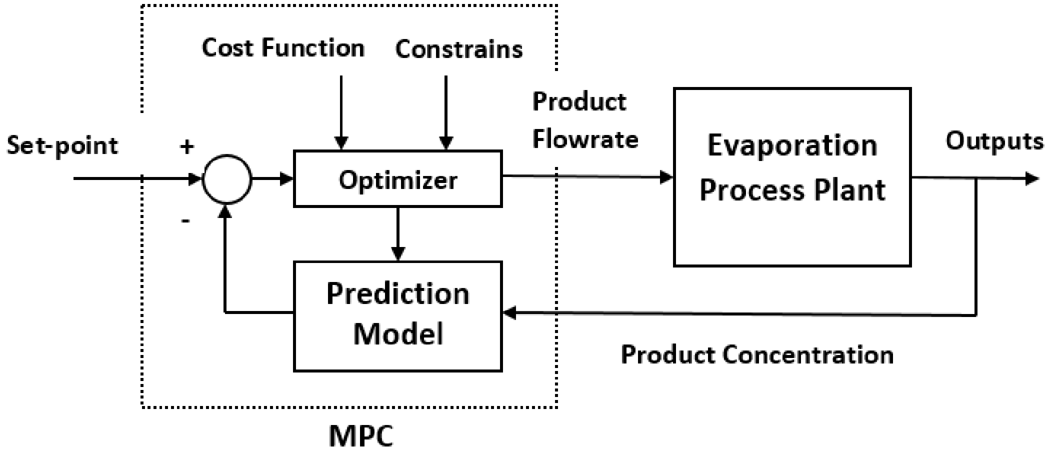


Figure 5.1 Diagram of MPC structure

The constrained optimisation technique Quadratic Programming (QP) is preferred to optimise the optimal manipulated variables. It attempts to minimise the objective function:

$$\begin{cases} \Phi = \int_{i=1}^{N_p} e(t+i)Qe(t+i)^T + \int_{i=1}^{N_c} \Delta u(t+i)R\Delta u(t+i)^T \\ e(t+i) = r(t+i) - \hat{y}(t+i) - d(t), \quad i = 1,2,\dots,N_p \\ d(t) = y(t) - \hat{y}(t) \end{cases} \tag{5.10}$$

In the meantime, in order to drive the output to the setpoint value, terminal weights are used by adding a constraint to support the optimisation problem. The adding term to the objective function shows in the following form:

$$\varepsilon_T = e(t+N_p)Q_T e(t+N_p)^T \tag{5.11}$$

Where, Q_T is the weighting matrix with larger weights than Q. For every optimisation step, the system controller model is called to calculate the cost function.

5.3.2 MPC tuning

Garicia et al (1989) have indicated that tuning an MPC controller consists of many trials and errors in most cases. The prediction horizon (N_p) normally equals to the settling time of controlled outputs. So in this evaporation simulation model, the prediction horizon is set of 10-time steps and the control horizon (N_c) is 2-time steps.

The MPC Simulink block is developed based on the equations, parameters and diagram structure to control the first effect product flowrate and temperature. The structure details have been shown in figure 5.2 as below.

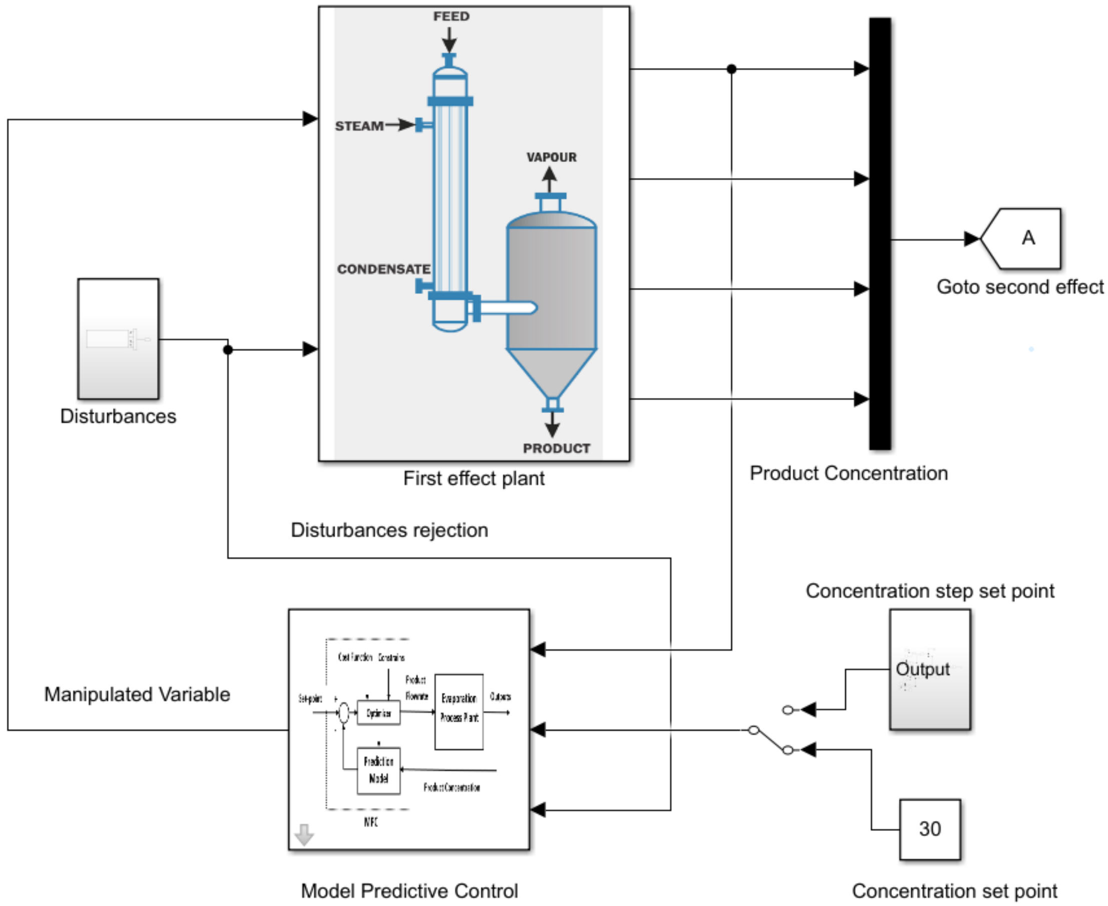


Figure 5.2 MPC MATLAB simulation blocks

There are not much literatures in how to selecte or design the best weight matrix Q and R for the MPC. Mostly used method is the trail and error to solve the problems by trying different values repeatedly. In this thesis, An MPC designer provided by MATLAB (toolbox) is applied to deal with the controller opzimization problem, specifically, the quadratic program (QP), at each control interval in order to support the controller development. It works by setting up the constrains of the plant inputs and outputs, and the inputs weight rate to call command to help with the cost function weight selection rather than selecting Q and R weight matrixs. For the constrains, product temperature must be less than 80℃ at any time during the evaporation process to protect the milk quality. Output concentration values are within ±0.2% of the desired set point.

5.4 Simulation Results with MPC

5.4.1 Setpoint tracking

The setpoint change is aiming at obtaining the MPC's tracking performance and testing its capture ability. In order to obtain more comprehensive situations, two trails with different setpoint value ranges of the product concentration are developed in the first effect. Both setpoints tracked curves are designed as a square wave. Trail 1 starts with an initial value of 5, increases to 30 at 240 seconds and then drops back to 5 at 720 seconds. Trail 2 is the same with trail 1 except for the initial value 20.

The total running time for both trails are 960 second and the model sampling time are 0.1s. Table 5.2 shows the setpoint change details.

Table 5.2 Setpoint change details

	Setpoint value change (%)	
Step time (Second)	**Trail 1**	**Trail 2**
240	5 to 30	20 to 30

720	30 to 5	30 to 20

According to figure 5.3, different disturbances have almost no influence on the settling time and overshoot. It only affects the error difference. So in the setpoint tracking performances, the simulation disturbance level was regulated at 20% for all the following results in this section.

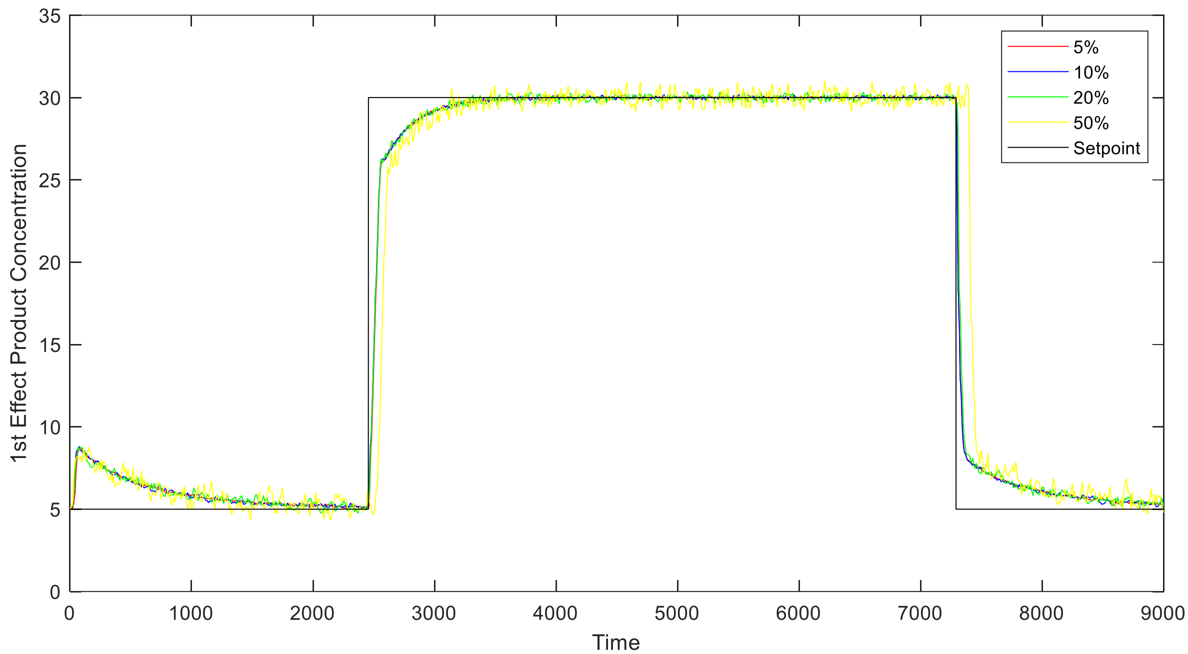


Figure 5.3 Setpoint tracking of 1st effect product concentration with different disturbance level

Figure 5.4 below shows the response of the MPC and PI controllers to the simultaneous setpoint step changes in product concentration for the first effect. The results have indicated that the different setpoint value range affects the tracking performance of the controllers. For both PI and MPC, more settling time is necessary with a large value range, but has less overshoot problems. On the contrary, by giving a small value change, both curves settle down more quickly but produce large overshoot values.

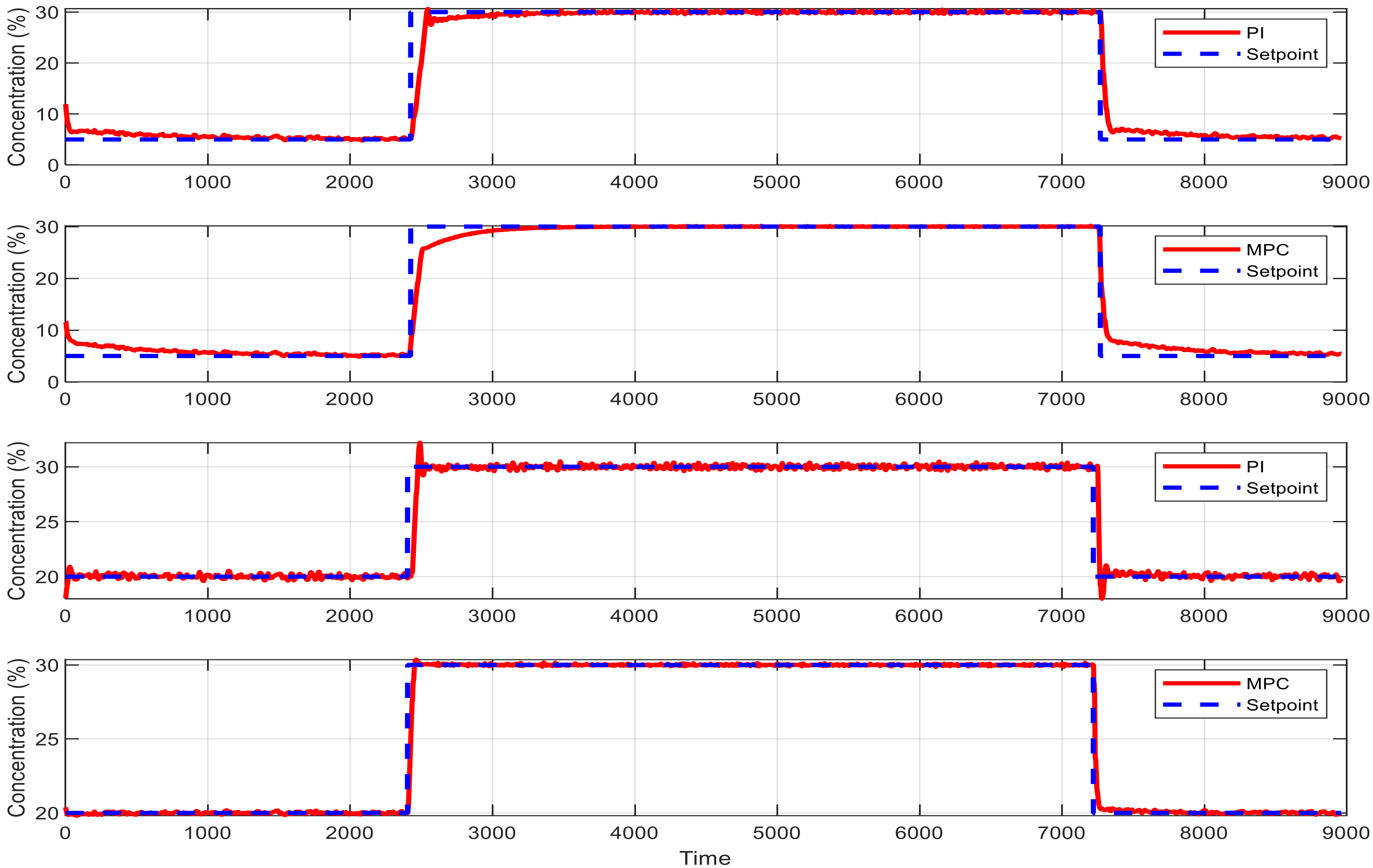


Figure 5.4 MPC and PI setpoint tracking response for a step-change in the production concentration

Overall, it is obvious from the figure that the setpoint changes occur at 240 seconds from 20% to 30%, but the settling times are different. The basic PI controller takes about 50 seconds to settle down and fluctuate around the desired setpoint value. Comparing with the PI strategy, the MPC controller exhibits much less overshoot and settles down more quickly than the PI controller.

The simulation results also show that both PI and MPC can achieve setpoint tracking when there are sharp value changes. In addition, no obvious offset in the control response is observed from figure 5.4 which means the errors can almost be neglected between the plant and the simulation model.

5.4.2 Constant Setpoint

In the real industrial process, a clear and certain control target is very important in the manufacturing plant. By achieving this in the simulation model, a desired constant setpoint was applied to obtain the accuracy of different controllers. The three desired control targets of product concentration for the three effects have been indicated in the previous chapter, which are 30%, 38%, and 52%. Different disturbances level (5%, 10% 20% and 50%) were added to the input signals to test the controllers' performances.

The following three figures (i.e. Fig. 5.5, 5.6 and 5.7) illustrated the MPC's performances of each effect with different disturbances levels. In order to make sure the figures are clear, 1000 random samples were taken from the simulation results.

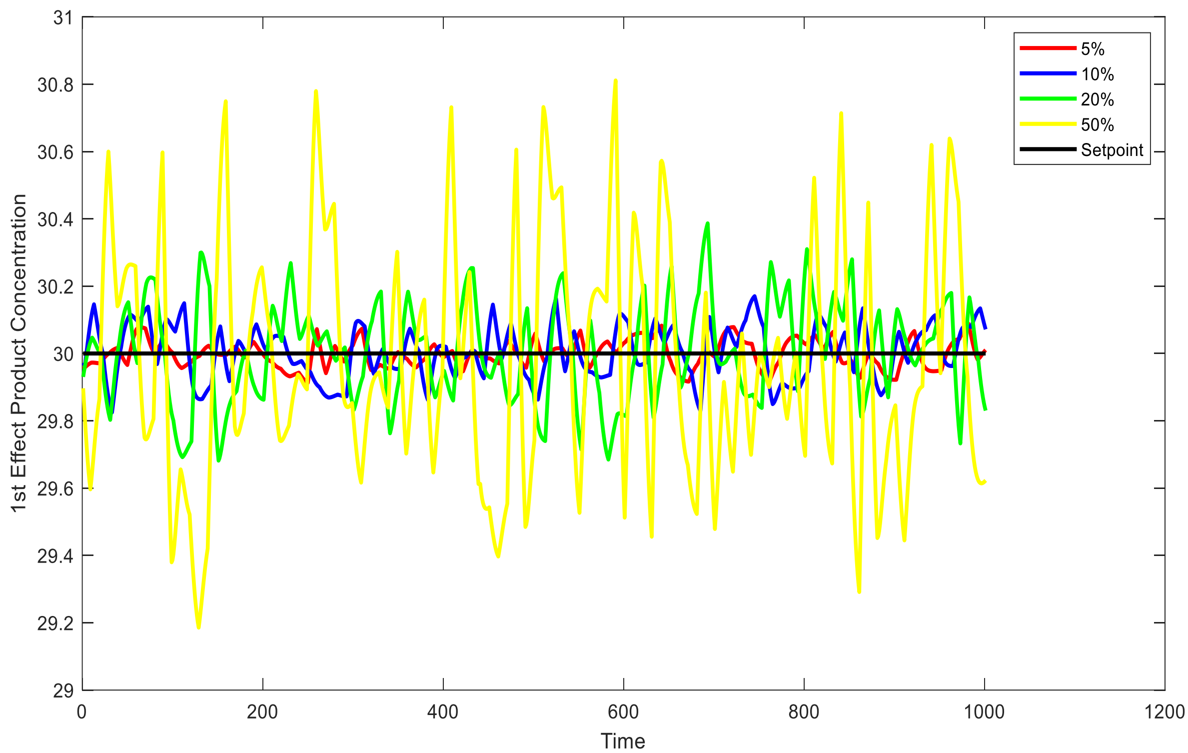


Figure 5.5 1st effect production concentration in a different disturbance with MPC

The MPC can achieve the desired control objectives to maintain the product concentration within a small fluctuating error of three effects under the same disturbance level. Even with the maximum 50% (normally 5% to 10% in the industrial plant) of the simulation input value-added as disturbances, the results are still within an acceptable range.

