The modelling and simulation of a drying process in a poultry by-product processing plant

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Abstract: This paper studies the modelling and simulation of a drying process in a poultry by-product processing plant. In a poultry by-product processing plant, the material is separated into liquids, fats and solids. The liquids are removed, either by pressing or drying, the fats are collected in a centrifuge and the solid mass is dried and manufactured into a protein-rich solid powder. The process contains two rendering operations, where heat is applied to the material to both remove extra moisture and to raise the temperature of the material. These processes are vital for the sterilization of the material as well as for the quality of the product. To improve the understanding behind the drying process, a mathematical model is created using first principles and the obtained model is complemented with experimental data from the real-world process to produce a dynamic model of the drying process. The model is built on a digital computer, using a simulation software. The model is validated with data from the real process plant and used to reveal the underlying dynamics that complicate the control of the process. The model enables the design of automatic control for the process in a safe environment.

Keywords: Drying, first principles, dynamic model

1. INTRODUCTION

Drying is a complex and still incompletely understood process (Mujumdar, 1987) that is used in many industries, e.g. agriculture (Delele *et al*., 2015), food (Canales *et al.*, 2001) and mining (Moon *el al*., 2014). The quality and the structure of the dried material, disturbances affecting the drying process and the highly nonlinear behaviour of the drying phenomenon all contribute to the difficulty of controlling the process with an acceptable accuracy. Manual control, where operators adjust the manipulated variables based on the behaviour of the process and based on their expertise in drying processes, is still used (Dufour, 2006). However, because of the increasingly competitive markets and the strong environmental incentives regarding energy usage, more effective control strategies are called for. Most automated control strategies require either an experience-based model or a numerical model of the process they intend to control. Many different models of drying processes have been developed, such as the models created by Iguaz et al. in 2003 and Abbasfard et al. in 2013, but they do not apply to any general case. Thus, the creation of a specific model for each unique case is warranted.

The objective of an indirect heater is to heat the material to a desired set-point whilst removing moisture from the material. This process of stabilizing raw material with heat is called rendering (van der Veen *et al.*, 2004). In an indirect heater the product and the heating medium are separated by a wall, thus not allowing the material to be in direct contact with air or other heating medium. The use of an indirect heater allows for various combinations of input materials without the risk of contamination or combustion (Kimball, 2001). Other benefits include energy efficiency, better product quality and environmental friendliness, since there is no need to clean the heating medium, e.g. steam, because of no direct contact with the material (Jamaleddine *et al.*, 2010).

A model of the drying process is created with first principles using moisture and energy balances. First-principles modelling is used since there is no possibility to gather all the necessary data from the process and the physical laws involved in the process are well known. The process is modelled and simulated in Simulink environment (Klee *et al.*, 2011) to improve the understanding behind the dynamics of the process and to allow simulation of different inputs and control strategies and their effect on the outcome. The model is validated using real-process data and the model is used to design and implement a control strategy for the process.

In Section 2, the theoretical background involving the drying phenomenon and heat-transfer is given. In Section 3, the experimental environment is introduced, and the dryer model is derived using mathematical modelling. In Section 4, the results of the simulations are presented and in Section 5, a proposal for the control strategy is presented. In Section 6, the conclusions of the paper are presented.

2. THEORETICAL BACKGROUND

In this Section, the fundamentals of both the drying process and the heat-transfer processes are introduced.

2.1 Drying

Drying is a process in which liquid is removed from a solid through evaporation (Yliniemi, 1999). When the drying process begins, two simultaneous phenomena occur:

- heat from a surrounding heating medium is transferred to the solid. In direct heating the heat is often provided by hot air and in indirect heating cases the is supplied through contact with a hot surface e.g. a metal plate

- moisture is transferred as a vapor from the surface of the solid and as a liquid or as a vapor within the solid

A large amount of different theories regarding moisture migration phenomena exist in the literature and the reader is referred to Öchsner *et al*. for an extensive review. However, the applicability of these migration mechanisms to real-world problems is still lacking. Often the design of dryers is based on the investigation of external conditions and the internal conditions are described by curves which state the rate of drying as a function of time. Typical drying rate curves and more thorough explanation is given in (Mujumdar, 1987) and (Yliniemi, 1999).

