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Model Predictive Control of a Double Effect Evaporator Via Simulation

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ABSTRACT

This paper is a study of the dynamic behavior of the double effect evaporator on the basis of energy and material balance under unsteady state conditions inside the evaporator. The simulation process was based on a model for the intensification of tomato juice. The mathematical model was used to study the effect of operational conditions, namely, the temperature of the feed, the flow rate of the feed, and the feed concentration. The dynamic behavior of the open system was studied by measuring the temperature response of the evaporators to the change of the staging function in the temperature of the feed, the feed flow rate and the feed concentration in the rate of (±10%, ±20%).The proportional-integral-derivative and model predictive controllers were applied to solve the difficult problem by determining the best operational conditions and avoid a sharp increase in temperature. Two methods are tested to control a wide range of operating conditions and simulation results show that there is good accuracy. The MPC controller is more accurate than the PID control and faster to reach the constant value.

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محاكاة السيطرة التنبئي الانموذجي لمبخر ثنائي التأثير

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الخلاصة

تضمن هذا العمل دراسة السلوك الديناميكي للمبخر ثنائي التأثيرات على أساس توازن الطاقة والمواد في ظل ظروف غير مستقرة داخل المبخر. استندت عملية المحاكاة إلى نموذج لتكثيف عصير الطماطم. تم استخدام النموذج الرياضي أيضًا لدراسة تأثير الظروف التشغيلية ، وهي درجة حرارة التغذية ومعدل تدفق التغذية وتركيز التغذية. تمت دراسة السلوك الديناميكي للنظام المفتوح عن طريق قياس درجة حرارة استجابة المبخرات لتغيير وظيفة التدريج في درجة حرارة التغذية ومعدل تدفق التغذية وتركيز التغذية بمعدل (± 10 ٪ ، ± 20 ٪). تم تصميم وتنفيذ نوعين من المسيطرات وهما التناسبي-التكاملي-التفاضلي والتنبئي الانموذجي لحل لمشكلة السيطرة الصعبة من خلال تحديد أفضل الظروف التشغيلية وتجنب زيادة حادة في درجة الحرارة تم فحص الطريقتين لمدى واسع من الظروف التشغيلية وبين برنامج المحاكاة دقة عالية في النتائج. كما ان مسيطر والتنبئي الانموذجي اكثر دقة وسريع الى الوصول الى الحالة المستقرة من المسيطر التناسبي-التكاملي-التفاضلي.

1. INTRODUCTION

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Evaporation is a process by which a substance in the liquid tends to convert into the gaseous phase without reaching its boiling point. The evaporated solvent, which will be eliminated from the liquid, mostly water, will then produce to a more concentrated product. Most of the manufacturers of this component use a multi-effect evaporator system to produce the suitable concentration that is demanded by the market. Evaporation process only occurs below the boiling point and on the surface of the liquid. Molecules on the surface of the liquid will absorb heat from the atmosphere and break the intermolecular bond and change its phase to gas. The fact that evaporation process also undergoes a vaporized process makes most of the people assume that evaporation and distillation is a similar process. By contrast, distillation is a modern separation technique and the separation happened at the specific boiling point of liquids. However, not in evaporation process where it could start although the liquid is not reaching the boiling point [1]. Evaporator may have a single unit, called single-effect, or multiple units, called multiple-effect. The most important advantage of multiple-effect over the single-effect evaporator is the economy. Multiple-effect scheme evaporates more water per kilogram of steam fed to the unit by reusing the vapor from one effect as the heating medium for the next [2].

Many researchers studied the model of multi-effect evaporators such as Stefanov and Hoo [3] who represented a distributed model of a multiple-effect falling-film evaporator plant and developed it to four-effect falling film evaporators to discover them, the results of this model showed the important phenomena of evaporation, the pressure dynamics of the plant as well as heating and condensation for different hydrodynamic regimes. Real factories steady state was compared with simulation results to validate the model, Kayaa and sarac [4] developed a mathematical model for multiple-effect evaporators. The results show that in the multiple-effect evaporations, the mathematical models for co-current, counter current and parallel flow operation types under with/without pre-heating cases were developed and the best operation for economic steam consumption is countercurrent operation with pre-heating while the worst case is parallel flow operation without pre-heating , Mohanty et al. [5] simplified a model for evaluating multiple impact evaporator systems based on the essential elements of process integration, representing the boiling point increase and variability in physical-thermal properties. The results showed new concepts of current analysis, temperature trajectory and internal heat exchanger for model equation formulation.

The simulation study became easier with the development of a computer code which is capable of simulating the steady state condition of a multiple effect evaporator. Khademi et al. [6] presented the steady-state simulation and optimization of a six-effect evaporator and the provision of its relevant software package. In this investigation, the modeling equations of each of the existing building blocks are written in steady-state conditions. The simulation results good agreement with design data. The results of optimization show that feed mass flow rate 51,408 kg/h and condenser pressure 7.6 kPa are optimized operating conditions for this system; also optimized operating time for operation of vaporizing unit in this refinery is the period of 187 days and the unsteady-state simulation is recommended for future work. With unsteady-state simulation, the economic influence of the optimized time of operation can be analyzed.

The simulation was carried out to study the closed-loop control performance using computer. Smith [7] concluded that multiple effect evaporator control is a problem that has been widely reported in the pulp and sugar industries. Such factors make effective evaporator control crucial to overall factory efficiency. The complexity and large number of interactions make single loop proportional- integral -derivative (PID) controller difficult and often sub-optimal. A Model Predictive Control (MPC) algorithm is presented as a different approach to solving the multiple input, multiple output problem. The results show that the subject of evaporator control has been investigated and some recent developments have been presented. Pan and Ning [8] studied the dynamic mathematical model of the evaporation system in sugar mill by mechanism analysis. PID and nonlinear adaptive predictive control algorithm are applied to the system. The error between the model output and the actual output is small which satisfied the control requirements. It shows that the model has a good predictive ability. Simulation experiments on the model show that the predictive control algorithm has better robustness and stability than the PID control algorithm. The purposes of this study are simulation of the double-effect evaporator on the basis of energy and material balance under unsteady state conditions and then applying the PID and model predictive controllers.

2. SIMULATION PROGRAM

2.1. Apparatus and Procedures

The double-effect evaporator with backward feeding arrangement used for tomato juice concentrate is shown in the Fig. 1. The two effects are numbered from left to right as Tank 1 and Tank 2, respectively. The raw juice having flow rate F, concentration X_f and temperature T_f enters Tank 2, and the steam with flow rate S and temperature T_s enters Tank 1. The mass holdups in the two tanks are defined as M_1 and M_2. V_1 and V_2 are the vapor flow rates from the overhead of two tanks with temperature T_1 and T_2, respectively. P_1 and P_2 are the product flow rates from the two effects with product concentration X_p and X_2, and temperature T_1 and T_2, respectively. For transfer of feed it required pump since the flow is from low pressure to higher pressure .The concentrated liquid is obtained in first effect. The energy flow in steam supplied to evaporation is transformed into multiple energy flows in vapors and condensates leaving evaporation [9]. Double-effect scheme evaporates more water per kilogram of steam fed to the unit by reusing the vapor from one effect as the heating medium for the next.