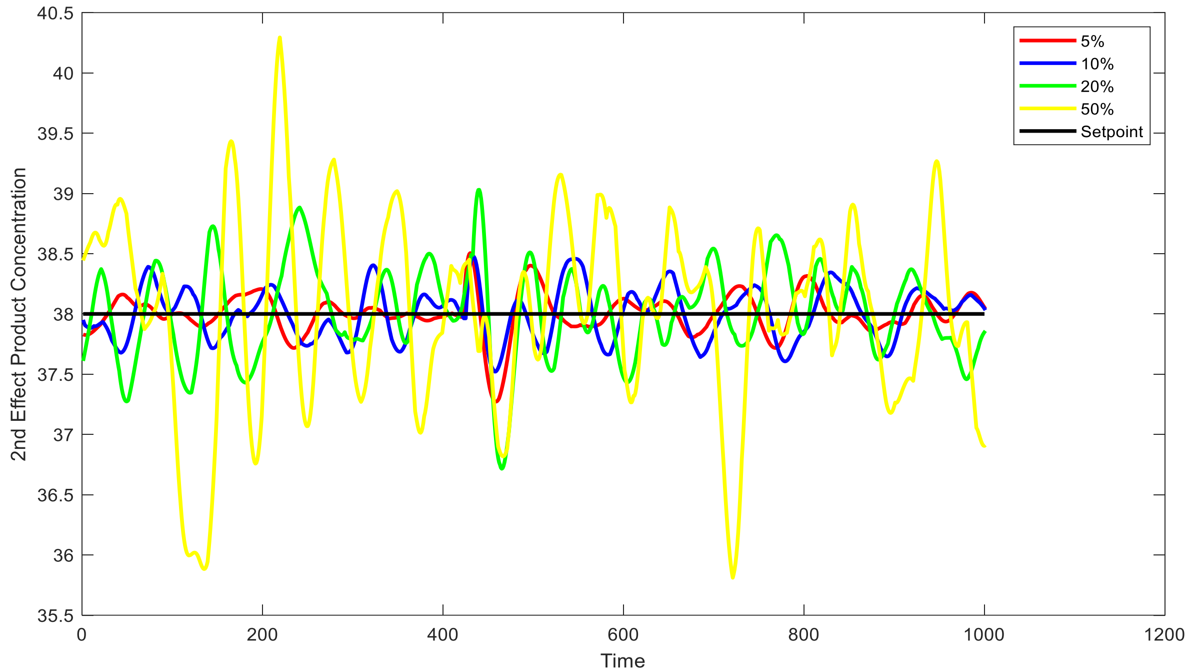


Figure 5.6 2nd effect production concentration in different disturbances with MPC

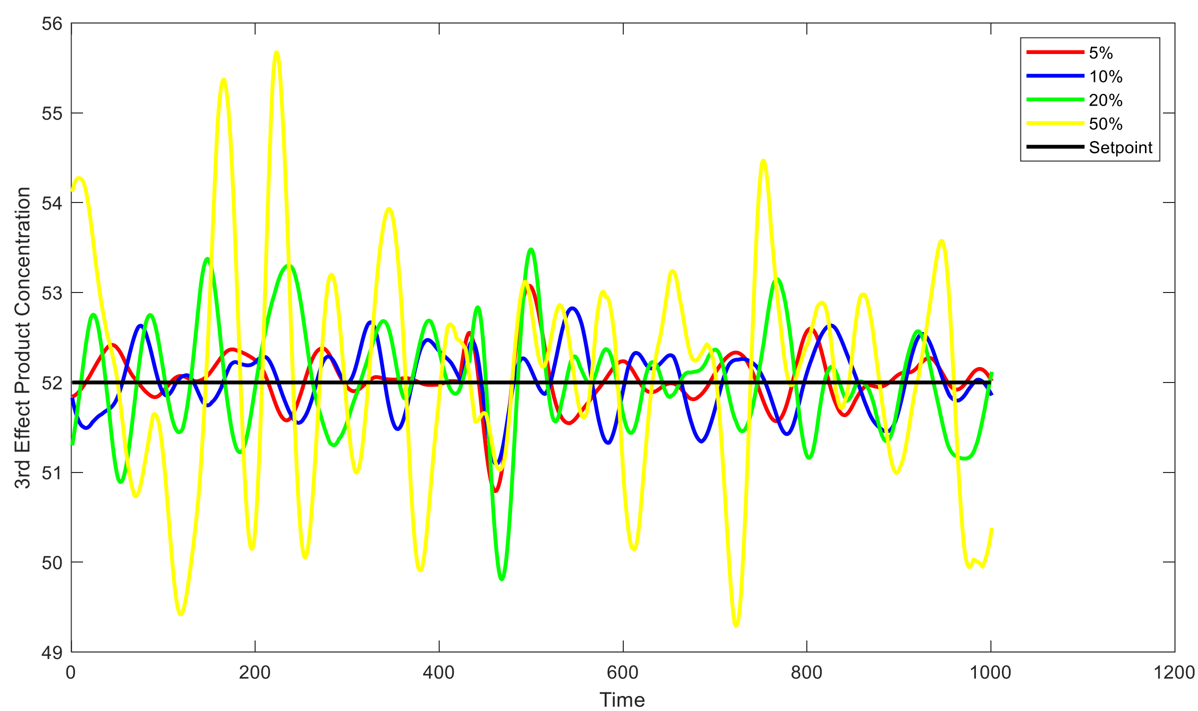


Figure 5.7 3rd effect production concentration in different disturbances with MPC

The increasing disturbance level can result in larger errors and fluctuations, which can be observed in the above figures of each effect. Meanwhile, both the figures and table 5.3 have indicated that under the same disturbances, the errors and fluctuation of the 1st effect are the smallest, then grow from the 2nd to the 3rd effect. One of the main reasons could be the accumulation of the variables error in the ahead effect impacts the controllability.

Table 5.3 Mean error and standard deviation in different disturbance level of each effect

	5%		10%		20%		50%	
	M error	Std	M error	Std	M error	Std	M error	Std
E1	0.0011	0.0388	0.0043	0.0754	0.0066	0.148	0.0232	0.3357
E2	0.0031	0.168	0.0045	0.2131	0.0162	0.3824	0.0179	0.7968
E3	0.0346	0.3047	0.0177	0.3555	0.024	0.6163	0.0647	1.2756

The error and standard deviation values are listed in table 5.3 above. The capital letter ‘E’ and ‘M’ stands for ‘effect’ and ‘mean’, ‘Std’ means standard deviation. The numbers in the table show that the MPC performs excellent for the simulation process control with 5%, 10% and 20% of the input variables values added as the system disturbances. Even with 50%, the results are still acceptable.

5.4.3 Results Comparison

Comparing with the PI control strategy, figure 5.8 indicated that the curve controlled by MPC has much smaller vibration.

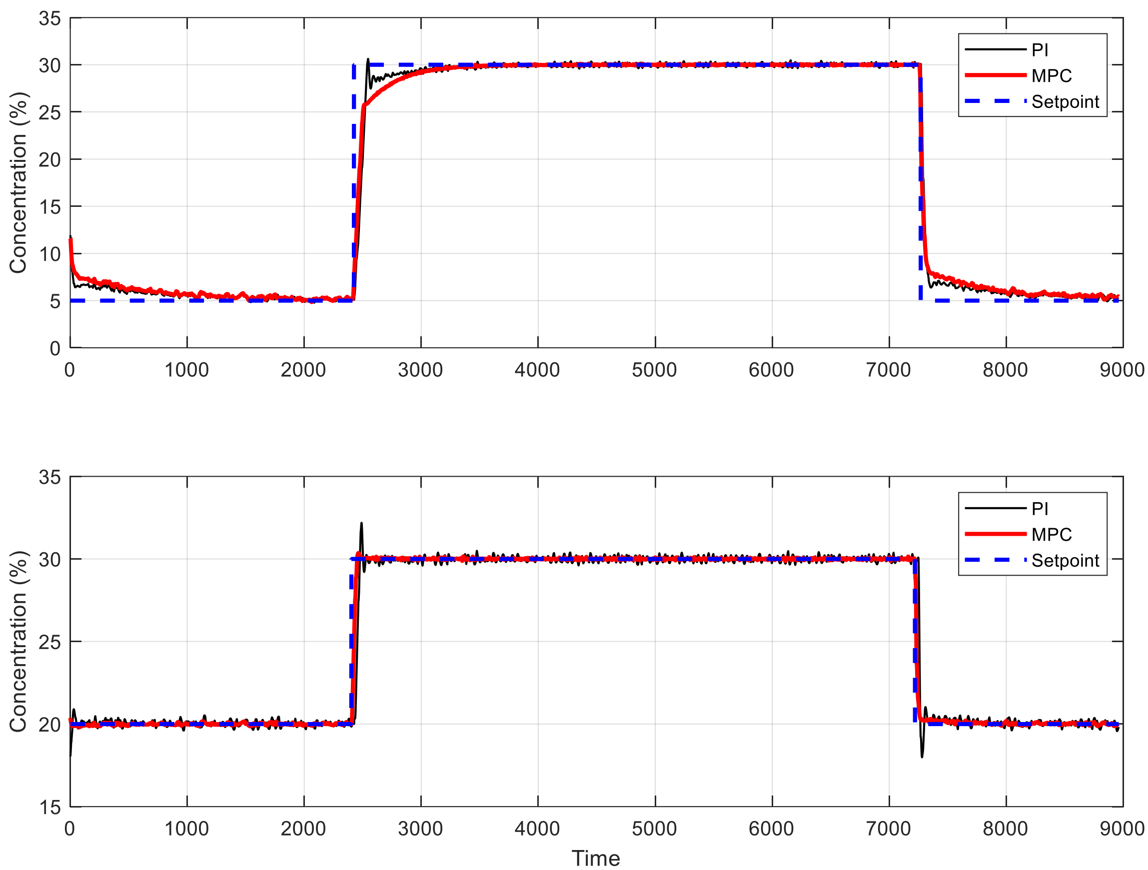


Figure 5. 8 Comparison of MPC and PI setpoint tracking response for step-change in product concentration

From the figure above, although both controllers provide stable and satisfactory performances, the results controlled by MPC are favourable to the production concentration. The standard deviation (Table 5.4) and the mean square errors

(Table 5.5) for the responses of both control strategies also show the improved control attained by the MPC.

Table 5. 4 Standard deviation for setpoint tracking test

Concentration setpoint change	PI control	MPC
From 5 to 30	10.531	7.1258
From 20 to 30	2.220	1.333

The mean square errors (MSE) here is used as an estimator to evaluate the errors between the simulation results and the desired values under MPC and PI controller. It is always non-negative and the closer to zero, the better. The following equation is applied to calculate the MSE value:

$$MSE = \frac{\sum_{i=1}^{n}[x_i - \hat{x}_i]^2}{n} \quad (5.12)$$

Where n is the sample size, x_i and $\hat{x}_i$ are the i^{th} modelling value and desired value respectively. The values show below:

Table 5. 5 Mean square errors for setpoint tracking test

Concentration setpoint change	PI control	MPC
From 5 to 30	4.9968	4.1971
From 20 to 30	0.4933	0.2616

5.4.4 Temperature and product flowrate results

Russell (1997) pointed out that the temperature remains stable in each effect because in a manufactory plant, the growth of concentration from effect to effect is well designed and the product flowrate through the effect is smooth in a relatively certain speed. However, the errors do increase through the evaporation process due to the accumulation. In this thesis, in order to observe and control the temperature responses, setpoint change of product concentration is designed from 28% to 32% (similar to the setpoint changes from 20% to 30% in Table 5.2) with 5% and 50% system disturbances respectively in the first effect.

Results controlled by PID and MPC are compared and shown in the two figures below.

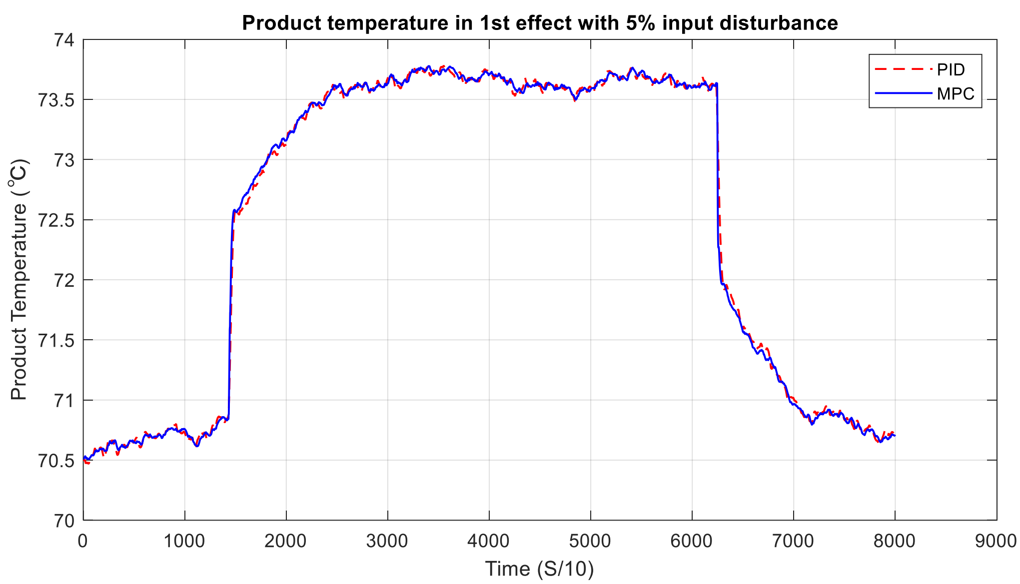


Figure 5.9 1st effect product temperature response to the concentration setpoint change with 5% disturbances

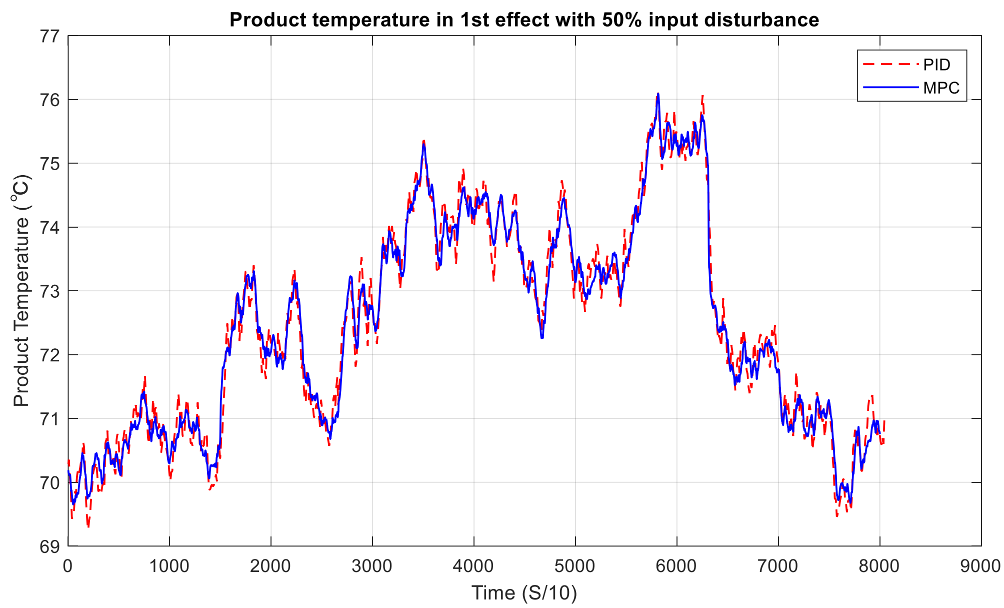


Figure 5.10 effect product temperature response to the concentration setpoint change with 50% disturbances

From both figures above, it is indicated that both PID and MPC provide good performances to control the temperature within a reasonable range (normally between 70℃ to 80℃ for the first effect). Additionally, with the increasing of the product concentration demands, the higher product temperature is needed to evaporate more water.

However, in order to maintain the product temperature stable, minimise the system disturbances are very essential. It is obvious that the product temperature in figure 5.10 is not as flat as its in figure 5.9. More stable product temperature makes the system easier to control and operate in the manufacturing plant.