2.2 Nomenclature

NOMENCLATURE

Symbol	Description (unit)
A	area (m^2)
B	constant
C	constant
C_p	specific heat capacity (J/kg·ºC)
c_i	specific heat capacity of a food component (J/kg·ºC)
D	constant
h	level of a volume element (m)
h_f	latent heat of vaporization (J/kg)
K	drying constant
m	mass (kg)
Q	rate of heat transfer (W/s)
Q_p	heat loss through the shell of the dryer (W/s)
q_{in}	mass flow into a volume element (kg/s)
q_{out}	mass flow out of a volume element (kg/s)
q_h	mass flow due to the level difference (kg/s)
q_{evp}	mass flow of the evaporated water (kg/s)
R	valve coefficient (m$^{(5/2)}$ ·s^{-1})
R_w	rate of drying (s^{-1})
T	temperature (ºC)
T_m	temperature of the material (ºC)
T_s	temperature of the steam (ºC)
T_{in}	temperature of the material flowing into a volume element (ºC)
T_{out}	temperature of the material flowing out of a volume element (ºC)
ΔT_{mean}	logarithmic mean temperature difference (ºC)
ΔT_A	temperature difference of the cold and hot streams at the start of a heat exchanger (ºC)
ΔT_B	temperature difference of the cold and hot streams at the end of a heat exchanger (ºC)
t	time (s)
U	overall heat transfer coefficient (W/m^2·K)
X	moisture content of the material ($kg_{water}/kg_{full\ product\ mass}$)
X_e	equilibrium moisture content ($kg_{water}/kg_{full\ product\ mass}$)
X_{in}	moisture content of the material flowing into a volume element ($kg_{water}/kg_{full\ product\ mass}$)
X_{out}	moisture content of the material flowing out of a volume element ($kg_{water}/kg_{full\ product\ mass}$)
x_i	mass fraction of a food component

2.3 Rate of drying

The rate of drying R_w is one of the most important parameters when describing a drying phenomenon (Iquaz *et al.*, 2003). It is unique for each scenario and must be experimentally determined for each case. The rate of drying should contain at least moisture-solid equilibrium data and a drying curve obtained in near conditions to the process it models (Lopez *et al.*, 2000).

Different models have been developed. The following drying rate equation was introduced by Lopez *et al.* in 2000:

$$R_w = K(X - X_e), \tag{1}$$

where *K* is a drying constant determined experimentally, *X* is the solid moisture content and X_e is the equilibrium moisture content. Singh *et al.* in 2012 proposed a simpler model for the drying rate and expressed it as an exponential equation. The model assumes the equilibrium moisture content to be relatively small compared to the actual moisture content, thus reducing the equation to be:

$$R_w = ae^{bX}, \tag{2}$$

where *a* and *b* are constants determined by drying rate experiments and *X* is the moisture content of the material. The use of a specific model for each case must be justified according to the material and process properties.

2.4 Heat transfer

For the process of drying to begin, a heat transfer method is required. The most common heat transfer methods are convection, conduction, infra-red radiation and dielectric heating. Usually multiple heat transfer methods are working simultaneously in a dryer; however, one method is usually dominant. In an indirect tube rotary drier this dominant method of transfer is conduction, in which the required heat is provided to the solid by contact with a heated surface, which in turn, is heated from the inside with hot steam. The mathematical theory behind heat conduction was first developed by Joseph Fourier in the nineteenth century (Fourier, 1955). For more information regarding different heat-transfer methods the reader is referred to Saxena *et al*. and Serth.

2.5 Overall heat transfer coefficient

Couper *et al.* in 2004 presented the transfer rate of heat between a solid and a heating medium in a heat exchanger as:

$$Q = UA\Delta T_{mean}, \tag{4}$$

where *Q* is the rate of heat transfer, *U* is the overall heat transfer coefficient, *A* is the area of contact surface between the heating surface and the solids and ΔT_{mean} is the logarithmic mean temperature difference defined as:

$$\Delta T_{mean} = \frac{\Delta T_A - \Delta T_B}{\ln\left(\frac{\Delta T_A}{\Delta T_B}\right)}, \tag{5}$$

where ΔT_A and ΔT_B are the temperature differences of the cold and hot streams at both ends of the heat exchanger.