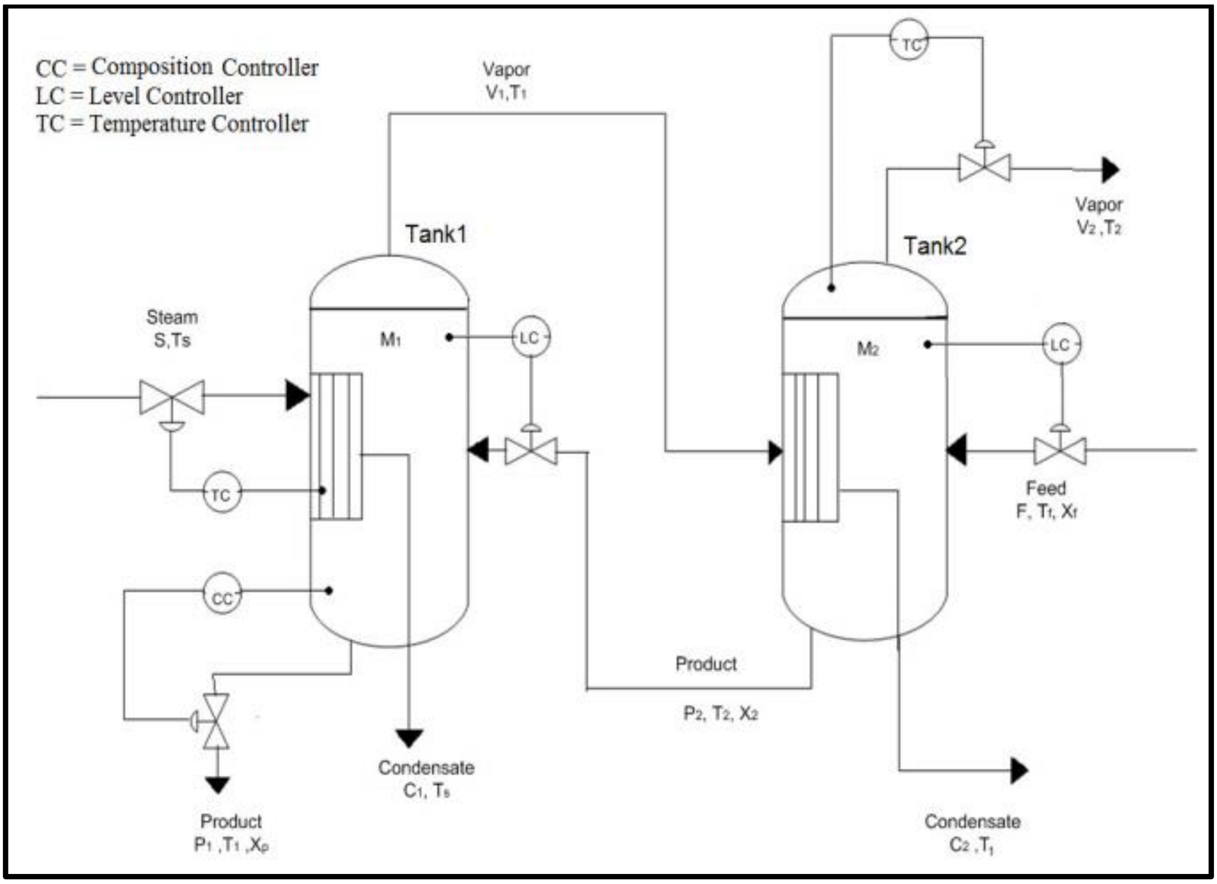

Fig. 1. Double-effect evaporator of tomato paste.

2.2. Process model

The modeling of a double-effect evaporator includes the formulation of total mass and component balances together with an energy balance. The dynamic model of the evaporator is derived for tomato concentrate based on the references No. [9-11]. The evaporation process involves mass and heat balance. The tomato juice is assumed as a binary solution of water and soluble solids, both considered inert in a chemical sense. The microscopically evaporator model consisted of a set of differential-algebraic equations that have been constructed based on conservative laws and empirical relationships. It should be noted that only the juice phase is considered for modeling. The assumptions involved in the formulation of model are listed below:

1. The heat losses to the surrounding are neglected.
2. Composition and temperature inside each evaporator are Homogeneous.
3. Overhead vapors are pure steam.
4. Liquid holdup is variable.
5. Vapor holdup is negligible
6. Latent heat of vaporization or condensation varied with temperature.
7. Boiling point of the solution is not elevated.
8. No reactions occurred.
9. The pressures inside the effects are constant [10],[11].

The total mass balance around the first and second effect is as follow:

First effect: $\frac{dM_1}{dt} = P_2 - P_1 - V_1$ (1)

Second effect: $\frac{dM_2}{dt} = F - P_2 - V_2$ (2)

Component (solid) mass balance around the first effect is the following:

$$\frac{D(M_1X_P)}{dt} = P_2X_2 - P_1X_P \quad (3)$$

$$M_1\frac{d(X_P)}{dt} + X_P\frac{d(M_1)}{dt} = P_2X_2 - P_1X_P \quad (4)$$

$$M_1\frac{d(X_P)}{dt} = P_2X_2 - P_1X_P - X_P\frac{d(M_1)}{dt} \quad (5)$$

Substituting equation (1) in equation (5) and becomes:

$$M_1\frac{d(X_P)}{dt} = P_2X_2 - P_1X_P - X_P\,(P_1 - P_2 - V_1) \quad (6)$$

Equation (6) can be written as:

$$M_1\frac{d(X_P)}{dt} = P_2X_2 - P_1X_P - X_PP_2 + X_PP_1 + X_PV_1 \quad (7)$$

Rearranging equation (7):

$$\frac{d(X_P)}{dt} = \frac{P_2\,(X_2 - X_P) + X_PV_1}{M_1} \quad (8)$$

Component (solid) mass balance around the second effect is the following:

$$\frac{D(M_2X_2)}{dt} = FX_F - P_2X_2 \quad (9)$$

$$M_2\frac{d(X_2)}{dt} + X_2\frac{d(M_2)}{dt} = FX_F - P_2X_2 \quad (10)$$

$$M_2\frac{d(X_2)}{dt} = FX_F - P_2X_2 - X_2\frac{d(M_2)}{dt} \quad (11)$$

Substituting equation (2) in equation (11) and becomes:

$$M_2\frac{d(X_2)}{dt} = FX_F - P_2X_2 - X_2(F - P_2 - V_2) \quad (12)$$

Equation (12) can be written as:

$$M_2\frac{d(X_2)}{dt} = FX_F - P_2X_2 - X_2F + X_2P_2 + X_2V_2 \quad (13)$$

Rearranging equation (13):

$$\frac{D(X_2)}{dt} = \frac{F\,(X_F - X_2) + X_2V_2}{M_2} \quad (14)$$

The steam flow rate to the first effect is obtained through energy balance on the first effect heat exchanger as:

$$S\ \lambda(T_S) = U_1A_1\,(T_S - T_1) \quad (15)$$

Rearranging equation (15):

$$S = U_1A_1\,(T_S - T_1) / \lambda(T_S) \quad (16)$$

Similarly, the vapor flow rate to the second effect is derived from the energy balance on the second effect heat exchanger as:

$V_1\lambda(T_1) = U_2\ A_2\,(T_1 - T_2)$ Rearranging equation (17):

$V_1 = U_2A_2\,(T_1 - T_2) / \lambda(T_1)$ (18)

In the following, the energy balance equations are derived.

The energy balance on the first effect is:

$$\frac{d[M_1 h(T_1,X_P)]}{dt} = P_2h(T_2,X_2)+S\lambda(T_S)-P_1h(T_1,X_P)-V_1H(T_1) \quad (19)$$

$$M_1 \frac{dh(T_1,X_P)}{dt} + h(T_1,X_P)\frac{dM_1}{dt} = P_2h(T_2,X_2) + S\,\lambda(T_S) - P_1h(T_1X_P) - V_1H(T_1) \quad (20)$$

$$M_1 \frac{dh(T_1,X_P)}{dt} = P_2h(T_2,X_2) + S\,\lambda(T_S) - P_1h(T_1,X_P) - V_1H(T_1) - h(T_1,X_P)\frac{dM_1}{dt} \quad (21)$$

By substituting equation (1)

$$M_1 \frac{dh(T_1,X_P)}{dt} = P_2[h(T_2,X_2) - h(T_1,X_P)\frac{dM_1}{dt}] + S\,\lambda(T_S) - V_1[H(T_1) - h(T_1,X_P) \quad (22)$$

Using equation (15)

$$M_1 \frac{dh(T_1,X_P)}{dt} = P_2[h(T_2,X_2) - h(T_1,X_P)] + U_1A_1(T_S - T_1) - V_1[H(T_1) - h(T_1,X_P) \quad (23)$$

This gives:

$$\frac{dh(T_1,X_P)}{dt} = \frac{P_2[h(T_2,X_2) - h(T_1,X_P)] + U_1A_1(T_S - T_1) - V_1[H(T_1) - h(T_1,X_P)}{M_1} \quad (24)$$