Product flowrate is another very important variables that can affect the concentration. Because the flowrate can decide that how long the materials are heated in each effect which result in a different removal amount of water. Mostly, in an industrial evaporation process, the product flowrate reduces from the first effect to the next.

Figure 5.11 shows the product flowrate in each effect with concentration setpoint change from 28% to 32%.

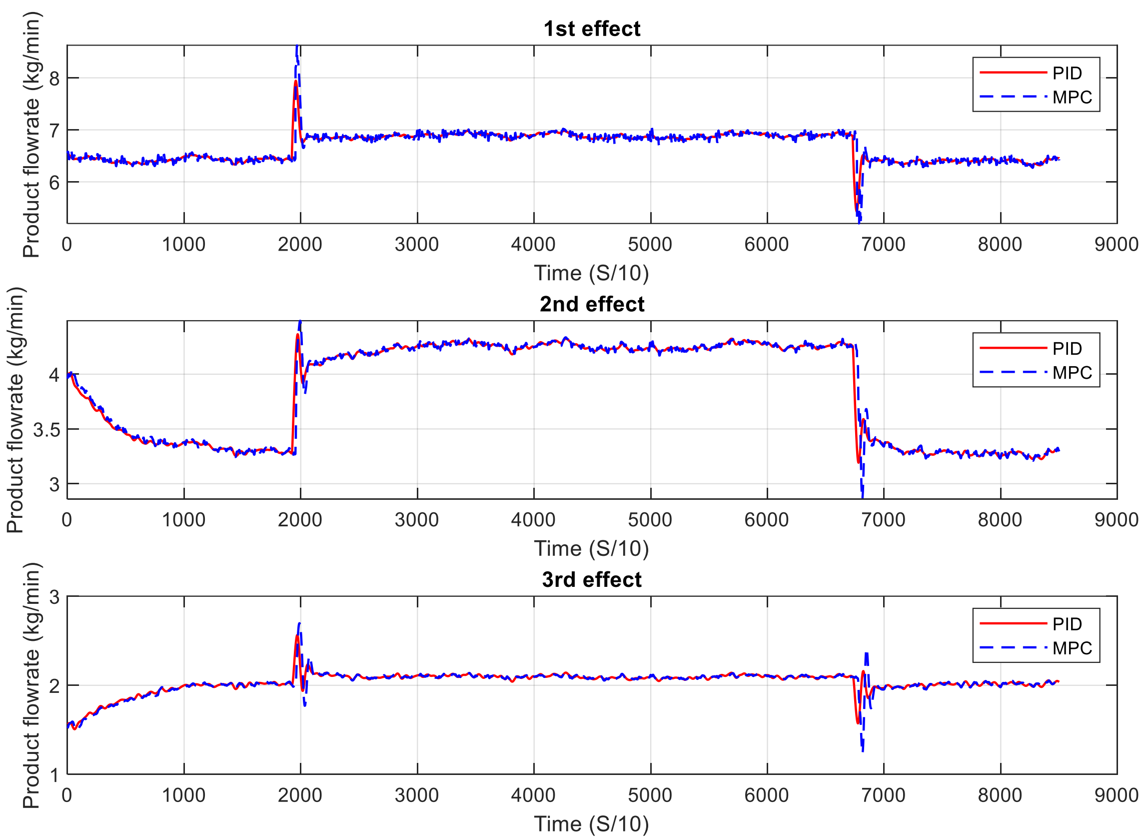


Figure 5.11 Product flowrate of each effect response to concentration setpoint change

The figure above indicated PID can maintain smoother flowrate of the product in each effect, but MPC works better on the concentration accuracy.

As a controlled variable during the evaporation process, product flowrate must be maintained within a reasonable range to ensure the concentration output performance. More fluctuations potencially enhance the control difficulties and increase the operation cost. Based on the flowrate comparison results above, MPC strategy costs more on the pump and circulation control than the PID.

5.5 Disturbance Rejection

In the real industrial manufacturing process, the disturbance rejection performance of the controllers must be tested. Because the unexpected disturbances mostly have a negative influence on the system performance and product quality. For an evaporation process, the most unexpected disturbance comes from the steam supply. To test the disturbance rejection performance, an impulse is applied to the steam temperature as a disturbance. This disturbance represented an unexpected temperature increase in the stem supply. The impulse is assumed to have an amplitude of 5℃ and added to the steam temperature at t=300 second.

Figure 5.12 described the output responses for the first effect product temperature for two control scheme with added disturbance. It is clear that MPC performs better in rejecting a sudden disturbance than the PI controller.

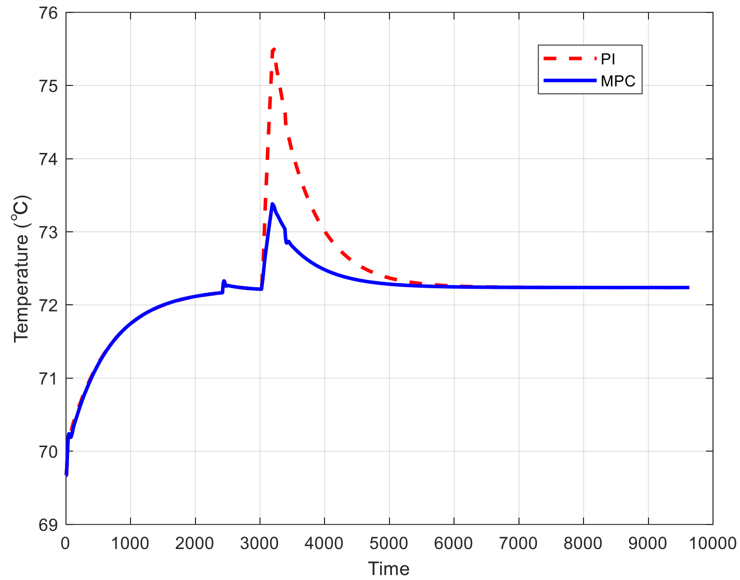


Figure 5.12 Disturbance rejection results controlled by PI and MPC

Meanwhile, other unexpected disturbances can be produced due to device error, control overshoot, transport delay or simulation environment, and finally enhance the difficulties of the system control. Different percentages (5%, 10%, 20%, and 50%) of the input variables value were added to the input signals to imitate the unexpected disturbances in real industrial plants. Results controlled by MPC and PI will be compared in order to obtain the ability to deal with system disturbances.

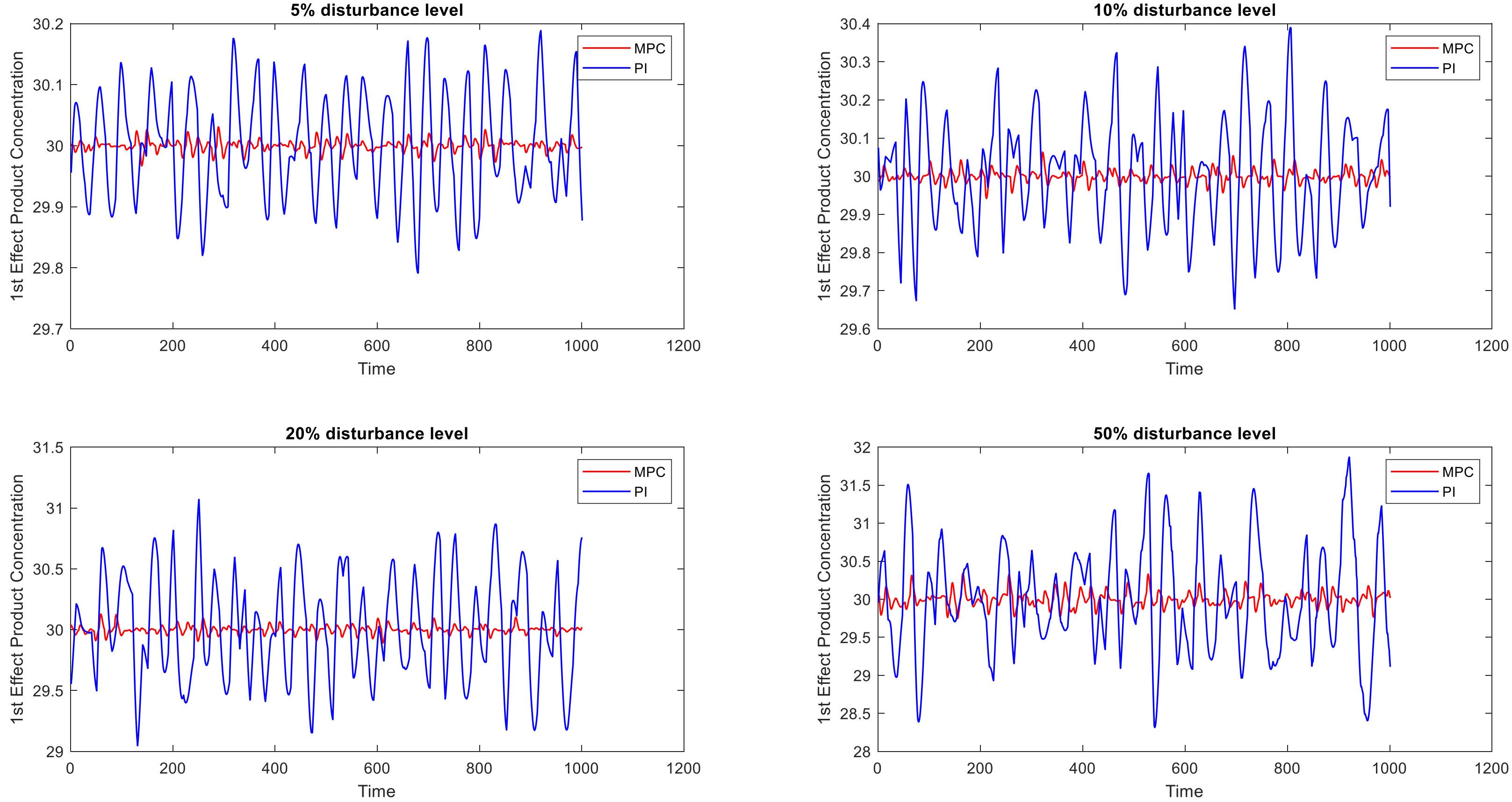


Figure 5.13 Comparison of MPC and PID performances on product concentration control with 5%, 10%, 20% and 50% disturbance level of the 1st effect

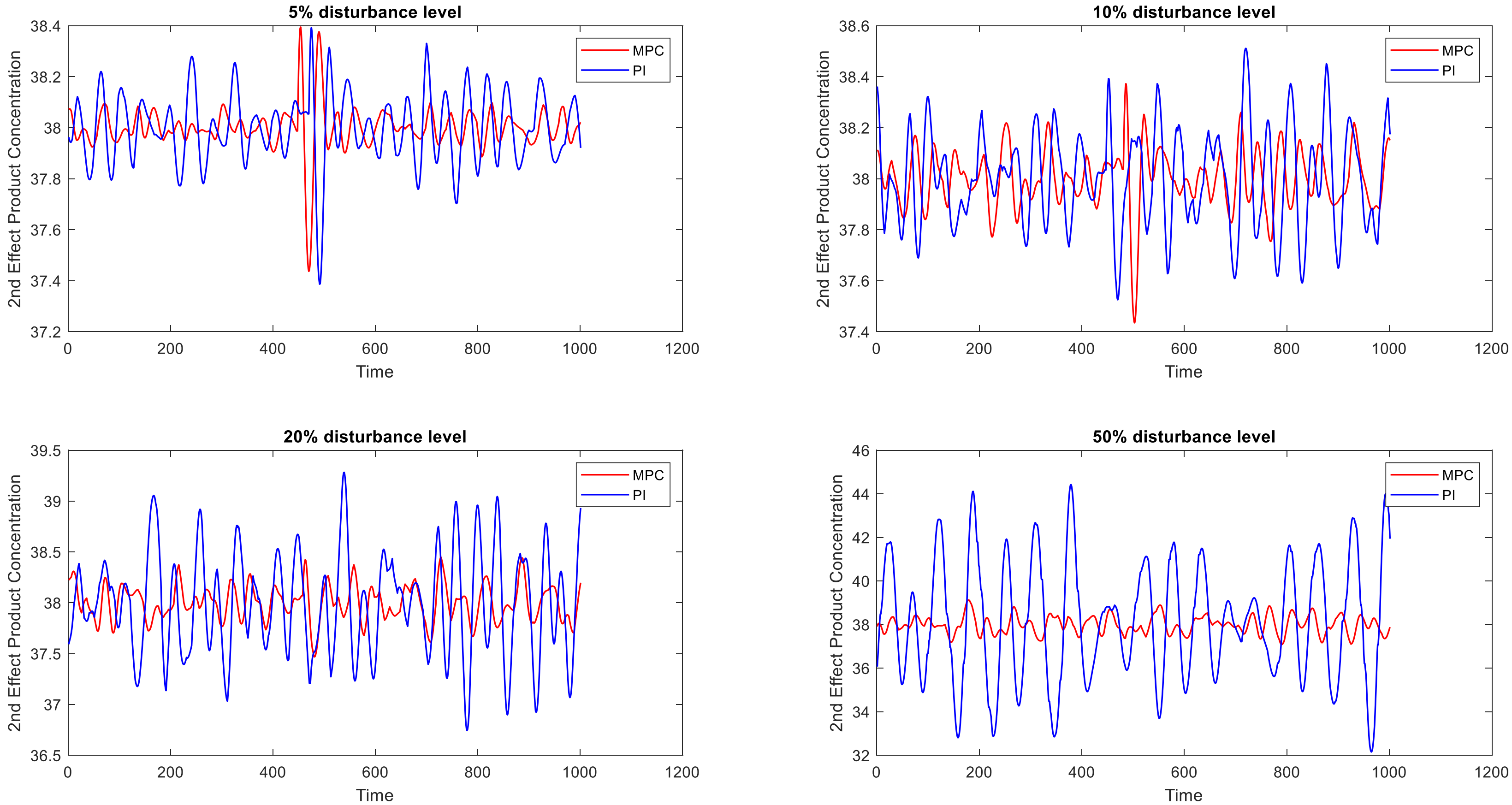


Figure 5.14 Comparison of MPC and PID performances on product concentration control with 5%, 10%, 20% and 50% disturbance level of the 2nd effect

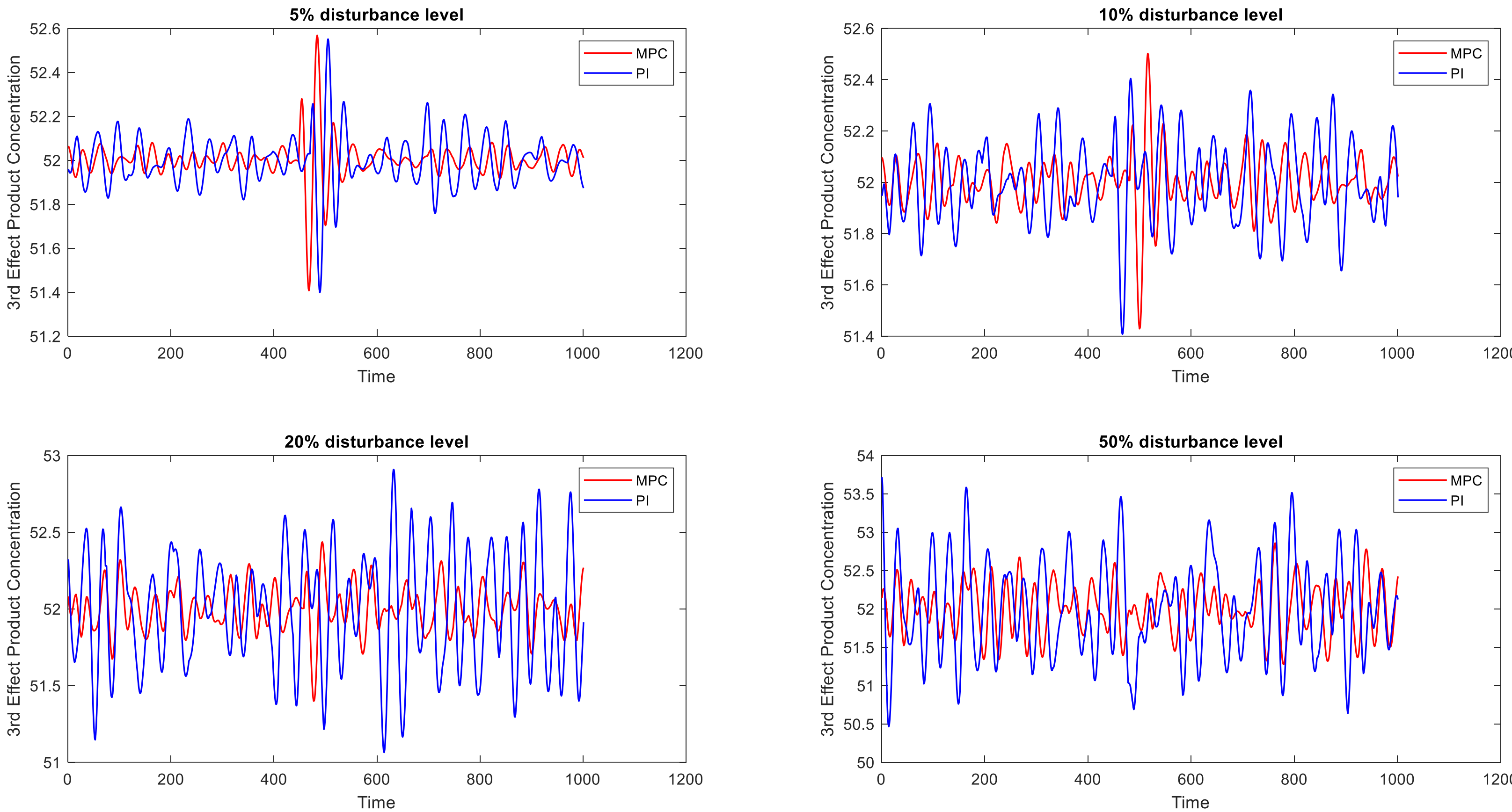


Figure 5.15 Comparison of MPC and PID performances on product concentration control with 5%, 10%, 20% and 50% disturbance level of the 3rd effect

The simulation results are shown in the three figures above (Fig. 5.13, 5.14, and 5.15). The figures indicated that MPC performs better than PI on disturbance rejection, especially when the unexpected disturbances are large.