However, in cases where one of the temperature streams remains a constant, such as heaters with a pressure-controlled steam heating, the logarithmic mean temperature difference is not applicable. In these cases, a temperature difference of the heating medium and the heated material can be employed to replace the term ΔT_{mean}. The transfer rate of heat then becomes:

$$Q = UA(T_s - T), \tag{6}$$

where T_s is the temperature of the heating medium, often steam or oil, and T is the temperature of the heated material.
The overall heat transfer coefficient U must be determined experimentally, through pilot plant experiments or process expertise.

3. EXPERIMENTAL ENVIRONMENT

In an indirect rotary tube disc dryer, the material is isolated from the heating medium. The heating medium passes through a series of discs that are located inside the shell and the heat is transferred to the material from the surface of those discs. Steam is used as the heating medium in the temperature range of 150 to 180 °C. The material is inserted into the dryer from one end and discharged from the other end, while the heating medium is flowing in countercurrent direction. The rotating discs are equipped with small paddles that increase the flow and the mixing of the material throughout the tube.

The heating medium can be repurposed to other plant processes since there is no odor, dust or emissions released in the heating process. The indirect heating method also eliminates the risk of contamination and the risk of combustion inside the dryer. The slow movement of the material through the dryer prevents degradation and abrasion to the material, thus not affecting the materials' final characteristics (Kimball, 2001).

3.1 Formulation of the drying model equations

A rotary tube disc dryer can be modelled as n consecutive interconnected volume elements, where the output of the first element is the input of the next element. A visualization depicting these well-stirred interconnected volume elements is shown in Figure 1.

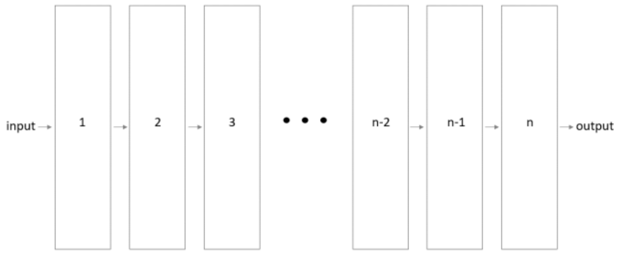

Figure 1. Interconnected volume elements.

Iguaz *et al.* in 2003 expressed the change in moisture content in a volume element as:

$$\frac{\partial(mX)}{\partial t} = q_{in}X_{in} - q_{out}X_{out} - q_{evp} \tag{7}$$

or as:

$$\frac{\partial(mX)}{\partial t} = m\frac{\partial X}{\partial t} + X\frac{\partial m}{\partial t} \ . \tag{8}$$

Combining (7) and (8) we obtain:

$$q_{in}X_{in} - q_{out}X_{out} - q_{evp} = m\frac{\partial X}{\partial t} + X\frac{\partial m}{\partial t} \tag{9}$$

and rearranging to obtain the change in moisture content:

$$\frac{\partial X}{\partial t} = \frac{1}{m}[q_{in}X_{in} - q_{out}X_{out} - q_{evp} - X\frac{\partial m}{\partial t}], \tag{10}$$

where X is the mass of water of the material related to the overall product mass ($0 < X < 1$), m is the total mass of the product, q_{in} is the total mass flow of the material flowing in, q_{out} is the total mass flow of the material flowing out and q_{evp} is the mass flow of the material evaporated from the product.