The energy balance on the second effect yields:

$$\frac{dM_2h(T_2,X_2)}{dt} = F[h(T_f,X_f) + V_1\,\lambda_2(T_1) - P_2h(T_2,X_2) - V_2H(T_2) \quad (25)$$

$$\frac{dh(T_2,X_2)}{dt} = \frac{F[h(T_f,X_f) - h(T_2,X_2) + U_2A_2(T_1 - T_2)] - V_2[H(T_2) - h(T_2,X_2)]}{M_2} \quad (26)$$

The enthalpy of the product (tomato juice) is given by the correlation

$$h(T,X) = (4.177 - 2.506X)\,T \quad (27)$$

The pure solvent vapor (steam) enthalpy is obtained using a polynomial regression equation of values from the steam tables as:

$$H(T) = 2495.0 + 1.958\,T - 0.002128\,T^2 \quad (28)$$

For the condensate streams, the pure solvent liquid enthalpy is also found from the steam tables as:

$$h(T) = 4.177\,T \quad (29)$$

The latent heat of vaporization can be computed as:

$$\lambda(T) = H(T) - h(T) = 2495.0 - 2.219\,T - 0.002128T^2 \quad (30)$$

Using the above correlations, the energy balance equations (24) and (26) have the following final forms:

$$\frac{dT_1}{dt} = \frac{P_2\,(4.177 - 2.506X_2)(T_2 - T_1) - U_2A_2(T_1 - T_2)] + U_1A_1\,(T_S - T_1)}{M_1\,(4.177 - 2.506X_P)} \quad (31)$$

$$\frac{dT_2}{dt} = \frac{F\,(4.177 - 2.506X_f)(T_f - T_2) + U_2A_2(T_1 - T_2)] + V_2[\,4.177T_2 - H(T_1)]}{M_1\,(4.177 - 2.506X_2)}$$

2.3. *Simulation Procedure*

The mathematical model was built for the evaporator in the form of a set of systems, and each system component with a set of subsystems which represents the mathematical model equations for evaporator. The Equations (8), (14),(31) and (32) are used in the simulation works of the open-loop system. The values of mathematical parameters of the double-effect evaporator for tomato system are shown in the Table 1.

Table 1.

Operating parameters for Tomato system [9].

Parameter	Unit	Value
M_1	(kg)	2268
M_2	(kg)	2268
F	(kg/hr)	26103
S	(kg/hr)	11023
P_1	(kg/hr)	5006
P_2	(kg/hr)	14887
V_1	(kg/hr)	9932
V_2	(kg/hr)	11165
X_f	(kg/kg)	0.05
X_1	(kg/kg)	0.2607
X_2	(kg/kg)	0.0874
T_s	(°C)	115.7
T_f	(°C)	85.0
T_1	(°C)	74.7
T_2	(°C)	52.0
A_1	(m^2)	102
A_2	(m^2)	412
U_1	(kJ/hr.m^2.°C)	5826
U_2	(kJ/hr.m^2.°C)	2453

3. RESULTS AND DISCUSSION

3.1. The dynamic behavior of the open loop system

Simulation results for the evaporator temperature responses at different step changes in feed flow rate, feed temperature and feed concentration were obtained in simulation.

3.1.1 Effect of Feed Temperature

The outlet temperatures of evaporator are increased by increasing the temperature of feed. The speed of temperature response of the second evaporator is larger than the first evaporator because of present delay time in the second effect of evaporation process and the first effect reach to uniform condition with more time delay. The dead time was noticed to decrease with increasing the temperature of feed. This is because increasing the temperature of feed results in the acceleration of the evaporation process. All those are shown in Figs. 2 and 3

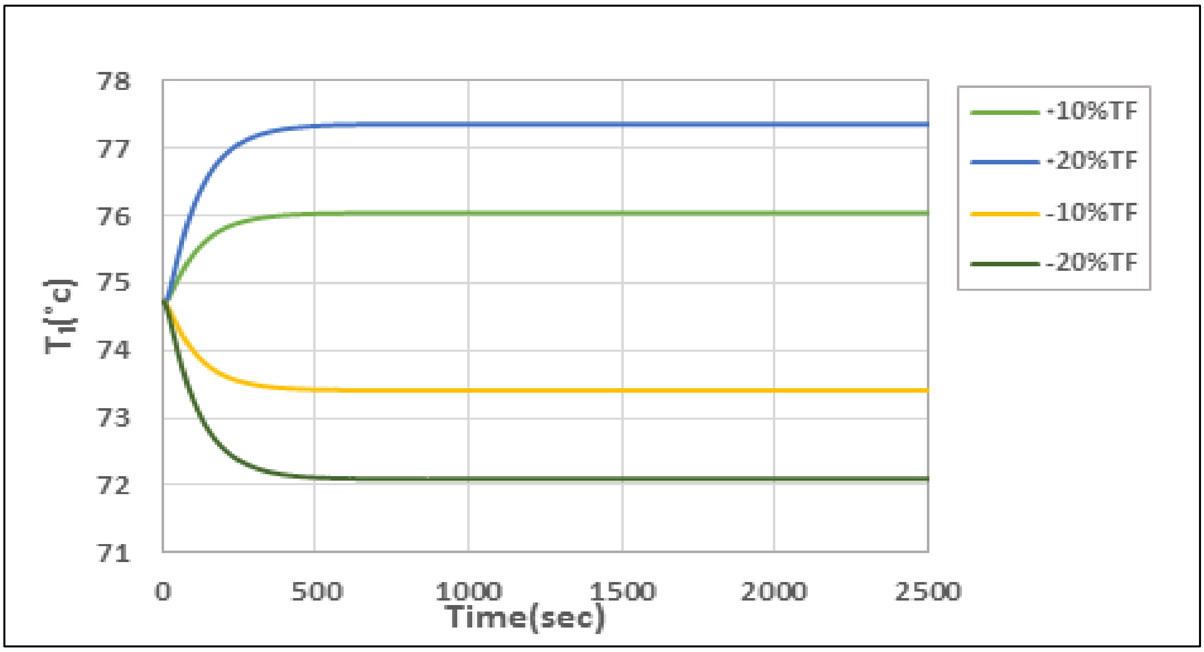

Fig. 2. The open loop simulation temperature response of first evaporator to step change ($\pm$10% and $\pm$20%) in temperature of feed.

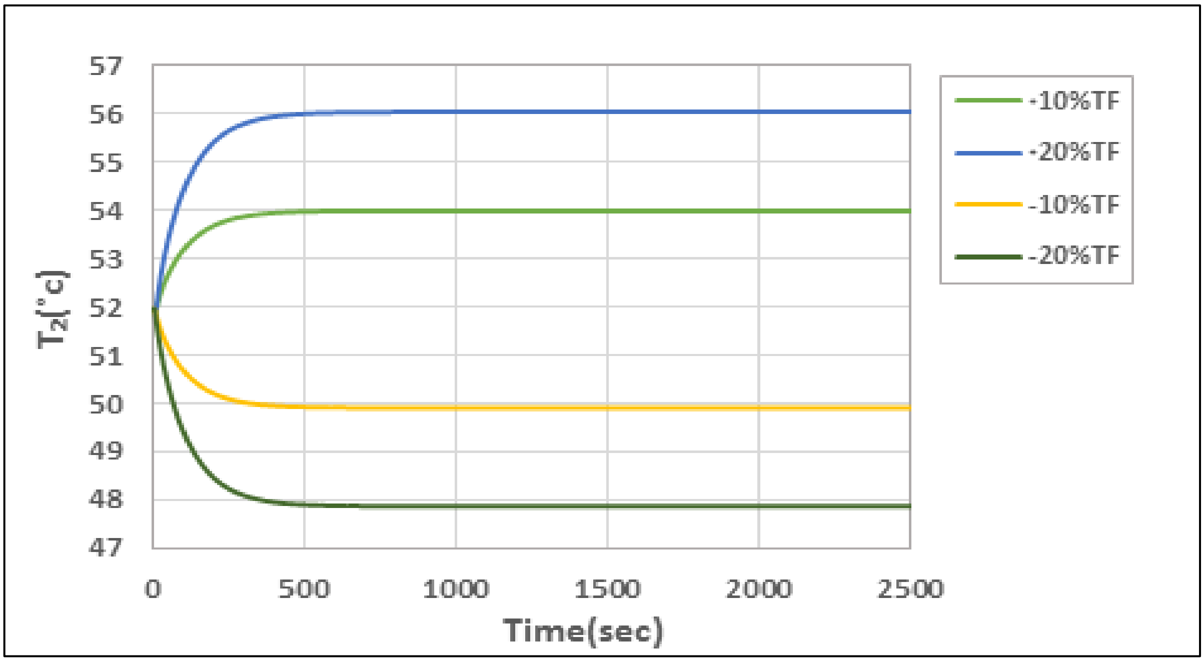

Fig. 3. The open loop simulation temperature response of second evaporator to step change ($\pm$10% and $\pm$20%) in temperature of feed.