5.6 Conclusions

The application of the MPC strategy has been demonstrated on the three effects of falling film evaporator simulation. Comparing with the PI controller, both methods can achieve the control target and maintain the simulation performances within an acceptable range. No obvious offset in the control response is observed in product concentration for both control strategies, which means the simulation model replicates the evaporation process very well, almost no errors between them.

However, MPC response appears to be better than that of the PI response of setpoint change tracking in product concentration. Although the MPC controller possesses a similar control performance on the settling time with PI in the setpoint tracking response, it has less overshoot problem.

In addition, PID and MPC have similar performance on the first effect temperature control. But lower disturbances level maintain the temperature more stable to reduce the difficulties of system operation and control. On the other side, PID can maintain smoother flowrate of the product in each effect than MPC, but MPC works better on the concentration accuracy.

On the disturbance rejection performance, MPC can handle a sudden disturbance that occurs during the simulation process and performs smaller overshoot and quicker response in temperature control than the PI controller. The temperature overshoot values are approximately 1℃ and 3.5℃ respectively for MPC and PI (Fig. 5.12). It also has a better result on the concentration accuracy control, which was more obvious with a large disturbance during the simulation process, comparing with PI strategy.

Chapter 6 Fuzzy Logic Control of the Three-effect Falling Film Evaporator

6.1 Introduction

It is generally believed that the conventional control methods have achieved huge success in the process control field based on the exact mathematical modelling of both the process and the controllers. However, there are still much uncertainty and ambiguity in the process itself which makes it complicated and more difficult to control. At this time, an intelligent system is required to handle these problems. Fuzzy logic control is one of the most applied intelligent control strategies.

The fuzzy concepts, which include fuzzy logic and fuzzy set theory, were firstly introduced by Prof. Zadeh in the 1960s as a new framework to represent vagueness and imprecision in the natural language. Not like Boolean logic control restricted to 0 and 1, fuzzy logic provides uncertain boundaries set, called ‘fuzzy set’, which defines intermediate values, also known as ‘membership’, between 0 and 1 to avoid absolute true or false only. This control strategy works more like how human makes a decision.

However, this new concept was not accepted by scientists and engineers due to the vagueness occurred in the theory in the early 1960s. Mamdani (1974) demonstrated the industrial application of fuzzy logic to control a steam engine, since then the fuzzy logic control theory started to be applied to majority process control fields and gained vast applications successfully.

In this chapter, the fundamentals of fuzzy set and fuzzy logic theory will be introduced. Type 1 and type 2 fuzzy sets will be presented. Meanwhile, Mamdani Fuzzy Inference System and Takagi-Sugeno Fuzzy Model (TS Method) will be introduced, and Mamdani type fuzzy inference system will be developed in MATLAB and implemented as a fuzzy control strategies to maintain the system performances. Controlled results will be compared with PI and MPC.

6.2 Fuzzy Sets

6.2.1 The Concept of Fuzzy Sets

The fuzzy sets can be normally considered as a simple extension of the notion of classical sets. The classical sets only contain two memberships (true or false) to all the global elements. Equation 6.1 is a very typical example of a classical set. The two memberships are '1' and '0', and x is the element in the set. χ_A is the characteristic function of set A.

$$\chi_A(x) = \begin{Bmatrix} 1, & if\ x \in A \\ 0, & if\ x \notin A \end{Bmatrix} \tag{6.1}$$

This classical set represents absolute truth and falseness. But in a fuzzy set, every object has a grade of membership in the set. The grade of membership arbitrarily runs from zero to one and is represented by a membership function $M(x)$. Every object x has its membership function value. Figure 6.1 shows the difference of two sets more directly.

The fuzzy set extends the range of variability of the membership function from the two-point set {0, 1} to the whole interval [0, 1]. That is why fuzzy logic is considered much closer in spirit to human thinking.

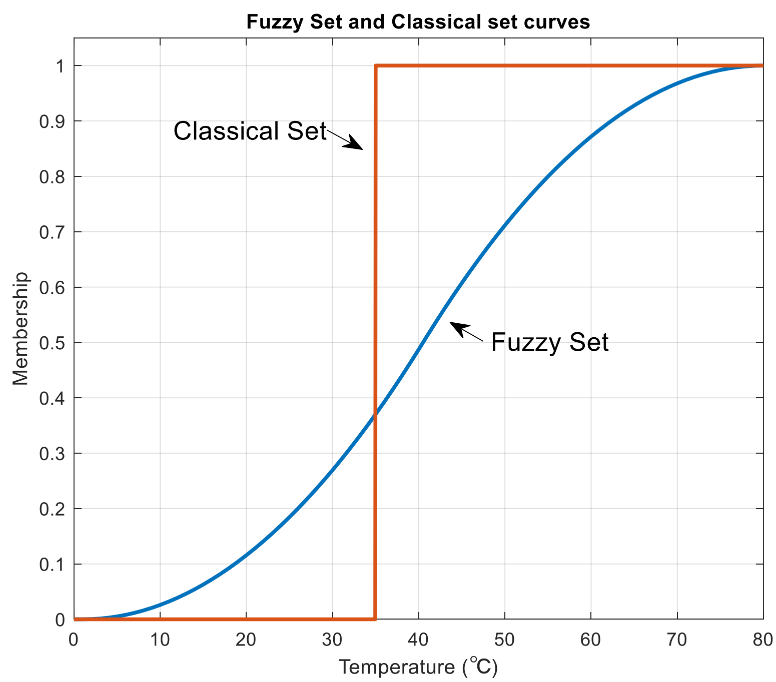


Figure 6.1 Fuzzy Set versus Classical Set curve

The general fuzzy sets form is very simple to be described by the equation below:

$$A = \{(x, \mu_A(x))\} \qquad (x \in X) \tag{6.2}$$

The A in Eq 6.2 represents the fuzzy set in the universe of discourse X, which is defined by a real value function μ_A, also known as membership function. It is used to calculate the degree of membership of x and indicate to which extent x belongs or is a member of the fuzzy set A. The value is between 0 and 1.

6.2.2 Type 1 and Type 2 Fuzzy sets

There are basically two different approaches to FLC design: type 1 and type 2. The general form of a fuzzy set (Eq 6.2) belongs to type 1 because that all the degree of membership ($\mu_A(x)$) can be calculated to acquire a certain value in this type of set.

However, only type 1 FLCs are not able to handle all the uncertainties, such as noisy data and changing environments, in the system. In the real industrial applications, there are eight sources of uncertainties which are very common in FLCs:

- The precision of the measurement devices
- Noise on the measurement devices
- Environmental conditions of the measurement devices
- Lack of modelling
- The meanings of the words that are used in the antecedents and consequents of rules can be uncertain (words mean different things to different people)
- Consequents may have a histogram of values associated with them, especially when knowledge is extracted from a group of experts who do not all agree.

- The uncertainty caused by some unvisited data that the fuzzy system does not have any predefined rules for.

Many types of research have proved that when there are a large number of uncertainties in the system, the type 1 FLCs mostly cannot achieve the desired control targets or the expected performances (Juang and Hsu, 2009; Khanesar et al, 2010; Biglarbegian et al, 2011). In order to deal with these large uncertainties, an improved fuzzy logic, type 2 fuzzy logic theory, was developed by Mendel and Karnik in 1999. The type 2 fuzzy set $\tilde{A}$ is represented as below:

$$\tilde{A} = \{((x,u), \mu_{\tilde{A}}(x,u)) \mid x \in X\ u \in J_x \subseteq [0,1]\} \quad (6.3)$$

Where $\mu_{\tilde{A}}$ is the type 2 fuzzy set membership function to calculate the membership degree $\mu_{\tilde{A}}(x,u)$ in which $0 \leq \mu_{\tilde{A}}(x,u) \leq 1$.

J_x in the Eq. 6.3 is called the primary membership of x. According to Mendel and Karnik's study, there is a secondary membership value corresponding to each primary membership value that defines the possibility for primary memberships. In the circumstances in which there are so many uncertainties, the membership degree determined by a type 1 fuzzy set crisp number in [0, 1] may not be precise. In these cases, applying type 2 fuzzy sets are a preferable option.

In summary, the most different part between type 1 and type2 fuzzy sets are the membership functions. All the membership functions of type 1 fuzzy sets are certain. However, the membership functions themselves of type 2 are fuzzy, which results in antecedents and consequents of the rules being uncertain. If the circumstances are so fuzzy, type 2 appears to be a more promising method than type 1 for handling uncertainties.

6.3 Fuzzy Logic Control System

6.3.1 Fuzzy logic controller structure

A fuzzy logic controller generally contains four components, fuzzifier, rules, inference, and defuzzifier. The following figure 6.2 indicated how these four parts linked together.

- Fuzzifier: it is used to convert crisp inputs values into fuzzy values. This process is also called fuzzification.
- Rules: also knows as Fuzzy Rule Base, it stores the knowledge (set of rules or/and if-then conditions) about the operation provided by the experts to manage the decision-making system.
- Inference Engine: Basically, it simulates human decisions to determine the matching degree of the current fuzzy input concerning each rule and decides which rules are to be fired according to the input field.
- Defuzzifier: the role of defuzzifier is exactly opposite to the fuzzifier which is to convert the fuzzy values into crisp values getting from the fuzzy inference engine. Then pass these crisp values into the process plant.

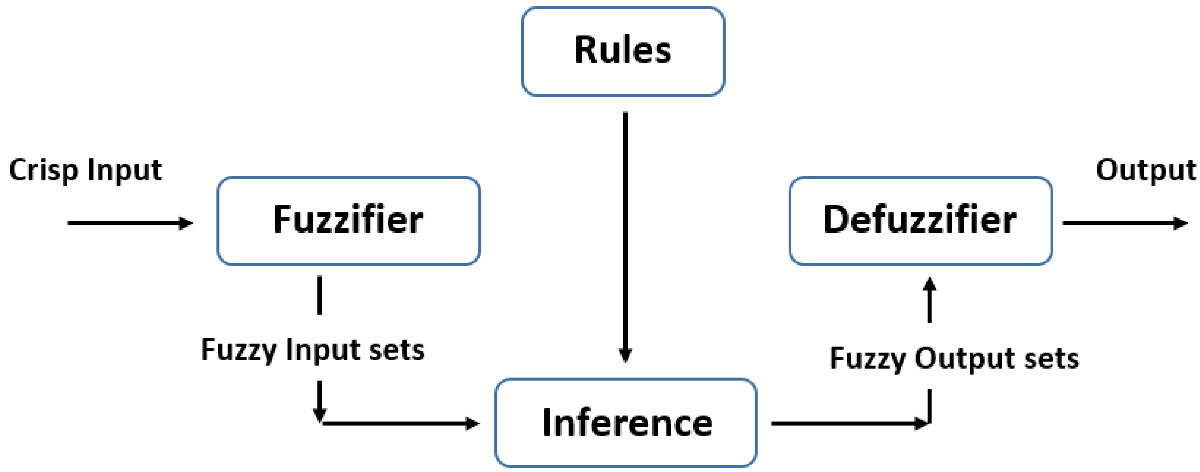


Figure 6.2 Fuzzy logic controller structure diagram

However, before designing a fuzzy control system, there are several basic assumptions should be considered:

- The plant is observable and controllable − It must be assumed that the input, output as well as state variables are available for observation and controlling purpose.
- Existence of a knowledge body − It must be assumed that there exist a knowledge body having linguistic rules and a set of input-output data set from which rules can be extracted.
- Existence of a solution − It must be assumed that there exists a solution.
- ‘Good enough’ solution is enough − The control engineering must look for ‘good enough’ solution rather than an optimum one.
- Range of precision − Fuzzy logic controller must be designed within an acceptable range of precision.

- Issues regarding stability and optimality − The issues of stability and optimality must be open in designing the Fuzzy logic controller rather than addressed explicitly.

6.3.2 The Membership Functions (MFs)

As we discussed above, the MFs play vital role in the fuzzy logci control process by determining the suzziness in a fuzzy set. They have several different shapes such as triangular, trapezoidal and Gaussian. ‘’The MFs can be of any shape and form as long as it maps the given data with desirable degree of memberships. The function itself can be an arbitrary curve whose shape we can define as a function that suits us from the point of view of simplicaity, convenience, speed, and efficiency’’ (Sadollah ,2018). However, only one requirement for the MFs is that the value range must be between 0 and 1. The triangular shape of MFs will be used in the fuzzy logic control in this thesis according to the problem size and type.

Inaddition, choosing the interval and number of MFs are important for a fuzzy logic controller, especially for the temperature control model. Mostly used fuzzy logic controllers have 3, 5, 7 or 9 MFs due to the system complexity and design requirements. Kosko and Mitaim (1996) indicated that more MFs increase the system computational time. Therefore, one of the system optimization methods for a fuzzy logic control is determing the number of MFs. In this thesis, a MATLAB fuzzy logic Simulink block is applied to help the controller development. Totally 9 input membership functions are designed to achieve the process control objectives.

6.3.3 Advantages and drawbacks

As discussed in chapter 2, there are many successful applications of fuzzy logic controllers in many different fields. But why fuzzy logic? Behrooz (2018) have pointed out several reasons:

- Imitate human thinking: this is the most important reason that fuzzy logic has been widely applied as a controller in industrial process control to make the system more intelligent. It is also useful in providing solutions to a complex system in many types of applications due to human reasoning and decision making characteristics.
- Relatively simple structure: the fuzzy logic structure is very simple (Fig. 6.2), which means It is easy to be constructed, and easy to understand as well.
- Robust: fuzzy logic is able to cover a huge range of operating conditions than other conventional controllers, and because of its robustness, no precise inputs are required. Even imprecise, distorted and error input information is also accepted by the system.
- Ambiguousness: basically fuzzy logic can perform very well when dealing with uncertainties, such as noise data, unexpected environment changes and so on.
- Flexible: intuitive knowledge base design and modification in the rules is allowed.
- Cheaper and efficiency: many studies have been proved that fuzzy logic strategies are comparatively cheaper than developing model-based or other

controllers in terms of performance, and provide more efficiency when applied in a control system.

However, there are some drawbacks of fuzzy logic controllers which could restrict the implementations:

- Accuracy: fuzzy logic controlled results are not always accurate enough due to its assumptions and lack of rules. This is the most concerning disadvantage, in most cases, in order to improve the accuracy, more fuzzy grades are needed to increase exponentially the rules.
- Ambiguity: many searches have indicated that by using fuzzy logic to solve a given problem can sometimes lead to ambiguity.
- Characteristics: It is difficult or even impossible to obtain the fuzzy logic controllers' characteristics because there is no mathematical description of the approach.
- Data: A lot of data and expertise are required to develop a fuzzy system.

6.3.4 Two Fuzzy Logic Inference Systems

6.3.4.1 Mamdani fuzzy logic system

As discussed, the fuzzy logic control method has achieved huge successful applications in many industrial processes. Among all these applications, there are two popular and widely used fuzzy inference systems called Mamdani and Takagi-Sugeno (TS) fuzzy logic systems.

Mamdani fuzzy system was firstly proposed in 1975 by Ebhasim Mamdani to imitate the human expertise behaviours in charge of controlling a steam and boiler combination system. The original purpose was to summarize the system operator's experience into a set of (linguistic) IF-THEN rules that could be used by a machine to automatically control the process.

For simplicity, the IF-THEN rules form is:

{ IF X is A, Then Y is B }

X and Y represent the inputs and outputs in the fuzzy system which could be physical variables, such as pressure, volume, temperature. A and B are the conditions and results which could be certain numbers or descriptive words like 'high', 'medium', 'big' or 'small'.

Mamdani systems are normally composed of several IF-THEN rules to carry out the fuzzy control. By using these rules, a function ϕ is defined to generates the numerical output y from the input values x ($y = \phi(x)$). Each rule may use different fuzzy set A_k and B_k (assume the index with subscript k). A standard Mamdani system is computing the truth value (TV) of combined propositions based on the following rules (Izquierdo, 2018):

TV (X is C or X is D| X=x)=max(TV(X is C| X=x), TV(X is D| X=x)) = max($\mu_C(x), \mu_D(x)$);

TV (X is C and X is D| X=x)=min(TV(X is C| X=x), TV(X is D| X=x)) = min($\mu_C(x), \mu_D(x)$).

Ion Iancu (2012) illustrated the steps of applying a Mamdani fuzzy inference process:

- Fuzzification: This is the first step for most of the fuzzy logic systems to transfer the crisp inputs into fuzzy inputs.
- Evaluate the antecedent for each rule: The membership values can be computed with given crisp input values. But when evaluating the antecedent for each rule, if the antecedent of the rule has more than one part, a fuzzy operator is applied to obtain a single membership value. Then the antecedent evaluation can be applied to the membership function of the consequent.
- Aggregation of the rule outputs: The membership function of all rule consequents previously clipped or scaled are combined into a single fuzzy set by using a fuzzy aggregation operator. The most commonly used aggregation operators are the maximum, the sum and the probabilistic sum.