Similarly, Iguaz *et al.* expressed the change in enthalpy in a volume element as:

$$\frac{\partial(mC_pT)}{\partial t} = q_{in}C_pT_{in} - q_{out}C_pT_{out} + UA(T_s - T_m) - q_{evp}h_f - Q_p \tag{11}$$

or as:

$$\frac{\partial(mC_pT)}{\partial t} = mC_p\frac{\partial T}{\partial t} + mT\frac{\partial C_p}{\partial t} + TC_p\frac{\partial m}{\partial t} \ . \tag{12}$$

Combining (11) and (12) we obtain:

$$q_{in}C_pT_{in} - q_{out}C_pT_{out} + UA(T_s - T_m) - q_{evp}h_f - Q_p = mC_p\frac{\partial T}{\partial t} + mT\frac{\partial C_p}{\partial t} + TC_p\frac{\partial m}{\partial t} \tag{13}$$

and rearranging to obtain the change in temperature:

$$\frac{\partial T}{dt} = \frac{1}{mC_p}\Big[q_{in}C_pT_{in} - q_{out}C_pT_{out} + UA(T_s - T_m) - q_{evp}h_f - Q_p - mT\frac{\partial C_p}{\partial t} - TC_p\frac{\partial m}{\partial t}\Big] \quad (14)$$

where C_p is the specific heat capacity of the material, U is the overall heat transfer coefficient, A is the area of contact between the heating plates and the material, h_f is the latent heat of vaporization and Q_p is the heat lost through the shell of the dryer. Even though the derivation of C_p is unnecessary, it is included for the sake of completeness and in refence to Iquaz *et al.*

The mass flow q through the process is considered to consist of two parts: a constant part which is caused by the rotating discs that move the material and a sliding part which is caused by the level difference between two adjacent volume elements. For example, in a case where the discharge screw is stopped altogether, the mass starts to accumulate in the last volume element; when the level of the material rises in the last element, some of the material slides back to the previous element and when the level of the second to last element increases, some of its material slides back to the third to last element and so on. In the case of the material being as fluid as water, the sliding would be near instant, but the material in this study does not slide as quickly. The flow is assumed incomprehensible and totally turbulent. The mass flow occurring due to this dynamic is modelled through

$$q_h = \mathrm{sqn}(h_{n-1} - h_n) \cdot R\sqrt{|h_{n-1} - h_n|}, \quad (15)$$

where R is the coefficient which determines the speed of the flow, here called the valve coefficient. The variables h_{n-1} and h_n are the levels of two adjacent volume elements. The basis for the equation is derived from the Bernoulli's law (Mujumdar, 1987).

3.2 Parameter acquisition

The physical parameters required in the equations must be determined either by experiments or by literature review. The most important parameters for accurate results are the drying rate function and the overall heat transfer coefficient. However, these two parameters are the most difficult ones to determine.

The mass flow of the material evaporated can be formulated as:

$$q_{evp} = R_w m \quad (16)$$

where R_w is the rate of drying.

A lot of research has been done to determine the rates of drying for various products and it has been found to vary depending on the material and the environment (Singh *et al.*, 2012). The drying rate for this model was determined by experimental tests since no previous research had been done on this kind of process to determine accurate drying rates. An experiment was conducted in the real process environment by temporarily stopping the dryer and measuring the moisture content of samples from five locations along the dryer length. The locations of the data gathering points were dictated by the four vents in the process, which made the sample gathering possible. The locations of the data sample gathering points are illustrated in Figure 2.

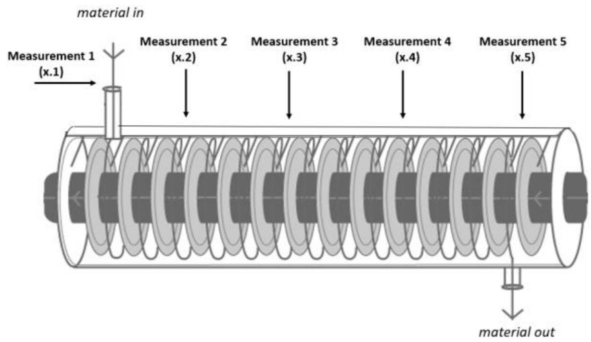

Figure 2. Locations of the sample gathering points.

A long metal pipe was used to gather a small piece of the material from the dryer. The samples were then put into a moisture measurement device, which weighs the samples and then heats them until they are free of any moisture. The final weight determines which was the original amount of moisture in the sample. The samples were taken approximately every hour for six consecutive hours.