3.1.2 Effect of Feed Concentration

In Figs. 4 and 5 observed that the outlet temperatures of evaporator are low effect when a disturbance in the concentration of feed. As the results showed that the increase or decrease in concentration of feed material ($\pm$10%, $\pm$20%) did not have any strong effect on the evaporator temperature.

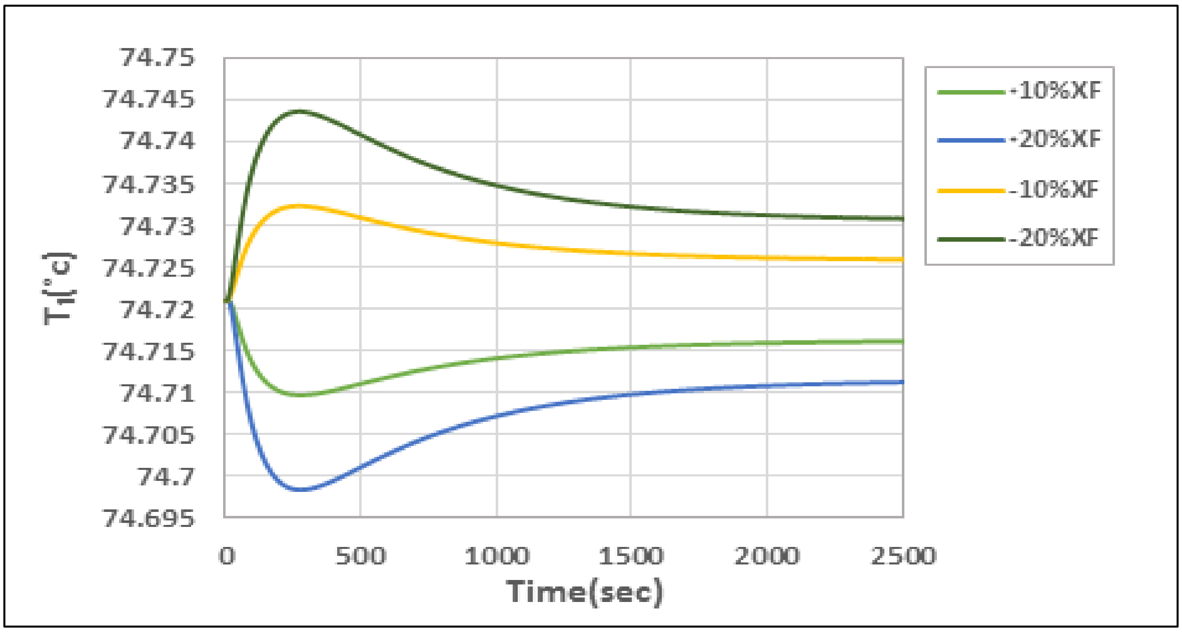

Fig. 4. The open loop simulation temperature response of first evaporator to step change (±10% and ±20%) in feed concentration.

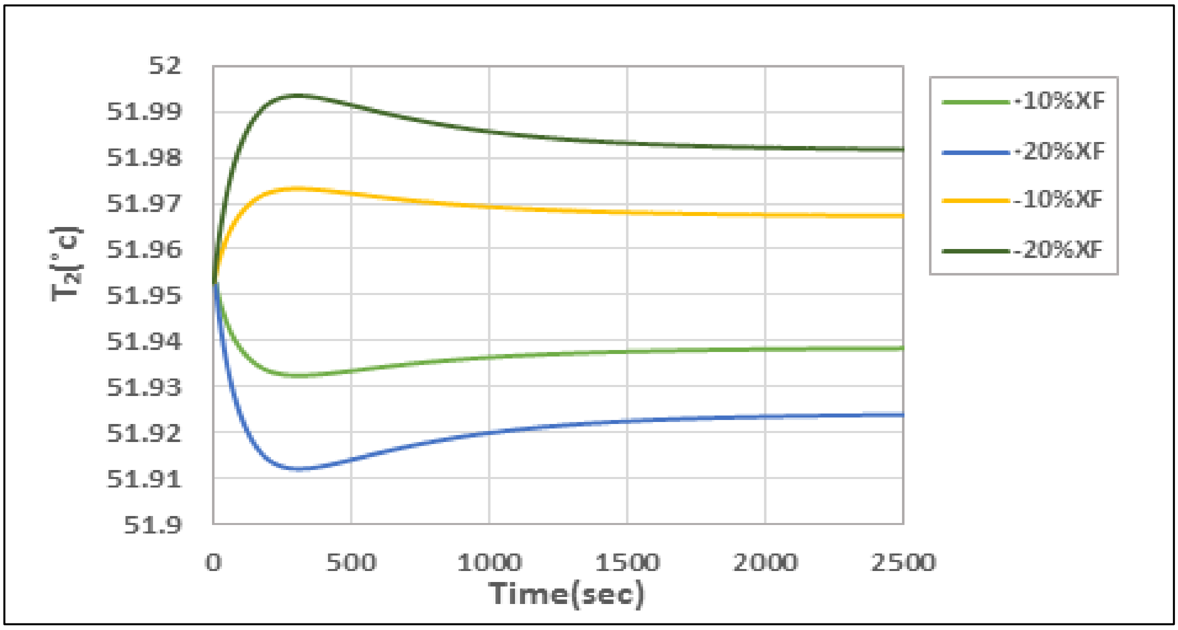

Fig. 5. The open loop simulation temperature response of the second evaporator to step change (±10% and ±20%) in feed concentration.

3.1.3 Effect of Feed flow Rate

Feed flow rate is one of the main causes in the existing of disturbances in evaporation process. The outlet temperature of each effect increase with the feed flow rate increase, because the feed temperature is larger than evaporator temperature, so caused accumulation of heat in the evaporator. All these are shown in Figs. 6 and 7.