- Defuzzification: The aggregated fuzzy set is transformed into one single crisp number in this step. In order to solve a decision problem, a certain value of the final output is more preferable than the fuzzy output.

6.3.4.2 Takagi-Sugeno (TS) Fuzzy Logic Systems

Another widely applied fuzzy logic system is developed by Takagi Sugeno and Kang in 1985. This model is described by fuzzy IF-THEN rules which represent local input-output relations of a nonlinear system. The main feature of a TS fuzzy model is to express the local dynamics of each fuzzy implication (rule) by a linear system model. The overall fuzzy model of the system is achieved by fuzzy 'blending' of the linear system models.

A typical fuzzy rule in a TS fuzzy model has the form:

$$\{ \text{ if } X \text{ is } A, \text{ and } Y \text{ is } B, \text{ then } z=f(X, Y) \}$$

Where A and B are fuzzy sets in the antecedent. $z=f(X, Y)$ usually represents a crisp function in the consequent, but it can be any appropriate functions which can describe

The output of the system within the fuzzy region specified by the antecedent of the rule. There is a specific situation that when the function is a constant, TS can be treated as a special case of the Mamdani fuzzy system.

The TS fuzzy inference process works in the following way:

- Fuzzification: same with the Mamdani inference system, transform the input form from crisp to fuzzy is the first step to make the system fuzzy.
- Computation: the fuzzy operator is applied to calculate the output of the inference system.

The most fundamental difference between Mamdani and TS fuzzy logic is the way the crisp output is generated from the fuzzy input (Kaur and Kaur, 2012). Mamdani-type uses the technique of defuzzification of a fuzzy output, while TS applies a weighted average to compute the crisp output. TS method contains more mathematical rules and has more adjustable variables than the Mamdani-type fuzzy logic system, which makes it more flexible in system design. Also, TS has a better performance on time-saving for the process because that weighted average can replace the time-consuming defuzzification process. However, Mamdani-type has certain output membership functions whereas the TS model does not. Due to these different features, the Mamdani-type fuzzy system is widely used in particular for decision support applications and the TS model works better with optimization and adaptive techniques.

6.4 Simulated Fuzzy Logic Control of the Evaporator

Similar to the control objectives of MPC, the fuzzy logic controller is also aiming to maintain the product concentration constantly and close to the desired setpoint value. Meanwhile, to reduce the different levels of disturbances and minimise the target vibration within an acceptable range are the other two control objectives. To achieve these objectives, product flowrate in each effect is treated as the controlled variable to regulate the final product concentration.

According to the fuzzy logic control structure (see Fig 6.2), the crisp inputs and the output in that figure represent the system feedback error of the concentration and the product flowrate in each effect, respectively. The error values are the concentration difference between the desired setpoint and the present value in the effect. The use of the error signal allows the controller to react to both disturbances and setpoint changes without a redefinition of the rules or reference sets.

Figure 6.3 below shows the input membership functions, involved nine linguistic terms to characterize each variable. The input error range is designed from -25 to 25 considering the actual industrial plant possibly conditions.

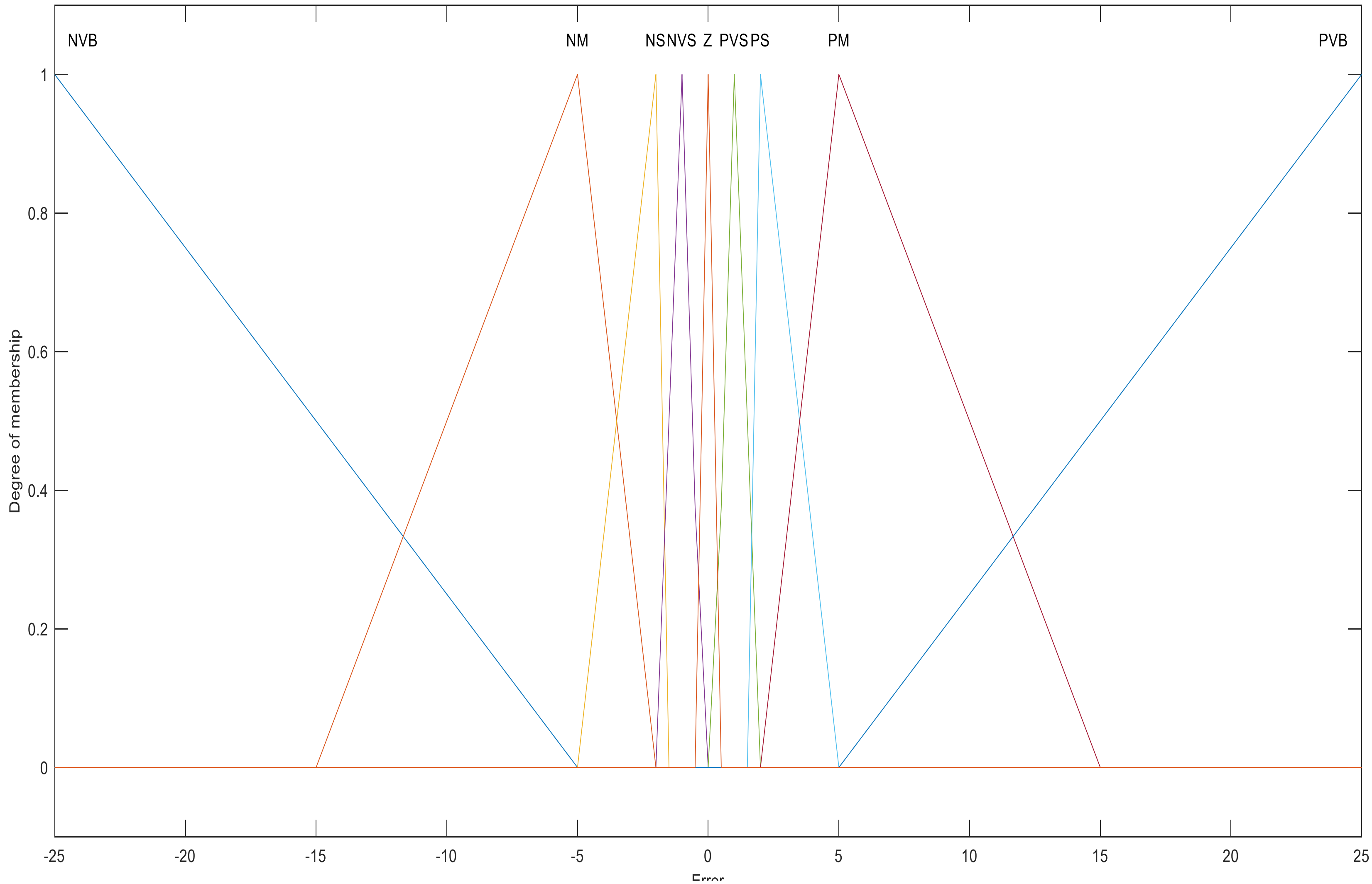


Figure 6.3 Fuzzy logic membership function of the input errors

The details of the input membership function are given in table 6.1.

Table 6. 1 Membership function of fuzzy logic controller input error

	Range	Fuzzy Set
NVG	[-30 -25 -5]	Negative very big
NB	[-25 -10 -5]	Negative big
NM	[-10 -5 -2]	Negative medium
NS	[-5 -2 -0.2]	Negative small
Z	[-0.2 0 0.2]	Zero
PS	[0.2 2 5]	Positive small
PM	[2 5 10]	Positive medium
PB	[5 10 25]	Positive big
PVB	[10 25 30]	Positive very big

Figure 6.4 illustrates the membership function of output changes of the product flowrate in the current effect with nine reference fuzzy sets defined and called: ‘NVBC (Negative very big change)’, ‘NBC (Negative big change)’, ‘NMC (Negative medium change)’, ‘NSC (Negative small change)’, ‘ZC (Zero change)’, ’PSC (Positive small change)’, ‘PMC (Positive medium change)’, ’PBC (Positive big change)’ and PVBC (Positive very big change)’.

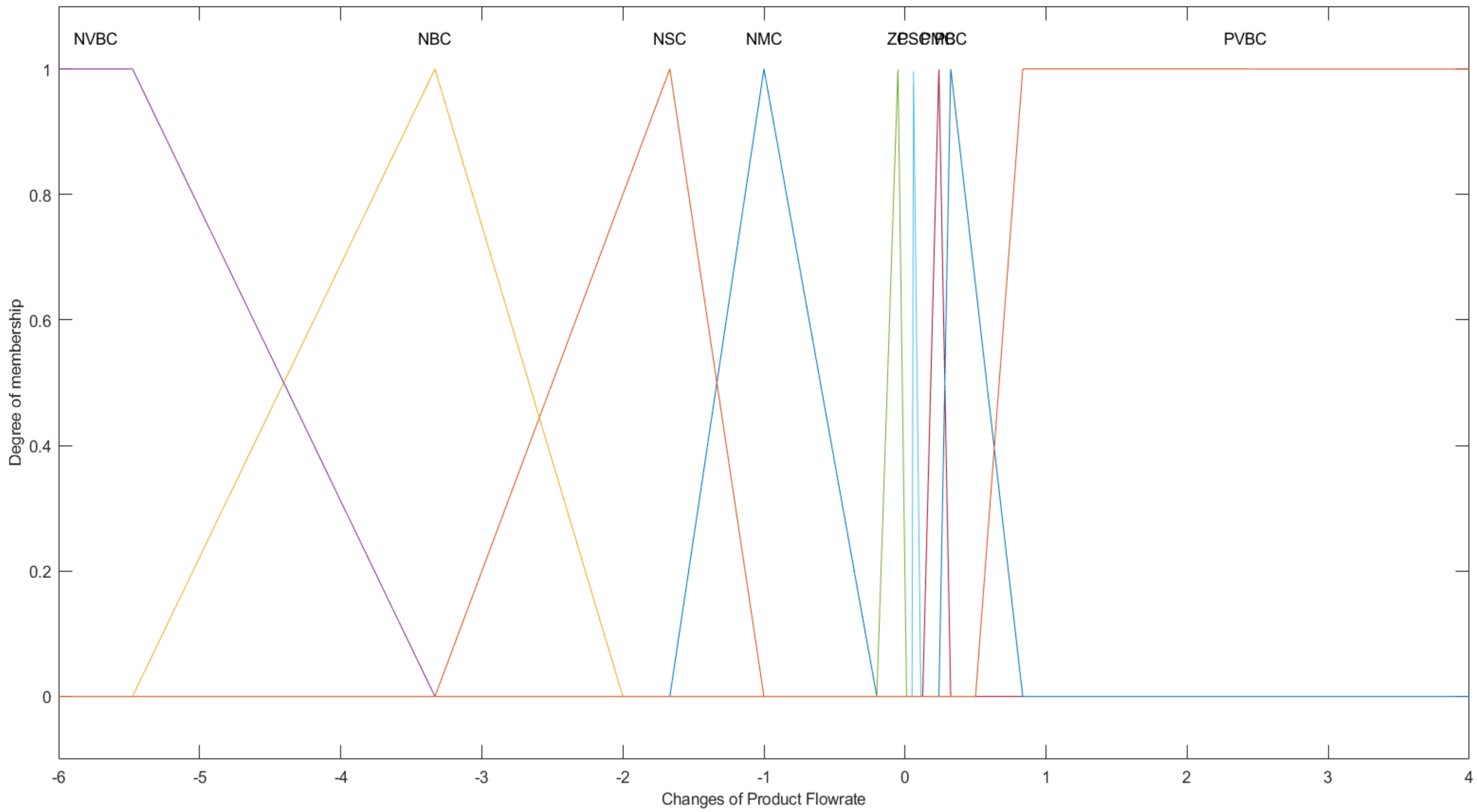


Figure 6. 4 Fuzzy logic membership function output

Different error ranges have a corresponding product flowrate change. The fuzzy logic rules set is:

1. If the error is NVB, then the product flowrate change is PVBC;
2. If the error is NB, then the product flowrate change is PBC;
3. If the error is NM, then the product flowrate change is PMC;
4. If the error is NS, then the product flowrate change is PSC;
5. If the error is Z, then the product flowrate change is ZC;
6. If the error is PVB, then the product flowrate change is NVBC;
7. If the error is PB, then the product flowrate change is NBC;
8. If the error is PM, then the product flowrate change is NM;
9. If the error is PS, then the product flowrate change is NS;

6.5 Fuzzy Logic Control Results

6.5.1 Constant Setpoint

The following three plots show the responses of the fuzzy logic controller to the designed constant setpoint (30, 38, and 52) on product concentration in each effect with different disturbance levels. 1000 samples have been selected randomly from each effect when the system output is stable.

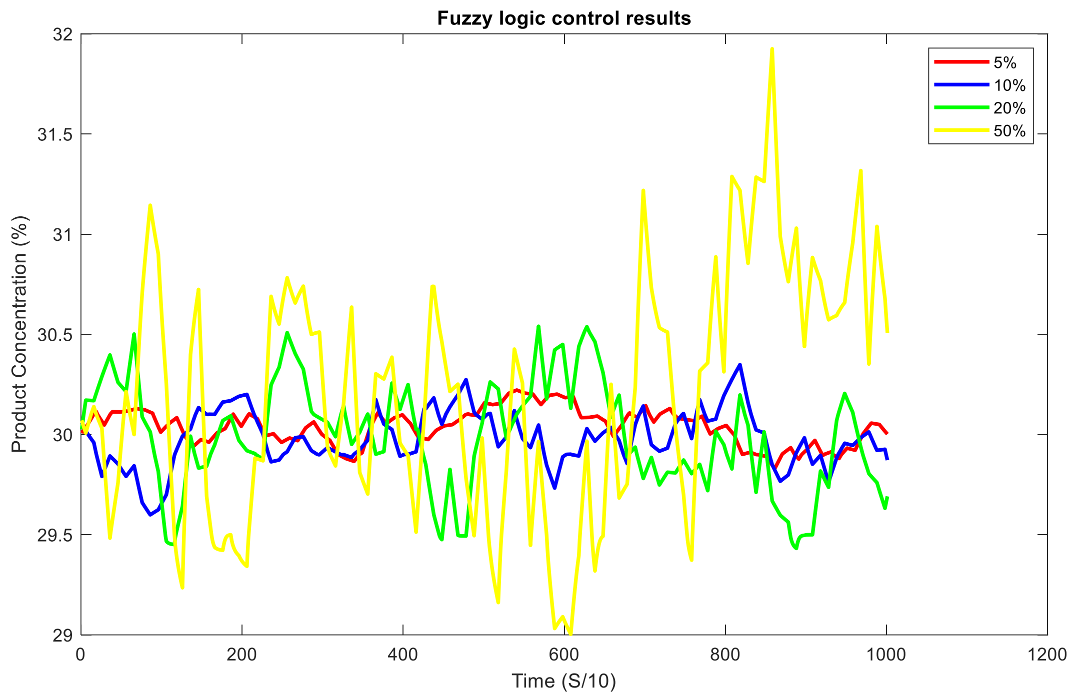


Figure 6.5 1st effect product concentration controlled by fuzzy logic with different disturbance level

Overall, with the increasing of disturbance percent, the performances of fuzzy logic are worse and worse. This is mostly due to the slow response of the fuzzy logic with the random disturbances and the disturbance accumulations. This has been indicated very obviously from figure 6.6 and 6.7 when the disturbance is more than 20% added, which the result is similar to the PID and MPC strategies'.

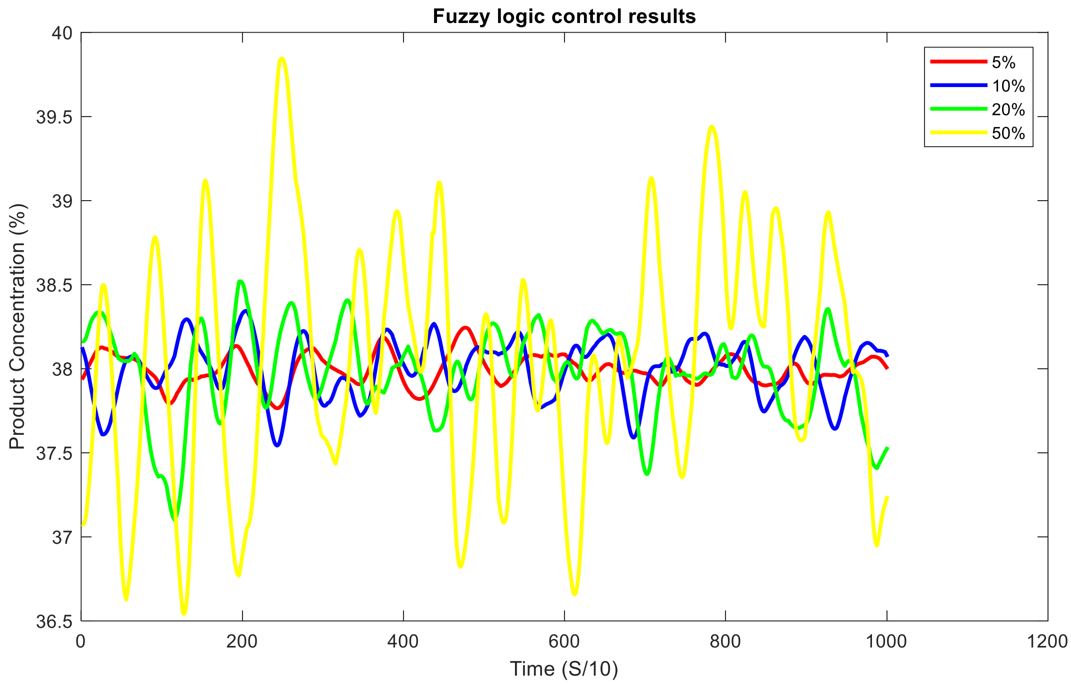


Figure 6.6 2nd effect product concentration controlled by fuzzy logic with different disturbance level

Besides, the simulation results also illustrated that fuzzy logic is not able to deal with the large random disturbance. According to figure 6.5, 6.6 and 6.7, once the input disturbance grows over 20%, the product concentration curves have larger amplitudes.