The experiment was difficult to conduct due to challenging process environment. However, the acquired data can be used to generate an estimate of the drying rate curve for the process. In Table I, the averages of the measurements are presented.

If we consider a fixed speed for the material flow and the physical locations of the sample gathering points, the drying rate has clear phases. The initial material moisture content drops down from 62.55% to 44.91% in approximately 500 seconds whereas the moisture content from 16.54% to 4.45% takes 1300 seconds. The drying rate is over 3.7 times higher in the beginning of the process than in the last part of the process. This correlates with the assumption that the material has free moisture in the beginning of the process, which is fast

TABLE I
THE AVERAGE MOISTURE CONTENT AND TEMPERATURE OF THE SAMPLES

Location	Temperature (^{0}C) of the material	Moisture content (%) of the material
1	54.0	62.55
2	99.1	44.91
3	100.7	31.73
4	103.3	16.54
5	117.4	4.45

to evaporate, and slows down when heat energy is required to both move the moisture from inside the material to the surface and to evaporate it. In Figure 3 the drying rate is plotted as a function of time and as a function of moisture content. The data is fitted using Curve Fitting Toolbox in Simulink.

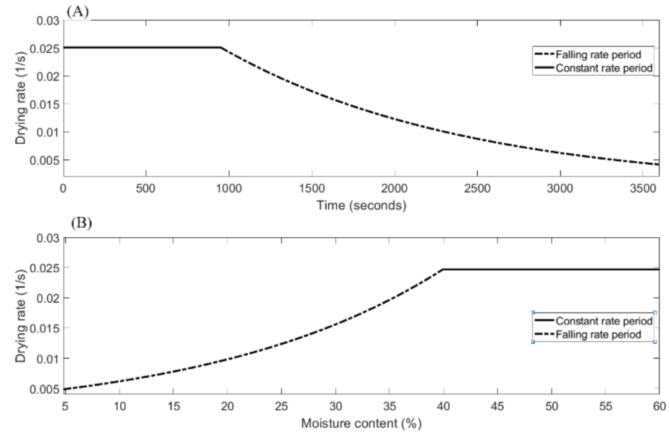

Figure 3. The drying rate curves. (A) Drying rate as a function of time. (B) Drying rate as a function of moisture content.

The drying rate as a function of moisture content is used in the simulations to determine the speed of which the material loses its moisture. The expressions for the functions in the simulator are presented below in (17):

$$\begin{cases} f(X) = D, & X > 40 \\ f(X) = B\exp(CX), & X < 40 \end{cases}, \tag{17}$$

where X is the moisture content of the material (*%*) and *B, C* and *D* are constants.

The drying rate function obtained this way is not ideal. However, it gives the model a rough estimate of how the drying rate changes in different phases of the process. More research into the drying rate of this specific poultry by-product material and this specific process is warranted.

The overall heat transfer coefficient U is another important parameter that is difficult to obtain precisely. No previous research exists into this kind of material in this kind of an environment. Herz *et al.* in 2012, Li *et al.* in 2005, Schlunder in 1981 and Nhuchhen *et al.* in 2016 have obtained heat transfer coefficient values ranging from 200 to 450 (W/m^2K) in drying of copper beads in a rotary dryer. A value of 200 (W/m^2K) was used as the starting point for the model and with information available on how much the material increases in temperature throughout the process in the real-process plant, the overall heat transfer coefficient was fine-tuned by simulating the drying process with different coefficient values. A realistic temperature value is obtained with an overall heat transfer coefficient value of 150 (W/m^2K).

The area of heating surface in the dryer changes when the levels of the volume elements change. This occurs especially in cases where the inflow of the material is disturbed, or the speed of the discharge screw is altered. As the heating surface in the dryer consists of the outer shell, the discs and the shaft in the middle of the dryer, the area of heating surface does not behave linearly. To calculate the area of heating surface in contact with the material, the dryer is assumed to be a horizontal cylinder. The calculation of the area of heating surface is achieved via an iterative Matlab function, which finds the corresponding solid mass height h for a given area A. With the knowledge of the solid mass height h the total heating surface area can be calculated.