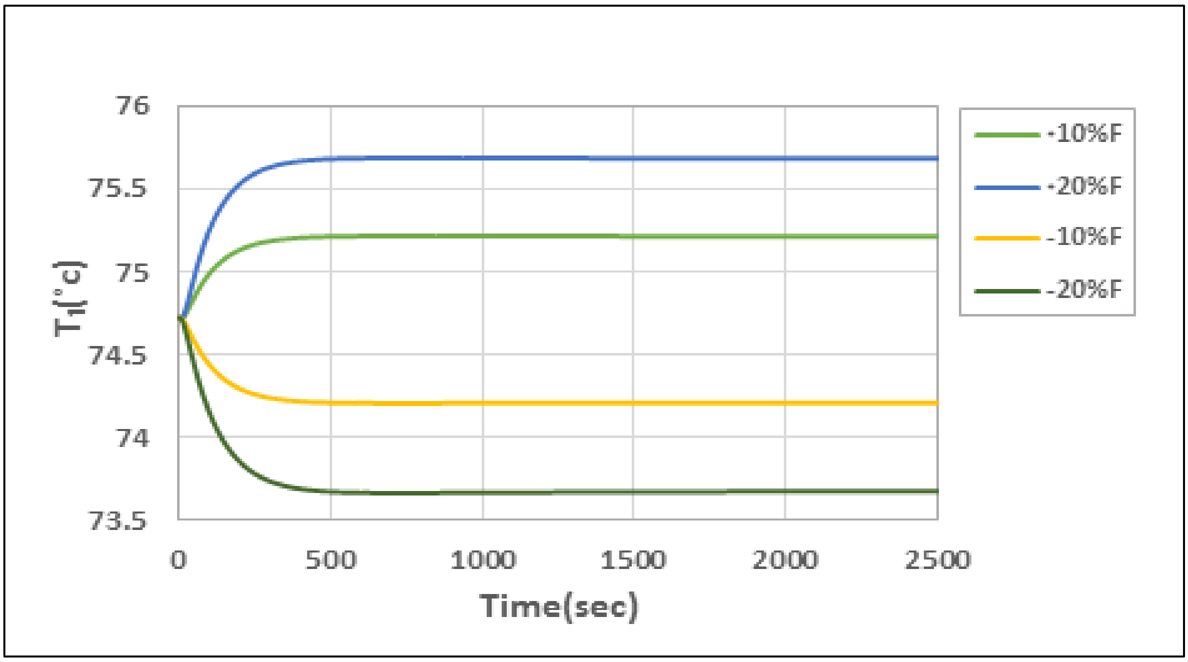

Fig. 6. The open loop simulation temperature response of first evaporator to step change (±10% and ±20%) in feed flow rate.

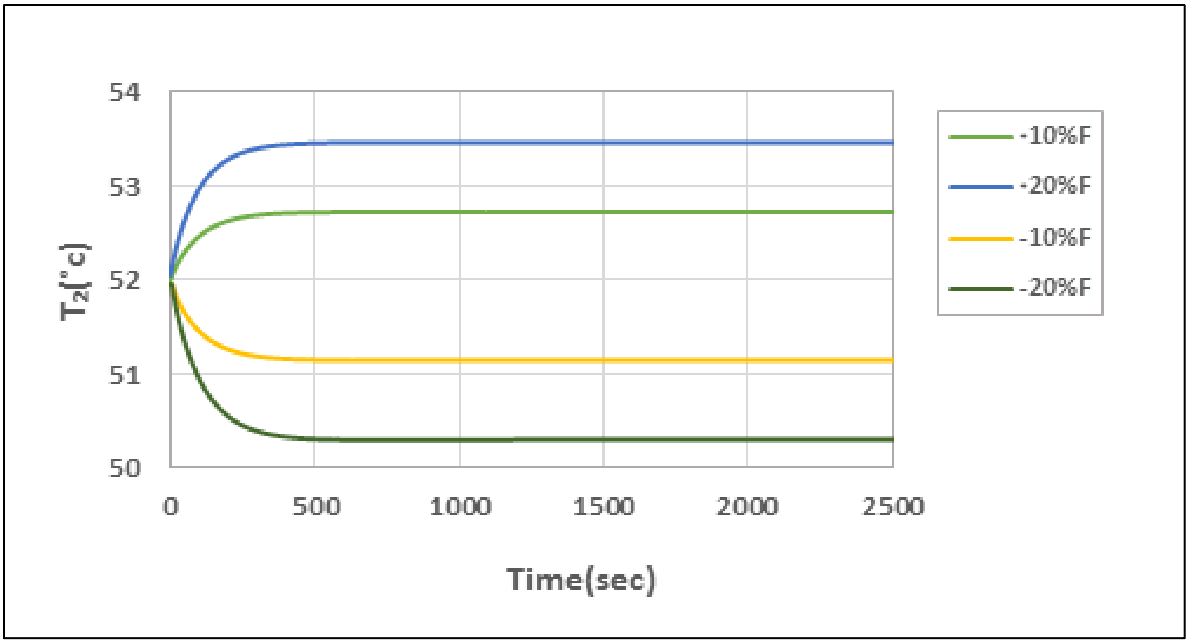

Fig. 7. The open loop simulation temperature response of second evaporator to step change (±10% and ±20%) in feed flow rate.

3.2. The dynamic behavior of the closed-loop system

3.2.1 Proportional-Integral-Derivative (PID) controller of the evaporator.

Initially, the outlet temperature of the first evaporator is 74.72 °C and 51.95 °C of the second evaporator which represent the desired value at 14887kg/hr feed flowrate and 85 °C feed temperature. The response of PID controller to step change in feed temperature and step change in feed flowrate on the same drawing in the rate of ±10% is shown in Figs. 8 and 9. Fig. 8 shows that the controller responses of first evaporator with the increase in feed temperature and of the feed flow rate in the rate of +10%. The first evaporator temperature increases about 0.48 by increasing of feed temperature and increasing about 0.18 by increasing of flow rate of feed. Fig. 9 shows that the increase in second evaporator temperature 0.6 and about 0.25 when increasing feed temperature and increasing feed flow rate in the rate of +10% respectably. So the second evaporator temperature is increasing is more than the first evaporator and it takes less time to reach the desired value.

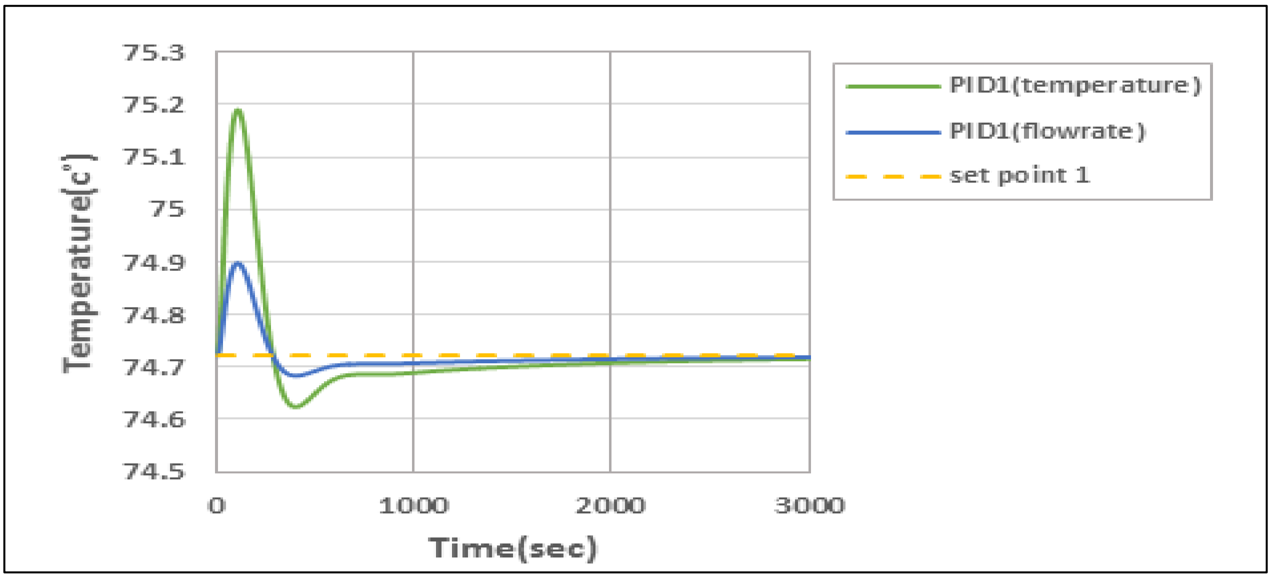

Fig. 8. Temperature response of first evaporator for PID single control to a step change in flow rate of feed and a step change in temperature of feed in the rate of +10% at set point =74.72℃.