Figure 6.7 3rd effect product concentration controlled by fuzzy logic with different disturbance level

The simulation results in statistics are described in Table 6.2. A data set, which contains 2000 continuous samples, will be considered to compute the mean and standard deviation values when the system is stable. The mean error and standard deviation values in the table below reflect how close the simulated results to the designed targets and the fluctuation respectively.

The table shows the same results which have been pointed out based on the figures above. The bigger disturbances were added as an extra input to the system, the worse fuzzy logic performances were.

Table 6. 2 Mean error and standard deviation in different disturbance level of each effect controlled by fuzzy logic

	5%		10%		20%		50%	
	M error	Std	M error	Std	M error	Std	M error	Std
E1	0.0197	0.0774	0.0624	0.1991	0.1238	0.3257	0.2318	0.7156
E2	0.0315	0.0841	0.0972	0.2135	0.1561	0.3782	0.2586	1.0053
E3	0.0504	0.1604	0.1101	0.2968	0.3749	0.5487	0.346	1.5680

6.5.2 Setpoint Tracking Results

In order to obtain the target tracking performances of the fuzzy logic controller, and test its capture ability, setpoint changes have been applied as the control objective in the simulation process. Two setpoint change stages are the same as what has been done to the PI and MPC controllers' in the previous chapter.

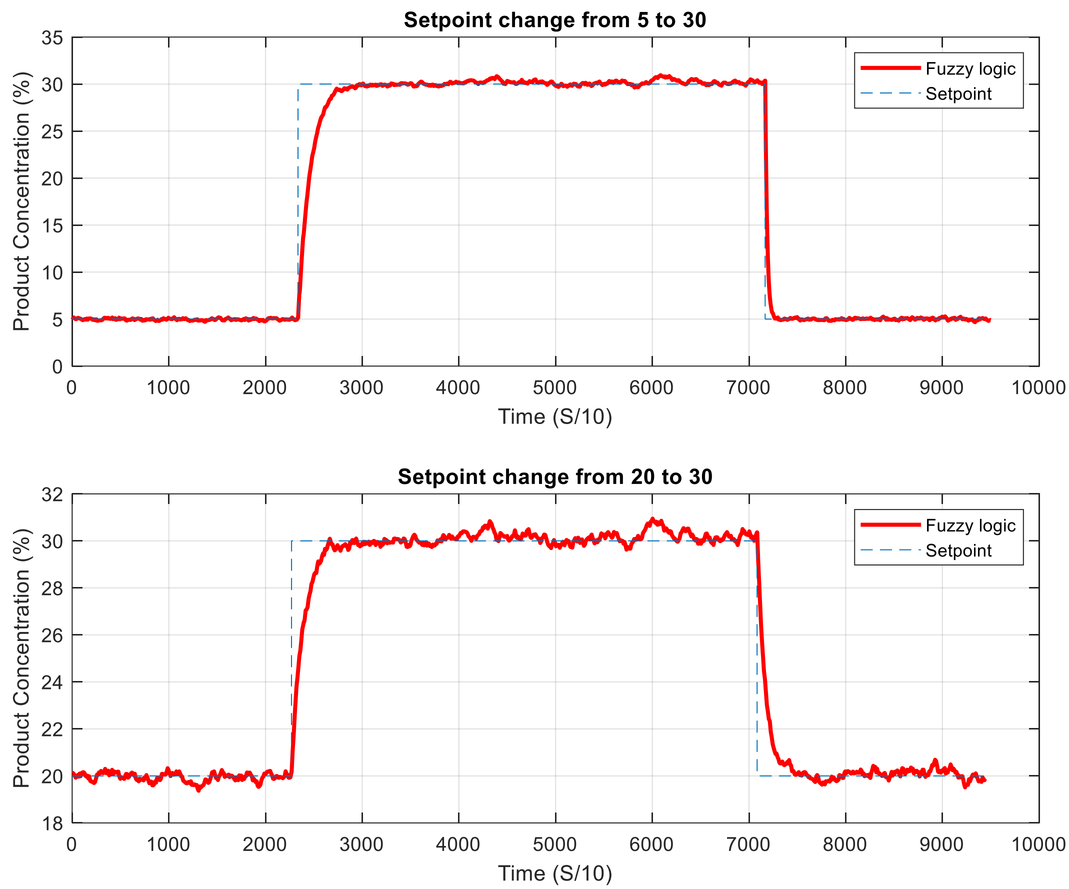


Figure 6. 8 Setpoint tracking response of the fuzzy logic controller

The above figure 6.8 indicated the setpoint tracking performances of the fuzzy logic controller. The two subfigures amplitudes are similar but seem different because of the different zoom ranges. Firstly, results show that the fuzzy logic strategy can achieve the tracking tasks for the simulation process relatively precisely. No obvious overshoot is observed and the settling time is within an acceptable range. Details will be discussed and compared in the next comparing section. Secondly, from the two subplots in figure 6.8, different setpoint change range has no evidently influence to the fuzzy logic controller's performances.

6.6 Results Comparison

The performances of PI and MPC have been discussed and compared in the previous chapter. In this section, the main tasks are obtaining the benefits and weaknesses of the fuzzy logic controller to compare with the other two strategies. 5% and 10% of disturbances were generated to the system to test the responses of different control strategies.

Figures 6.9 and 6.10 describe the simulation results. The two subfigures on the left side in each plot are the tracking performances of the controllers with different setpoint change. The other two subfigures in each plot on the right side are the magnified area of when the curves firstly reach the higher desired target.

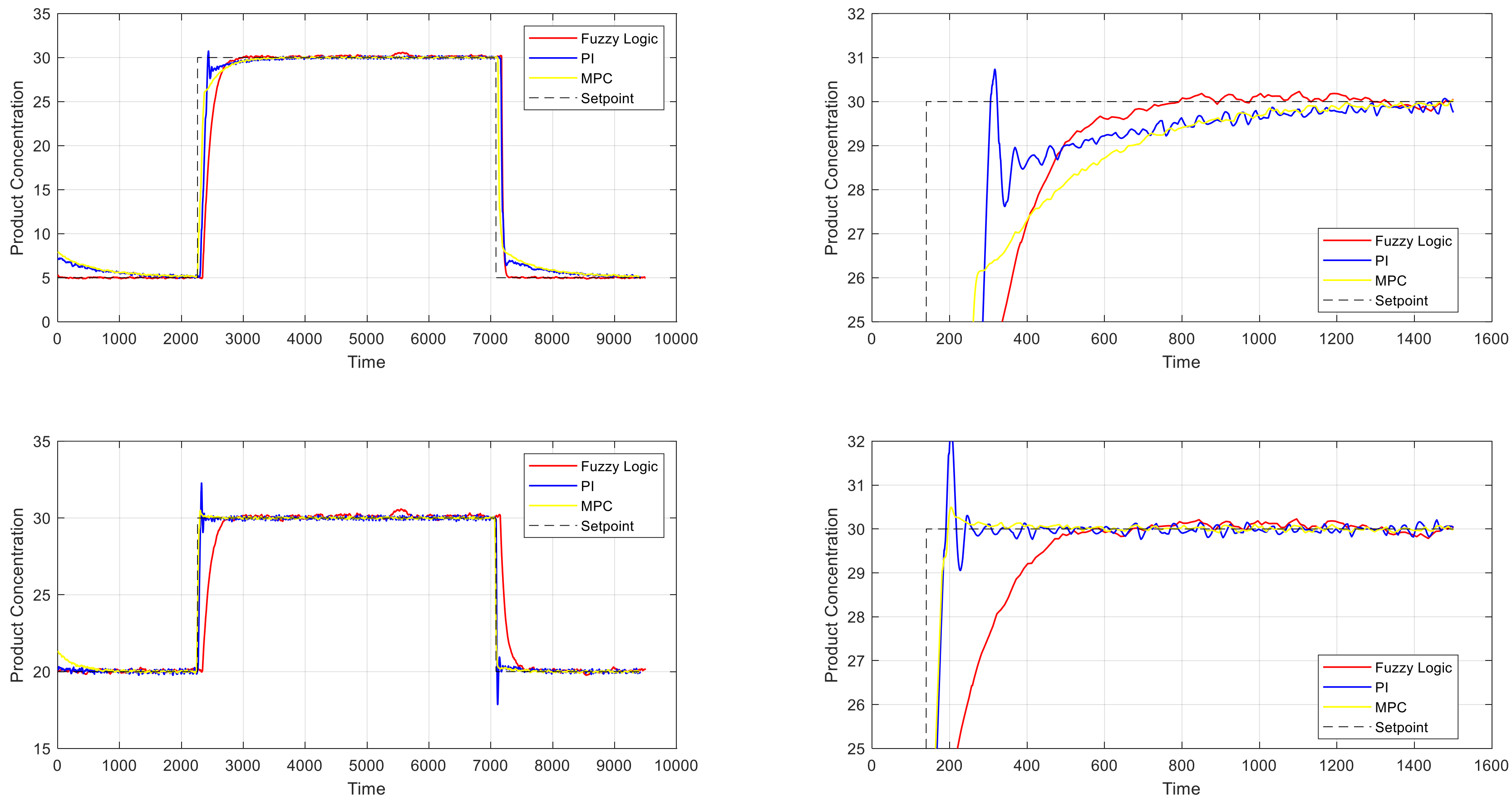


Figure 6.9 Setpoint tracking results comparison with 5% disturbances

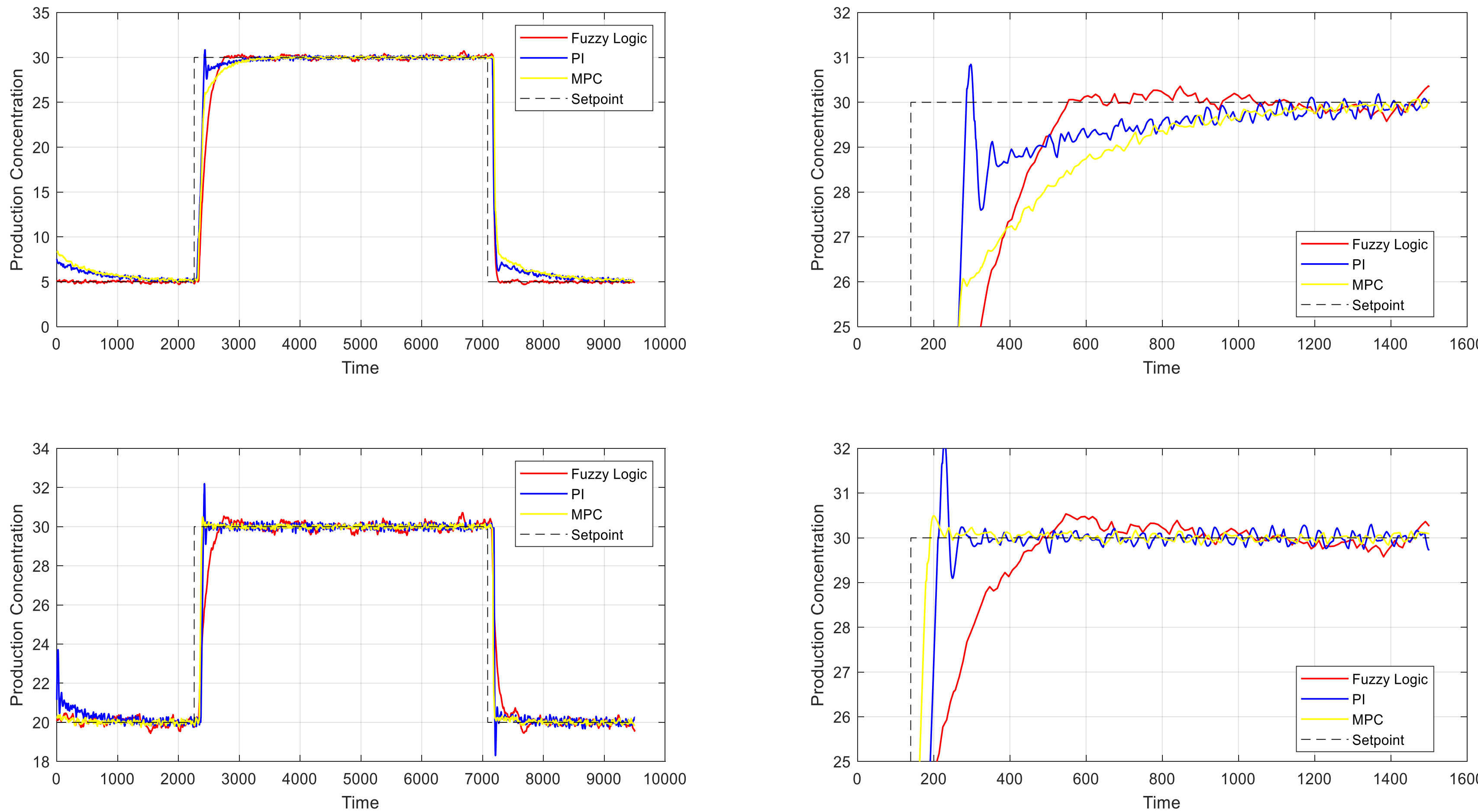


Figure 6.10 Setpoint tracking results comparison with 10% disturbances

It has been proved that disturbance rejection performance of fuzzy logic is relatively weak than the other two controllers. According to the two figures above, the results controlled by both PI and MPC have no evident variation of the amplitudes, which indicates these two control strategies are able to solve at least 10% system disturbance. Fuzzy logic has a similar performance on the tracking accuracy when the disturbance level is 5%. However, it starts to reveal the larger amplitude once the disturbance increases to 10% shown in the lower right subplot in figure 6.10.

Even fuzzy logic is not as good as PI and MPC on the accuracy control with more than 10% disturbance added, it has its benefits over the other two controllers. Firstly, the setpoint change range has almost no impact on fuzzy logic control, such as the settling time and results of accuracy. Secondly, no obvious overshoot can be obtained from the fuzzy logic controller, and it is settling down faster than PI and MPC for the big setpoint change range.

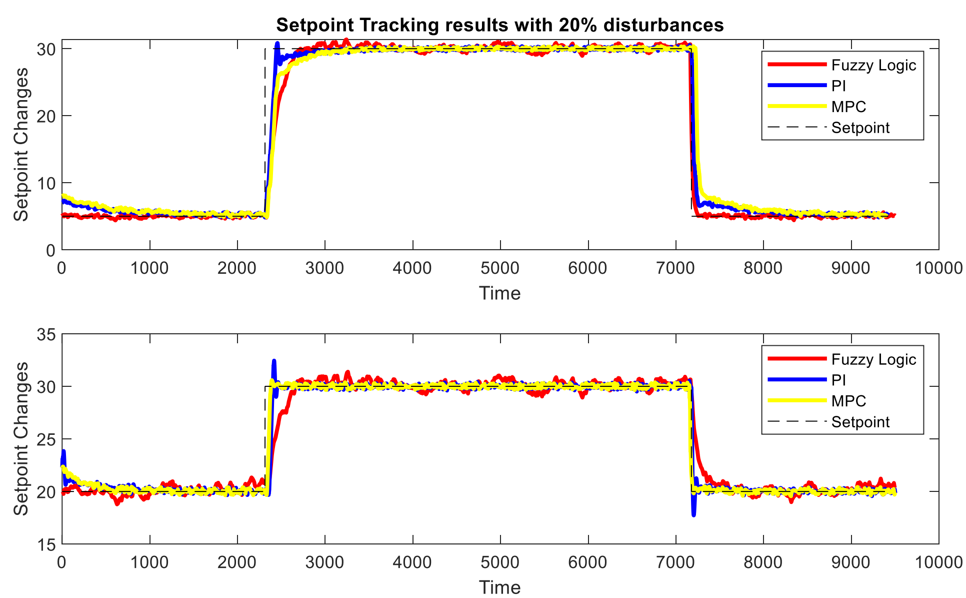


Figure 6. 11 Setpoint tracking results comparison with 20% disturbances

To increase the disturbance level to 20%, Figure 6.11 above illustrated the overall performance of three controllers. The fuzzy logic appears to have a bigger and bigger amplitude than PI and MPC for both setpoint change ranges. But the system is still controllable by using PI and MPC.

6.7 Summary

In this chapter, the application of fuzzy logic control strategy has been demonstrated on the three-effect falling film evaporator simulation.

Comparing with the PI and MPC controllers, they all can achieve the control target and maintain the simulation performances within an acceptable range. But the fuzzy logic controller is not recommended for the processes with large disturbances. In this chapter, the input with 5% and 10% disturbance level controlled by fuzzy logic have been compared and considered to be acceptable.

No obvious offset in the control response is observed in product concentration for all three control strategies, which means the simulation model replicates the evaporation process very well.

MPC response appears to be better than those of the PI and fuzzy logic control responses of setpoint change tracking in product concentration, especially for the small range of setpoint change. Meanwhile, MPC is the best strategy for the desired objective accuracy control, followed by PI and fuzzy logic. PI performs a little bit better than fuzzy logic control on the accuracy, but has similar results when the disturbance level is 5%.