The specific heat capacity of the material in question is calculated through:

$$C_p = \sum c_i x_i, \tag{18}$$

where c_i is the specific heat capacity of a food component and x_i is the mass fraction of that food component. The specific heat capacities of the food components are gathered in research conducted by Choi and Okos in 1986 and the mass fractions are gathered from information at the process plant. The specific heat capacity used in the calculations is 3500 (J/kg^oC) for the solid material and 4190 (J/kg^oC) for water. The combined specific heat capacity varies depending on the amount of water in the material.

The valve coefficient R that determines the speed of the flow due to level difference in each volume element is not possible to be determined through experiments due to unavailability of necessary measurements from the process. After studying simulation results and real process data along with comments from the process operators, the mass flow due to level difference is set to be 15% of the amount of the constant mass flow. This sets the valve coefficient R to be 0.00012 $m^{5/2}/s^{-1}$. The obtained value is an estimate and would require further research to be determined more accurately. Other parameters used in the balance equations are obtained from the process plant. A compilation of the parameters and their values is shown in Table II.

A dynamic simulator is built based on first principles modelling. The simulator is built in MATLAB computational environment using the Simulink extension. Even though the simulations are done with a digital computer, the process is continuous in nature. Because of that, the calculations are done in continuous time.

TABLE II
PARAMETERS USED IN THE BALANCE EQUATIONS

Parameter	Value	Unit
Tin	50	ºC
Xin	0.62	-
qin	0.00060	m^3/s
Qp	0	W/s
U	150	W/m^2K
B	0.003890	-
C	0.04627	-
D	0.02470	-
hf	2258	kJ/kg

The dryer is divided into 40 interconnected volume elements, where the output variables of the first volume element are used in the next volume element as input variables. The dryer is controlled manually by plant operators. The manipulated variables are the steam pressure and the speed of the discharge screw. The discharge screw reacts instantly to control changes, whereas the steam valve is modelled as a first-order process with a time constant of 8 minutes.

4. RESULTS

The purpose of this study was to provide extra information about the complex drying process and to determine the underlying dynamics that dominate the process. The model is simulated in different scenarios to both validate it and to expose the underlying dynamics that should be taken into account when operating the process.

The control of the dryer is difficult if the process behaviour is not understood. The simulations are thus focused on demonstrating the underlying dynamics of the process. The dryer measurements are scarce, and the inflow material varies both in temperature and in moisture content. The claws on the heating plates move the material through the process as they should, however, this dynamic is not consistent in speed and the actual velocity of the material moving through the process is unknown. Because of these difficulties, the model cannot predict the outflow temperature accurately. Luckily, we are more interested in the dynamics of the process than the actual temperature value since the final temperature value can be fine-tuned according to real process tests. The understanding of the process operation also allows for automated control to be implemented and enables the planning of similar processes in future plants.

The model is first validated with real-process data. The steam pressure is altered periodically while the discharge speed is altered at the same time. The steam pressure and discharge screw speed alterations are shown below in Figure 4.

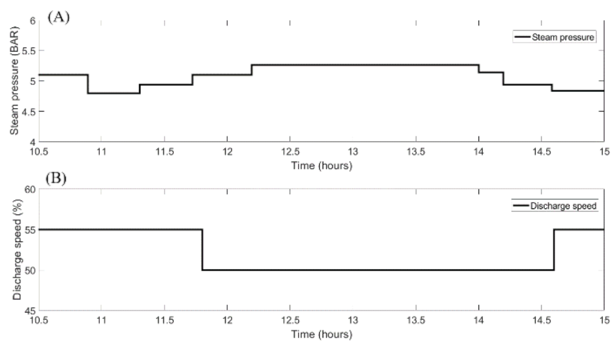

Figure 4. The alterations to control variables at the real process plant. (A) Steam pressure alterations. (B) Discharge speed alterations.