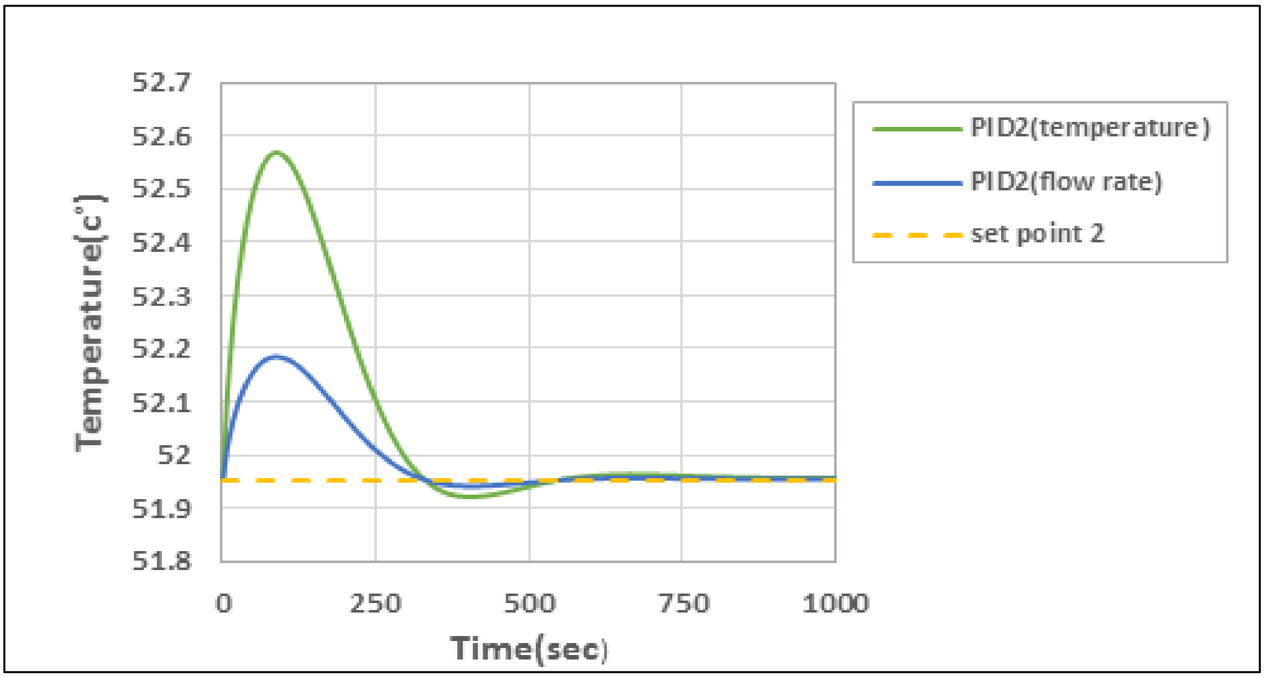

Fig. 9. Temperature response of second evaporator for PID single control to a step change in flow rate of feed and a step change in temperature of feed in the rate of +10% at set point =51.95℃.

Decrease in feed temperature and feed flow rate in the rate of -10% as shown in Figs. 10 and 11 decreases temperature response of each evaporator and decrease in the second evaporator is more than that in the first evaporator. The Table 2 shows the integral absolute error (IAE) for PID controllers.

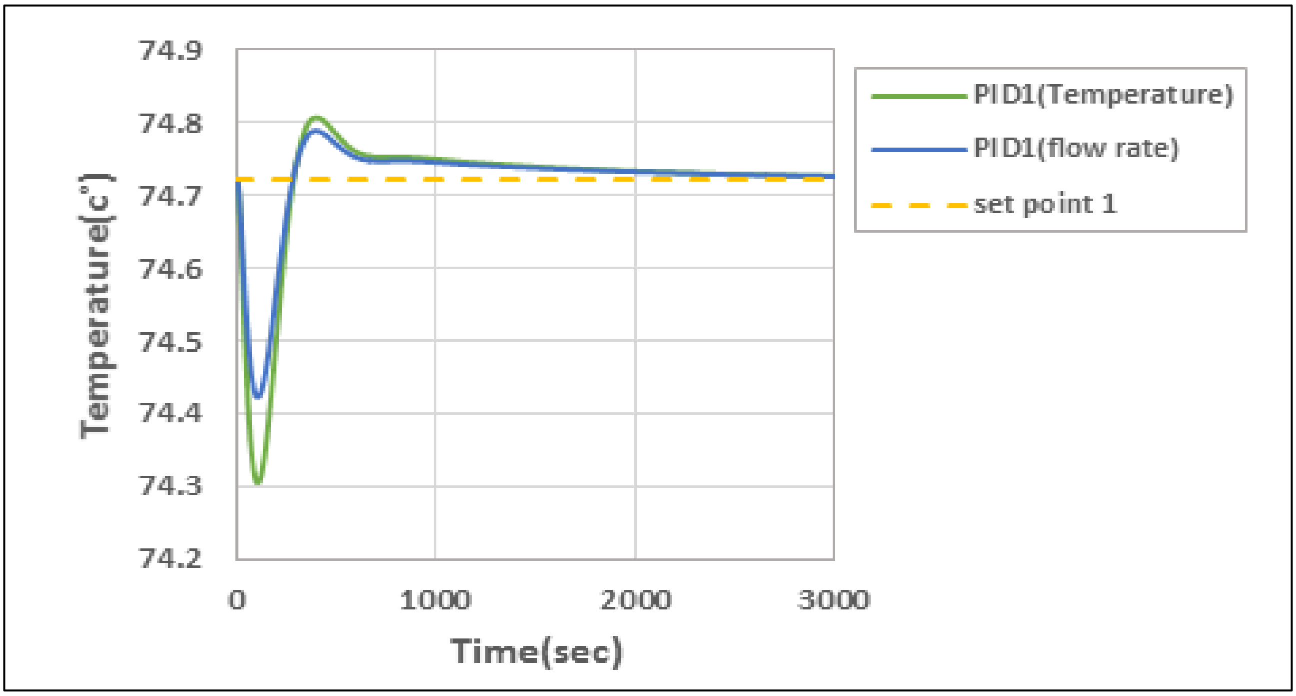

Fig. 10. Temperature response of first evaporator for PID single control to a step change in flow rate of feed and a step change in temperature of feed in the rate of -10% at set point =74.72°C.

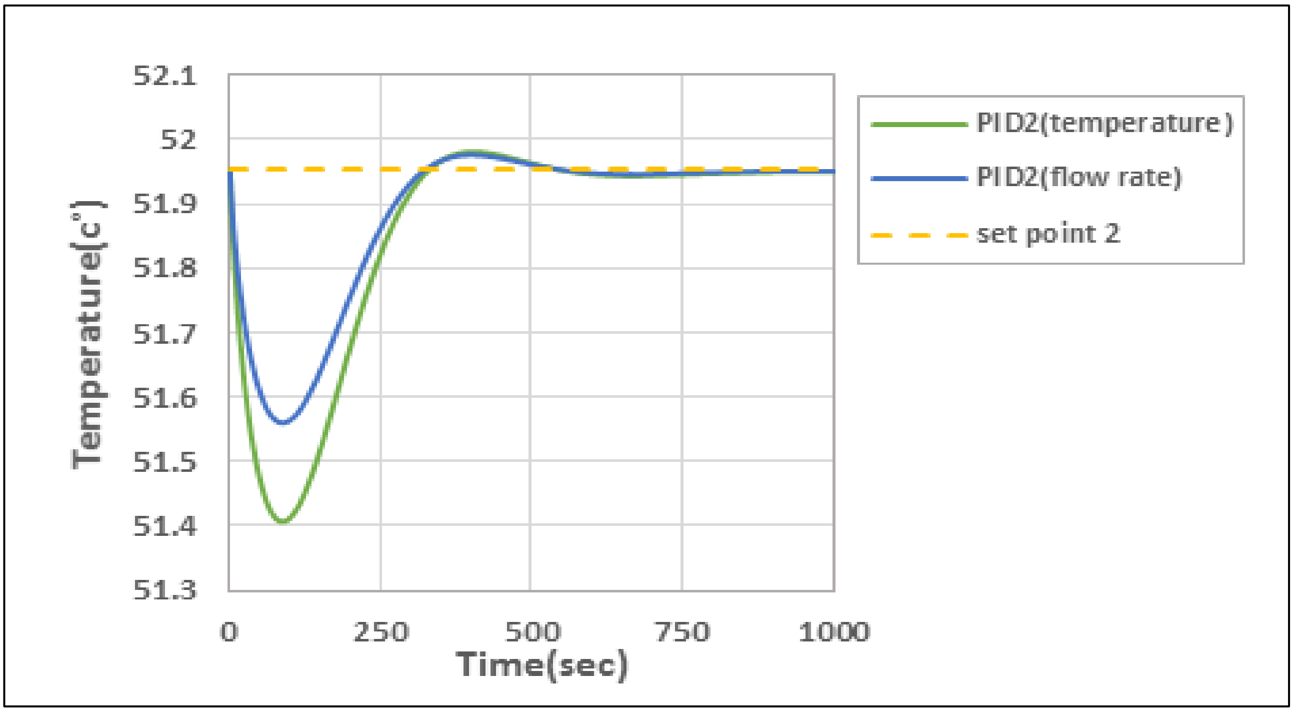

Fig. 11. Temperature response of second evaporator for PID single control to a step change in flow rate of feed and a step change in temperature of feed in the rate of +10% at set point =51.95°C.