Fuzzy logic control is not as good as the other two controllers on disturbance rejection performance. However, it has the smallest overshoot and settling time than MPC and PI.

In conclusion, after considering all the results, MPC is the most appropriate for this evaporation process among these three control strategies. Fuzzy logic has no overshoot and less settling time, but it is difficult to deal with large disturbance levels. PI is generally better than fuzzy logic on the accuracy, but not as good as MPC. However, when the setpoint change starts, PI is the fastest to respond among the three controllers.

Chapter 7 Conclusions and Future Work

This chapter summarises the objectives and results which have been achieved in this thesis and makes suggestions for the future development in areas closely linked to this work.

7.1 Overview of the completed objectives

In this thesis, several key issues concerning the industrial evaporation process which include system modelling and validating, conventional and advanced control strategies development, optimisation as well as the comparison of the performances for different controllers.

The ultimate objective of this thesis was the development of both conventional and advanced controllers to demonstrate an approach to the industrial evaporation process system control. The controllers developed in this thesis involve:

- A conventional PID controller with the gains tuned by using the Zeigler-Nichols method.
- An auto-tuning PID controller with an online auto-tuning approach.
- Model predictive controller with a constrained optimisation technique Quadratic Programming to optimise the manipulated variables, and a terminal weight objective function is applied to support the optimisation process.

- A fuzzy logic controller with the Mamdani inference system, which follows the designed IF-THEN rules to achieve the control objectives.

There are several other primary tasks have been completed in this thesis. For example:

- A comprehensive mathematical model of three-effects falling film evaporator was developed and employed for control purposes.
- Process variable and result values and trends were compared with similar previous research study to evaluate the model accuracy.
- Control results with different disturbance levels added to the input were obtained and compared in this thesis.

7.2 Conclusions

According to the simulation results controlled by different controllers discussed in chapters 4, 5 and 6, all the control strategies developed in this thesis could achieve the desired control objectives.

Model predictive control (MPC) strategy had the best overall performances than other controllers on system accuracy, overshoot, and settling time-based on the simulation results in this thesis. Fuzzy logic and PID based controller offered similar results with a small percentage of disturbance added to the input because fuzzy logic is weak to deal with the large disturbance. Once the disturbance level increased to more than 10%, it had no obvious impact on MPC and PI but affected the performances, especially the output accuracy, of the fuzzy logic controller. The performances of Auto-tuning and Conventional PID controllers were acceptable but relatively weak for this complex evaporation process comparing to the MPC; however, each control strategy has its advantage on specific aspects.

As mentioned above, the desired product concentration exiting each effect was 30%, 38% and 52% respectively in this thesis. MPC had a great performance on accuracy with the desired constant setpoint. The error range could be maintained from 0.02% to 0.1% with different disturbances from 5% to 20%. However, with the same level of disturbance, the error range controlled by PID based controllers and fuzzy logic controller were very similar from approximately 0.2% to 2%, and 0.2% to 3.5%.

It was also proved that the PID gains calculated by an online auto-tuning method provided by MATLAB were better than those calculated through the Zeigler-Nichols method. Although auto-tuning PID performed a little bit better than conventional PID, it took time to collect the system input/output data to evaluate the response frequency (online auto-tuning). Meanwhile, MPC was also a good choice for disturbance rejection than the other three control methods.

About the setpoint tracking performances, both MPC and PID based controllers had a quicker response than fuzzy logic controller. With the increasing concentration setpoint range, the response time of the PID controller was the fastest, which is within about 20 seconds, followed by MPC and fuzzy logic controller. However, fuzzy logic controller was the best for overshooting control among these three, followed by MPC. PID based controller had poor performance on overshooting results than the other two control methods.

TSA was used to validate the system model and estimate the model quality successfully. The three-effect falling film evaporator mathematical model was developed precisely enough to carry out various control methods to ensure the process accuracy and consistency.

7.3 Future work

Though substantial progress in the control, optimisation, and comparison of the multi-effect evaporation process has been made in recent years, there are still a large number of contributions and improvements can be made for future development. Several recommendations are given here for future work, some of which reflect the limitations of this research study.

7.3.1 Model improvement

The simulation model developed in this thesis is extended and improved based on Newell and Lee's single-effect evaporator. Within the scope of this research, obtaining real industrial process data to validate the simulation model was not possible. The developed mathematical model is considered to be good enough for control purposes, but it could be improved more precisely if there was data from a real industrial plant. If so, the simulation results would be more accurate and convincing. Meanwhile, the 'black box' model can be developed according to these industrial data and be used in more operations for the evaporation process, such as output prediction.

7.3.2 Controllers Development

In this thesis, only two advanced control strategies (MPC and FLC) were developed and applied to the simulation model. There are more alternative advanced control technologies, like neural network control, robust control, and adaptive control suitable for this simulation model. Hybrid control systems, such as fuzzy PID, neural network MPC, adaptive PID and so on, are also potential methods to improve the control system for future work.

7.3.3 Preheater and Spray Dryer Development

Pre-heating is a heat treatment process of the product until a certain temperature before it fed into the first evaporation effect. Spray dryers have been introduced in chapter 2 and widely implemented in the food industry to produce powder products. Adding these two parts into the evaporation process would create a complete powder process plant, which is not for the milk product only, but also for other solution powder products.

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Appendix A:

SIMULINK block diagrams

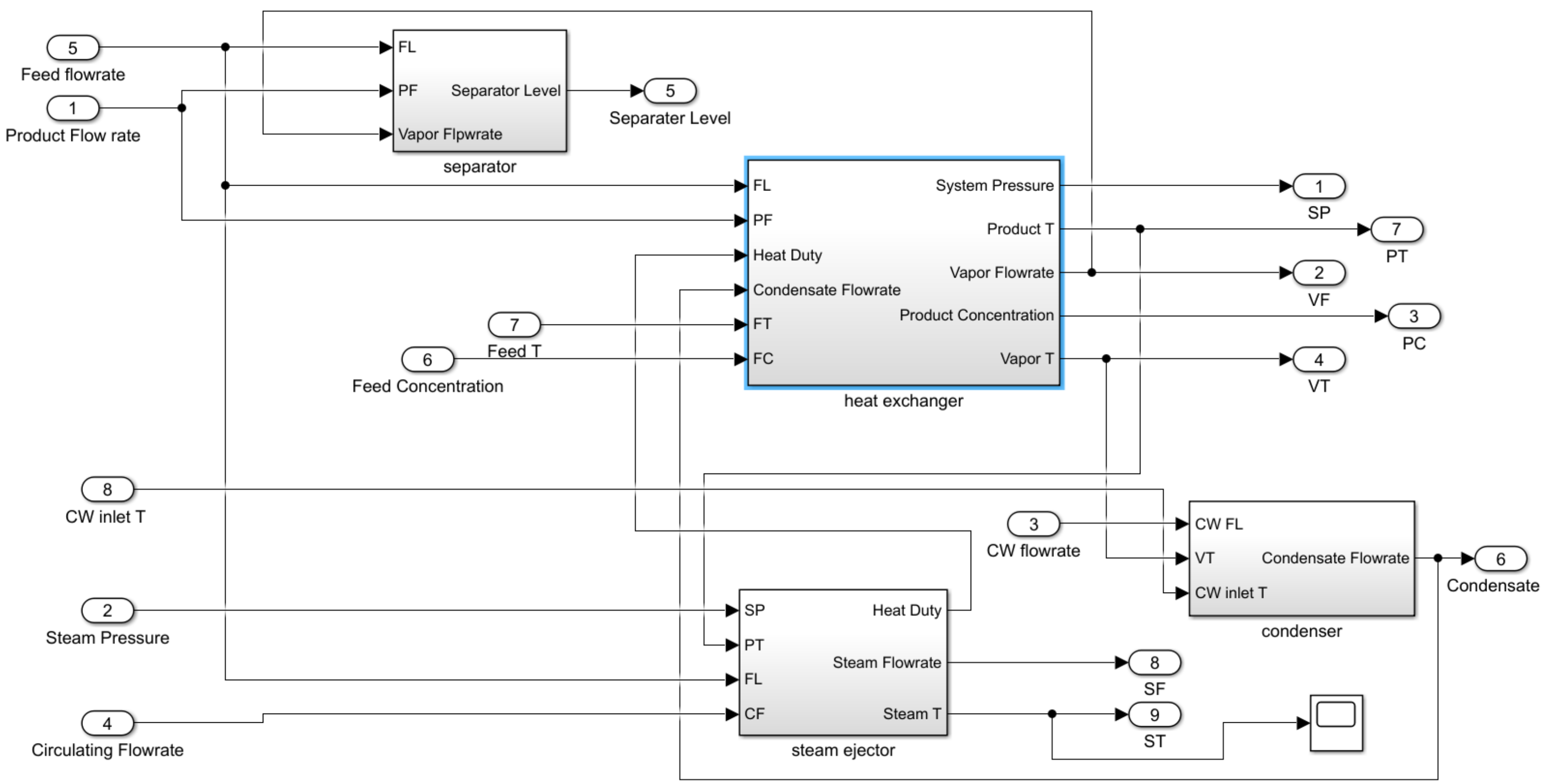

Figure A-1 Falling film evaporator effect sub-system block

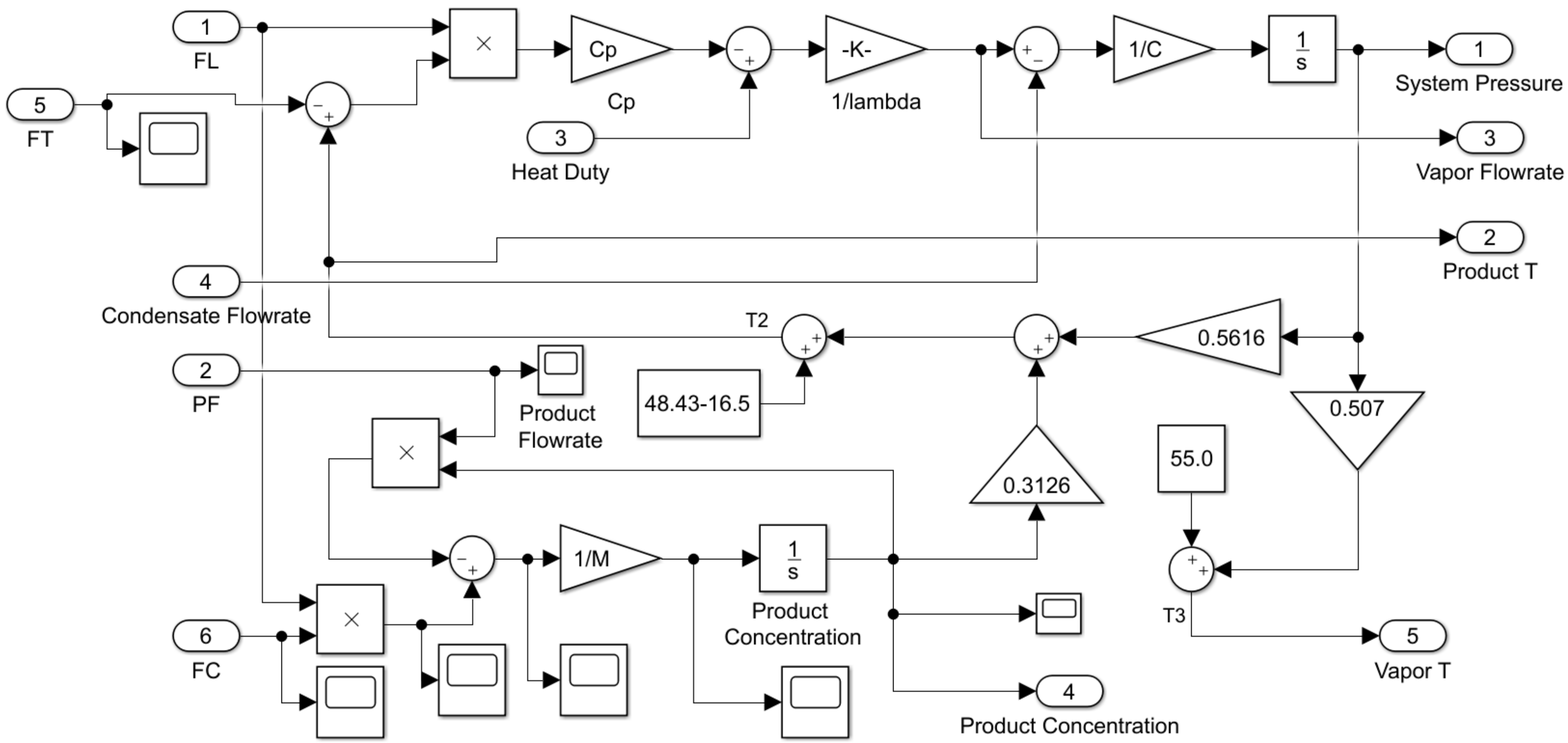

Figure A-2 Heat exchanger sub-system block

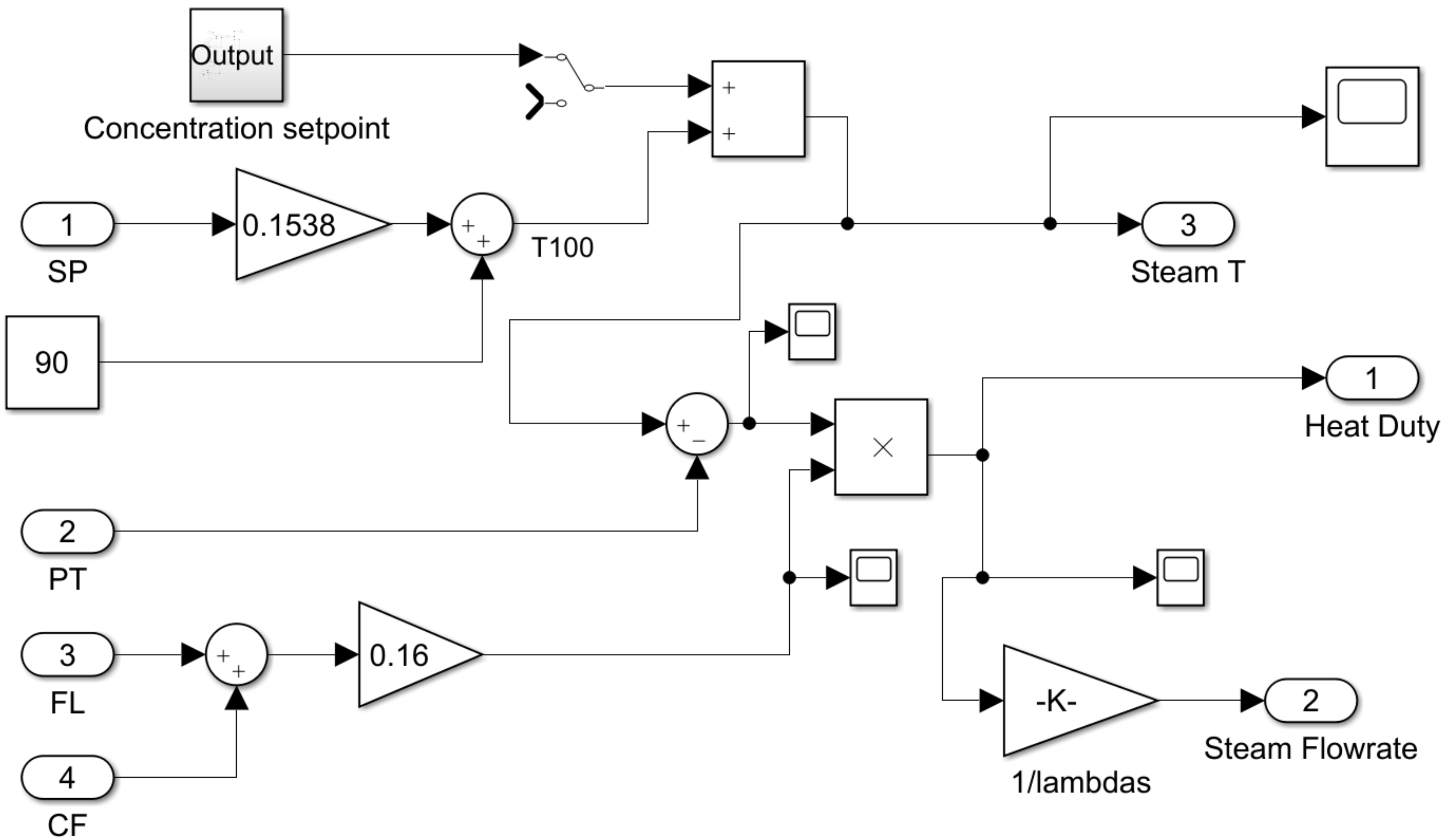

Figure A-3 Steam ejector sub-system block

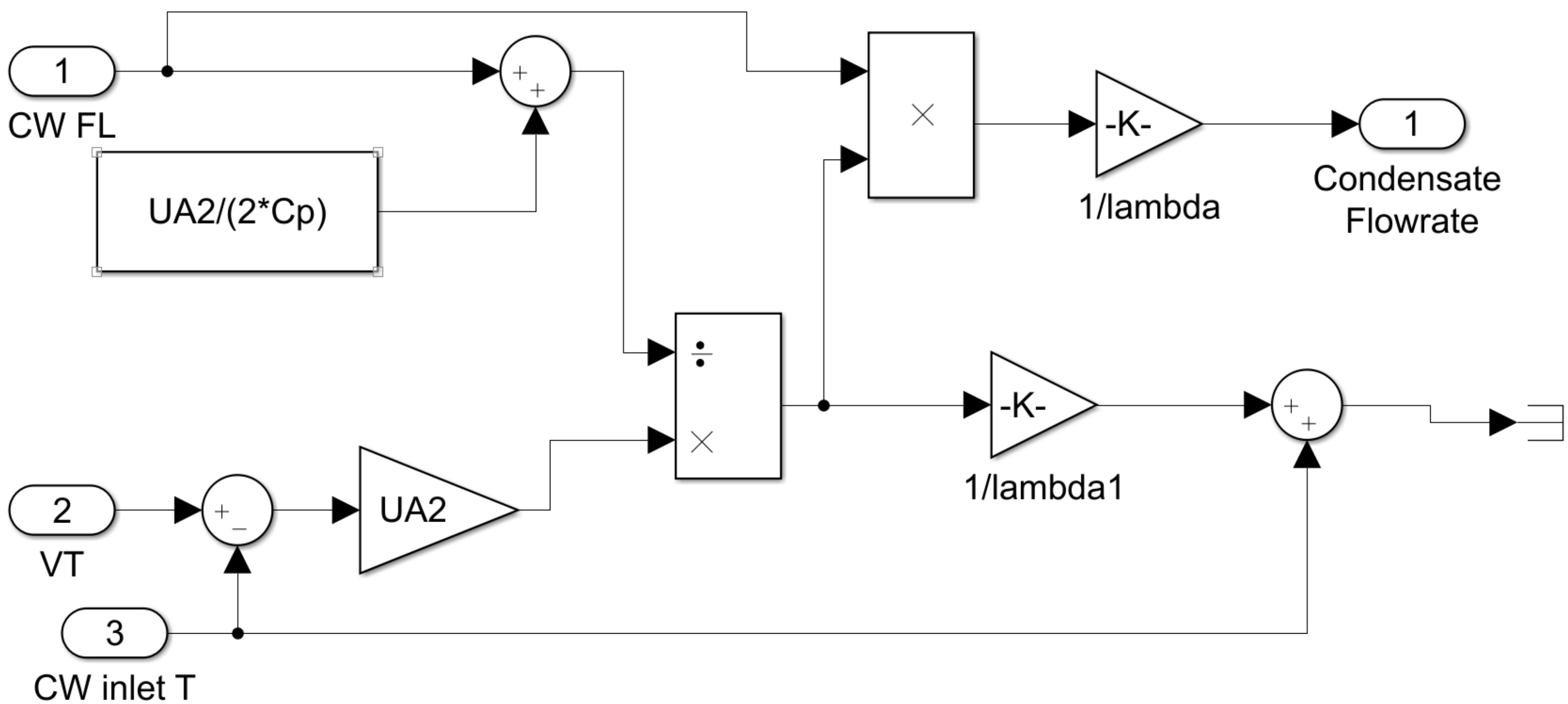

Figure A-4 Condenser sub-system block

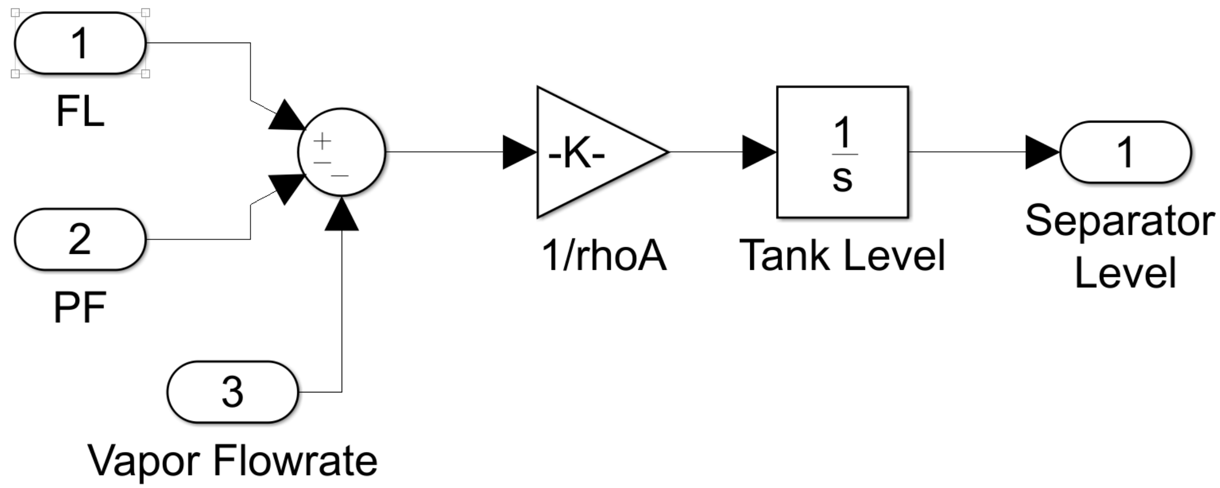

Figure A-5 Separator sub-system block

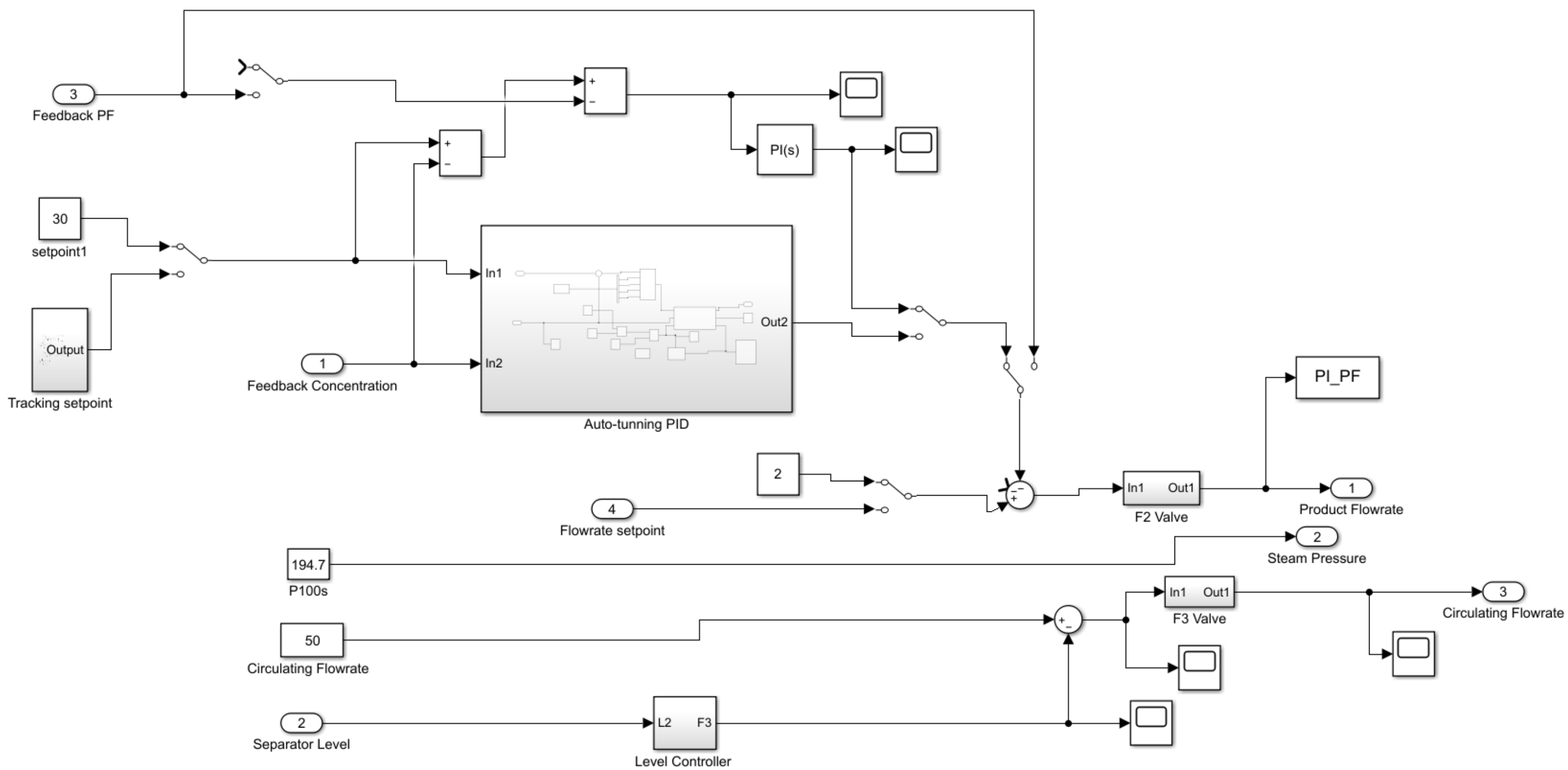

Figure A-5 PID and Auto-tuning PID controller sub-system block

Appendix B:

MATLAB Script Commands

```
%% Disturbances of input
% clear all, close all,
F=10;        % Product feed flowrate
X=5;         % Product feed concentration
TI=40;       % Product temperature
TW=25;       % Cooling water temperature
p=0.05;      % Disturbance percentage
T=2400;      % Running time
a=10+(F*p)/2;b=10-(F*p)/2;
f1 = (b-a).*rand(T+1,1) + a-10;
time=0:1:T;
Time=time';
F1=[Time,f1]; % Product feed flowrate disturbance value

a1=5+(X*p/2);b1=5-X*p/2;
x1 = (b1-a1).*rand(T+1,1) + a1-5;
X1=[Time,x1]; % Product feed concentration disturbance value

a2=40+TI*p/2 ;b2=40-TI*p/2;
t1 = (b2-a2).*rand(T+1,1) + a2-40;
T1=[Time,t1]; % Product temperature disturbance value

a3=25+TW*p/2;b3=25-TW*p/2;
t200 = (b3-a3).*rand(T+1,1) + a3-25;
```

```
T200=[Time,t200]; % Cooling water temperature disturbance value



%% PID AND AUTO-TUNING PID RESULTS COMPARISON OF THREE EFFECTS
figure(1)
x=PI.data(2000:end);
y=auto.data(2000:end);
plot(x,'r');hold on;
plot(y,'b');hold off;
xlabel('time (s)')
ylabel('concentration %')
title('Effect 1 milk concentration')
legend({'conventional PID','auto PID'},'Location','northeast')

figure(2)
x=PI1.data(2000:end);
y=auto1.data(2000:end);
plot(x, 'r');
hold on;
plot(y, 'b');
xlabel('time (s)')
ylabel('concentration %')
title('Effect 2 milk concentration')
legend({'conventional PID','auto PID'},'Location','northeast')


figure(3)
x=PI2.data(2000:end);
```

```
y=auto2.data(2000:end);
plot(x, 'r');
hold on
plot(y, 'b');
xlabel('time (s)')
ylabel('concentration %')
title('Effect 3 milk concentration')
legend({'conventional PID','auto PID'},'Location','northeast')



%% MPC WITH DIFFERENT DISTURBANCES RESULTS OF THREE EFFECTS
figure(1)
plot(MPC5_e1(2000:3000),'r','LineWidth',2), hold on;
plot(MPC10_e1(2000:3000),'b','LineWidth',2),hold on;
plot(MPC20_e1(2000:3000),'g','LineWidth',2),hold on;
plot(MPC50_e1(2000:3000), 'y','LineWidth',2), hold on;
plot(set_point1(2000:3000),'k','LineWidth',2),
hold off

figure(2)
plot(MPC5_e2(2000:3000),'r','LineWidth',2), hold on;
plot(MPC10_e2(2000:3000),'b','LineWidth',2),hold on;
plot(MPC20_e2(2000:3000),'g','LineWidth',2),hold on;
plot(MPC50_e2(2000:3000), 'y','LineWidth',2), hold on;
plot(x,y,'k','LineWidth',2);
hold off
```

```
figure(3)
plot(MPC5_e3(2000:3000),'r','LineWidth',2), hold on;
plot(MPC10_e3(2000:3000),'b','LineWidth',2),hold on;
plot(MPC20_e3(2000:3000),'g','LineWidth',2),hold on;
plot(MPC50_e3(2000:3000), 'y','LineWidth',2), hold on;
plot(x,y1,'k','LineWidth',2); hold off;



%% COMPARISON OF THREE CONTROLLERS (PID, MPC AND FUZZY LOGIC) WITH DIFFERENT
DISTURBANCE LEVELS OF EACH EFFECT

N1=2000; N2=4000; % Samples range
figure(1) % 5% disturbances, 1st effect,
subplot(2,2,1)
plot(MPC5_e1(N1:N2),'r','LineWidth',1), hold on;
% 5% disturbance, 1st effect,MPC controller
plot(PI5_e1(N1:N2),'b','LineWidth',1), hold on;
% 5% disturbance, 1st effect, PID controller
plot(FL5_e1(N1:N2),'K','LineWidth',1), hold on;
% 5% disturbance, 1st effect, Fuzzy Logic controller

subplot(2,2,2) % 10% disturbance, 1st effect
plot(MPC10_e1(N1:N2),'r','LineWidth',1), hold on;
plot(PI10_e1(N1:N2),'b','LineWidth',1), hold on;
plot(FL10_e1(N1:N2),'K','LineWidth',1), hold on;

subplot(2,2,3) % 20% disturbance,  1st effect
plot(MPC20_e1(N1:N2),'r','LineWidth',1), hold on;
```

```
plot(PI20_e1(N1:N2),'b','LineWidth',1), hold on;
plot(FL20_e1(N1:N2),'K','LineWidth',1), hold on;

subplot(2,2,4) % 50% disturbance,  1st effect
plot(MPC50_e1(N1:N2),'r','LineWidth',1), hold on;
plot(PI50_e1(N1:N2),'b','LineWidth',1), hold on;
plot(FL50_e1(N1:N2),'K','LineWidth',1), hold off;


figure(2)
subplot(2,2,1) % 5% disturbances, 2nd effect
plot(MPC5_e2(N1:N2),'r','LineWidth',1), hold on;
plot(PI5_e2(N1:N2),'b','LineWidth',1), hold on;
plot(FL5_e2(N1:N2),'K','LineWidth',1), hold on;

subplot(2,2,2) % 10% disturbances, 2nd effect
plot(MPC10_e2(N1:N2),'r','LineWidth',1), hold on;
plot(PI10_e2(N1:N2),'b','LineWidth',1), hold on;
plot(FL10_e2(N1:N2),'K','LineWidth',1), hold on;

subplot(2,2,3) % 20% disturbances, 2nd effect
plot(MPC20_e2(N1:N2),'r','LineWidth',1), hold on;
plot(PI20_e2(N1:N2),'b','LineWidth',1), hold on;
plot(FL20_e2(N1:N2),'K','LineWidth',1), hold on;

subplot(2,2,4) % 50% disturbances, 2nd effect
plot(MPC50_e2(N1:N2),'r','LineWidth',1), hold on;
plot(PI50_e2(N1:N2),'b','LineWidth',1), hold on;
plot(FL50_e2(N1:N2),'K','LineWidth',1), hold off;
```

```
figure(3)
subplot(2,2,1) % 5% disturbances, 3rd effect
plot(MPC5_e3(N1:N2),'r','LineWidth',1), hold on;
plot(PI5_e3(N1:N2),'b','LineWidth',1), hold on;
plot(FL5_e3(N1:N2),'K','LineWidth',1), hold on;

subplot(2,2,2) % 10% disturbances, 3rd effect
plot(MPC10_e3(N1:N2),'r','LineWidth',1), hold on;
plot(PI10_e3(N1:N2),'b','LineWidth',1), hold on;
plot(FL10_e3(N1:N2),'K','LineWidth',1), hold on;

subplot(2,2,3) % 20% disturbances, 3rd effect
plot((MPC20_e3(N1:N2)),'r','LineWidth',1), hold on;
plot(PI20_e3(N1:N2),'b','LineWidth',1), hold on;
plot(FL20_e3(N1:N2),'K','LineWidth',1), hold on;

subplot(2,2,4) % 50% disturbances, 3rd effect
plot(MPC50_e3(N1:N2),'r','LineWidth',1), hold on;
plot(PI50_e3(N1:N2),'b','LineWidth',1), hold on;
plot(FL50_e3(N1:N2),'K','LineWidth',1), hold off;



%% FUZZY LOGIC CONTROL RESULTS with DIFFERENT DISTURBANCE LEVELS of each effect
S1=2000;S2=3000; %Sample range
figure(1) % 1st effect results in different disturbance levels
plot(FL5_e1_sp30(S1:S2),'r','LineWidth',2), hold on;
plot(FL10_e1_sp30(S1:S2),'b','LineWidth',2),hold on;
plot(FL20_e1_sp30(S1:S2),'g','LineWidth',2),hold on;
```

```
plot(FL50_e1_sp30(S1:S2), 'y','LineWidth',2), hold on;
hold off

figure(2) % 2nd effect results in different disturbance levels
plot(FL5_e2_sp30(S1:S2),'r','LineWidth',2), hold on;
plot(FL10_e2_sp30(S1:S2),'b','LineWidth',2),hold on;
plot(FL20_e2_sp30(S1:S2),'g','LineWidth',2),hold on;
plot(FL50_e2_sp30(S1:S2), 'y','LineWidth',2), hold on;
hold off

figure(3) % 3rd effect results in different disturbance levels
plot(FL5_e3_sp30(S1:S2),'r','LineWidth',2), hold on;
plot(FL10_e3_sp30(S1:S2),'b','LineWidth',2),hold on;
plot(FL20_e3_sp30(S1:S2),'g','LineWidth',2),hold on;
plot(FL50_e3_sp30(S1:S2), 'y','LineWidth',2), hold on;
hold off
```