Table 2.

The integral absolute error (IAE) for PID controllers

Variable of step change	Values of step Change	PID1	PID2
Feed temperature, °C	+10%	1.6	114.8
	-10%	1.35	102.04
Feed flow rate(kg/hr)	+10%	3.31	43.6
	-10%	9.03	72.7

3.2.2 Model predictive controller (MPC) of the evaporator.

Fig. 12 shows that the increase in the first evaporator temperature is about 0.09 when increasing feed temperature in the rate of +10% and about 0.03 when increasing feed flow rate in the rate of +10%. The Fig. 13 shows that the increase in the second evaporator temperature is about 0.02 when increasing feed temperature in the rate of +10% and increase about 0.01 when increasing feed flow rate in the rate of +10%. So that the increase in first evaporator temperature is more than in the second evaporator temperature, and the speed of first evaporator to the desired value is less than that of the second evaporator. The Table 3 shows the integral absolute error (IAE) for MPC controllers.

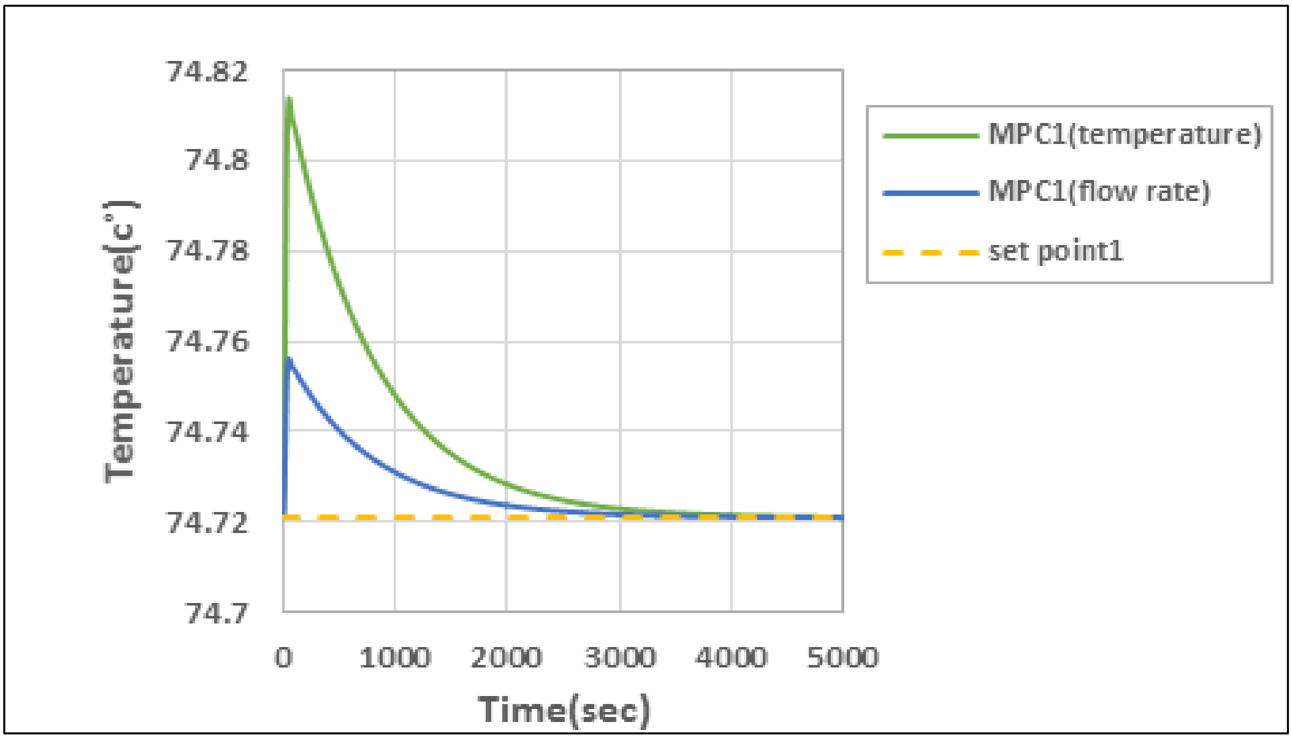

Fig. 12. Temperature response of first evaporator for MPC single control to a step change in flow rate of feed and a step change in temperature of feed in the rate of +10% at set point =74.72°C.

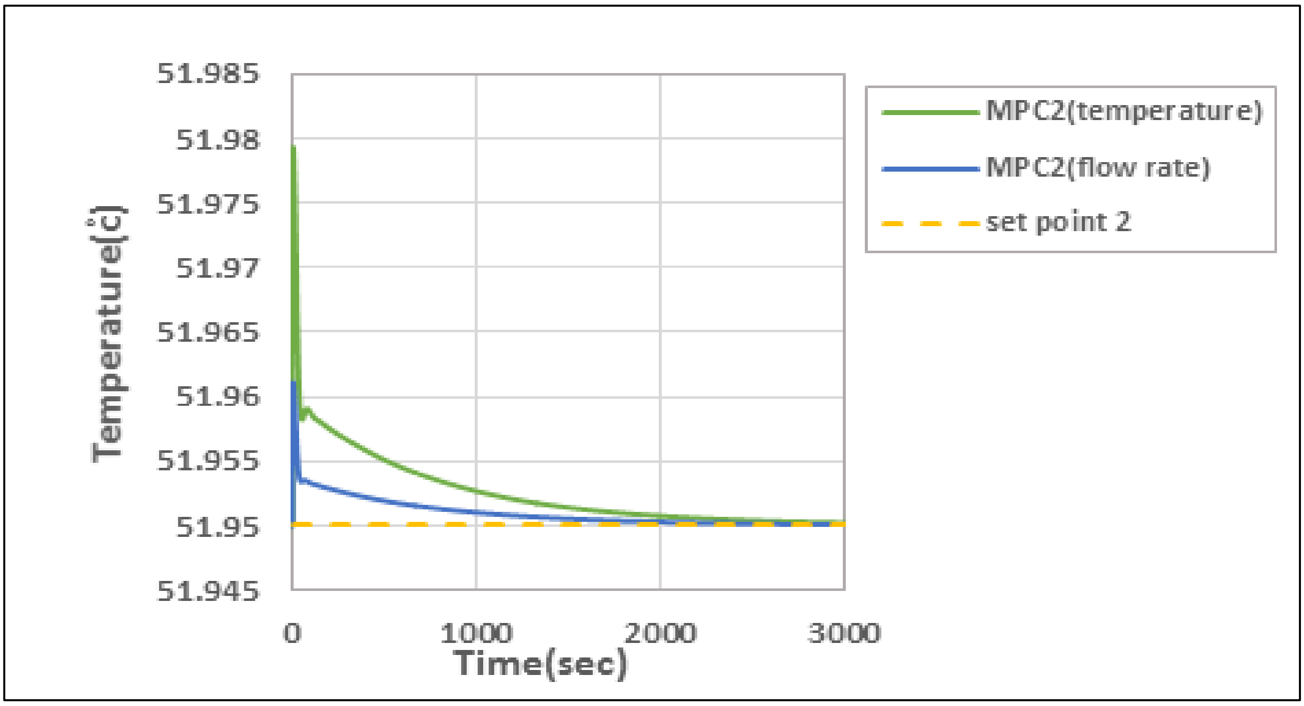

Fig. 13. Temperature response of the second evaporator for MPC single control to a step change in flow rate of feed and a step change in temperature of feed in the rate of +10% at set point =51.95℃

Figs. 14 and 15 show that the first evaporator temperature decrease is more than the second evaporator temperature when decreasing in feed temperature and feed flow rate in the rate of -10% and the speed of the second evaporator to the desired value is more than that of the first evaporator

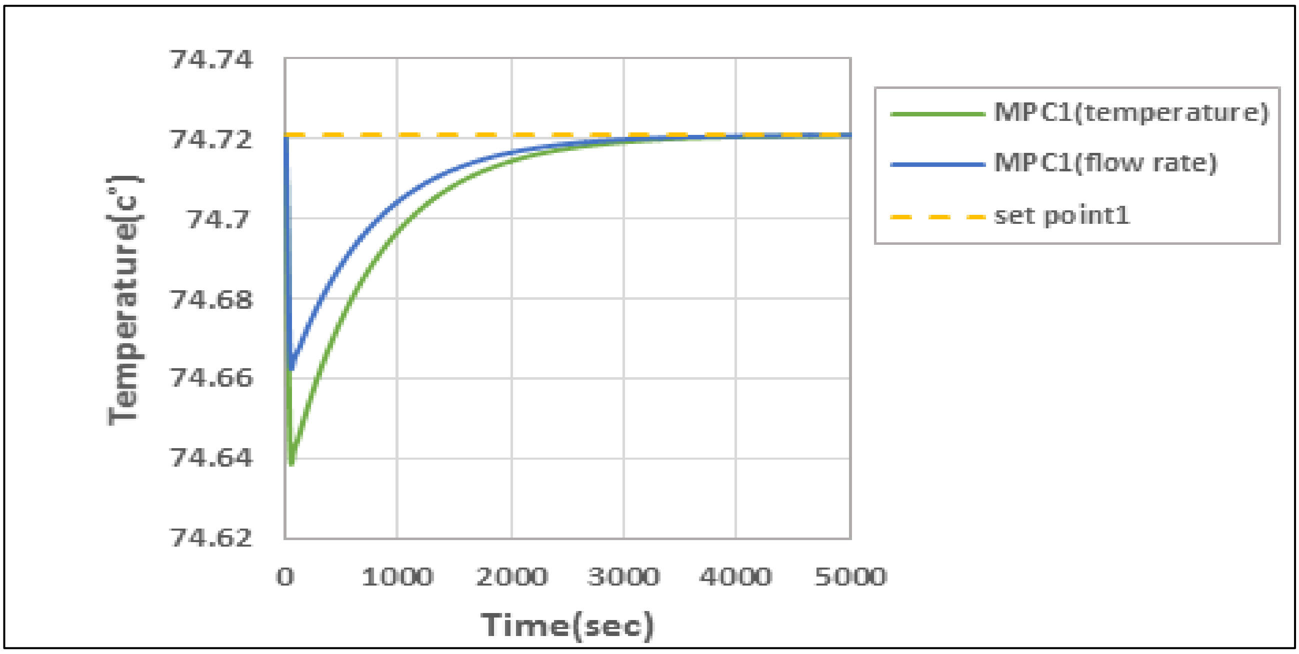

Fig. 14. Temperature response of first evaporator for MPC single control to a step change in flow rate of feed and a step change in temperature of feed in the rate of -10% at set point =74.72℃.

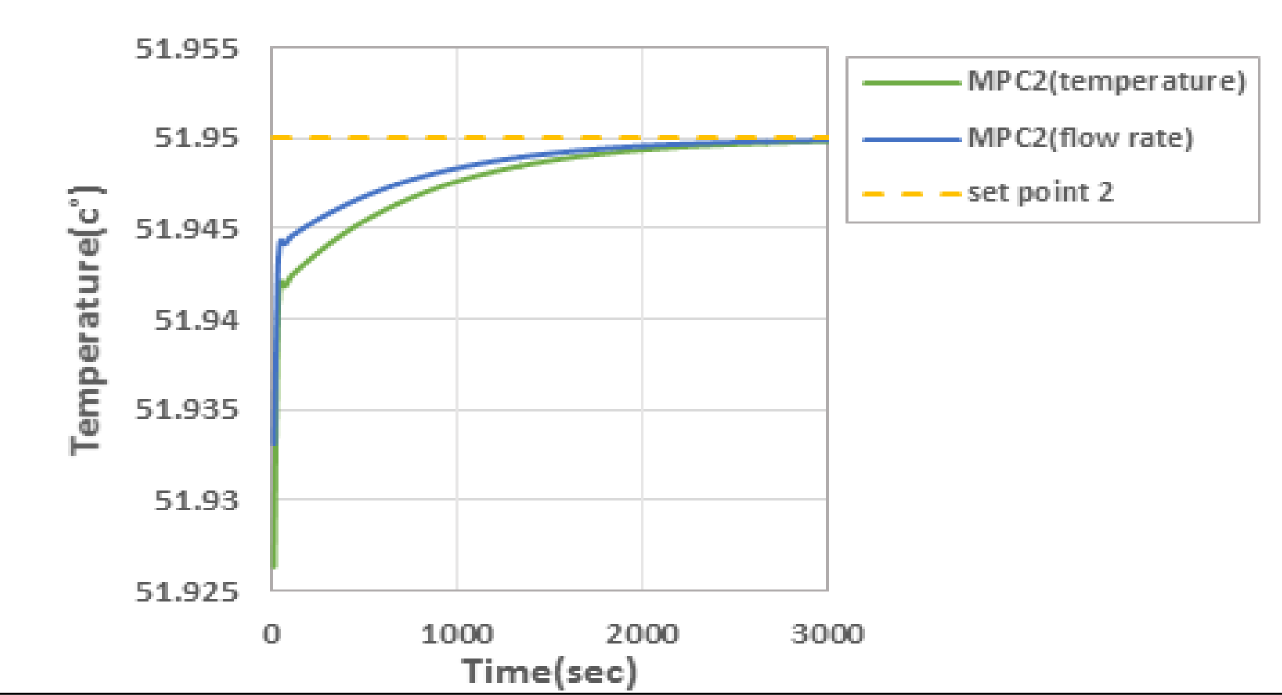

Fig. 15. Temperature response of the second evaporator for MPC single control to a step change in flow rate of feed and a step change in temperature of feed in the rate of +10% at set point =51.95℃.

Table 3.

The integral absolute error (IAE) for MPC controllers

Variable of step change	Values of step Change	MPC1	MPC2
Feed temperature, °C	+10%	74.1	7.92
	-10%	65.83	7.034
Feed flow rate(kg/hr)	+10%	27.72	2.97
	-10%	45.83	4.92

4. CONCLUSIONS

The study of dynamic simulations showed that the process is strongly influenced by temperature more than other variables, where increasing the temperature of the feed leads to an increase in the temperature produced by the evaporators. In addition to that, the increase in the flow rate of the feed leads to an increase in the heat of the evaporators, but with a lower effect of the heat of the feed while the feed concentration showed an effect that is almost no heat generated from evaporators. Two methods are tested to control a wide range of operating conditions and simulation results show that there is good accuracy. The MPC controller is more accurate than the PID control and faster to reach the constant value, but that is at the expense of other factors such as (steam) which were not taken into account, where the rate of flow increases at MPC controller. The responses of PID controllers to two steps change in set point shows that the PID controller is efficient and can perform the control of the process over wide range of operating variables.

Nomenclature

P Flow rate of liquid product (kg/hr).

V Flow rate of vapor (kg/hr).

F Flow rate of feed input (kg/hr).

X Mass fraction (kg solids/ kg stream)

S Flow rate of steam input to the first effect (kg/hr)

λ Latent heat (kJ/kg).

U Overall heat transfer coefficient (kJ/hr.m².℃)

A Heat transfer area (m²)

T Temperature (℃)

h(T,X) Enthalpy of liquid product(tomato juice).

H (T) Enthalpy of pure vapor solvent (saturate steam) at temperature T (kJ/kg).

h(T) Enthalpy of pure liquid solvent (condensate) at temperature T (kJ/kg).

Subscripts

f Feed

p Final product

S Steam

Sp Set point

0 Steady state

1 First effect

2 Second effect

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