These alterations are also conducted in the dryer model to simulate the behaviour of the model with the same inputs. The temperature curves from both the real process plant and the simulation are shown in Figure 5. The simulated temperature curve reacts to steam pressure changes faster than the real process. This might be due to the simulation receiving the changes instantly, while the real process temperature measurement device is in the bottom of the dryer and might not receive the altered material temperature as fast. The model also assumes the material to be uniform in temperature, while this is not the case in the real process. Nevertheless, the model correctly predicts the dynamics of the process temperature even during multiple control actions.

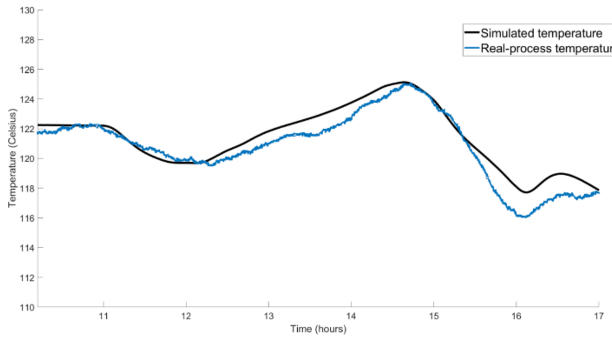

Figure 5. The temperature of the material in the simulations and in the real process plant.

The control actions are simulated separately to demonstrate their individual effect on the process. The effect of increasing or decreasing the speed of the discharge screw is demonstrated in Figure 6.

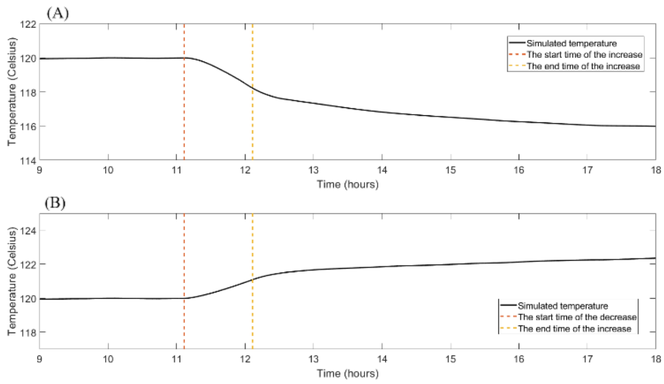

Figure 6. The effects of increasing and decreasing the discharge screw speed. (A) Increasing the discharge speed. (B) Decreasing the discharge speed.

In Figure 6, the discharge speed is altered for one hour and then returned to the original value. The material temperature reacts quite fast but keeps changing slightly for a long time period after the change has been reversed. If the discharge screw is stopped altogether, the effect is more severe, as illustrated in Figure 7. The discharge screw is stopped for 15 minutes, resulting in a quick increase in temperature. However, the temperature of the material keeps increasing for several hours after the discharge stop has ended and the process becomes difficult for process operators to stabilize.

Another severe disturbance to the material temperature is caused by a sudden lack of material flowing in. The sudden drop in material flow in the first phase of the drying process decreases the amount of heating surface the material comes in contact with during its residence time in the dryer and ultimately lowers the final temperature. The lack of material also causes the levels of the volume elements to start changing, which complicates the control of the process even more. The disturbance lowers the temperature over a long time period, affecting the final temperature for up to several hours. In Figure 8, the inflow of the material is decreased drastically for 30 minutes and its effect on the final temperature of the material is shown.

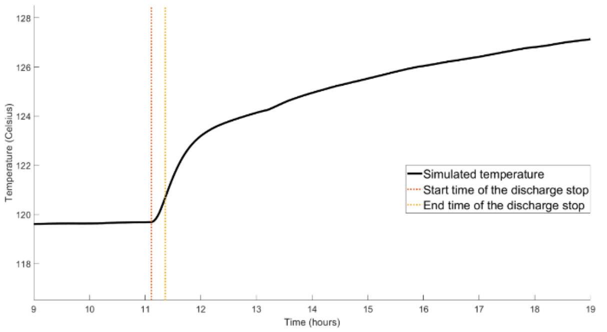

Figure 7. The effect of stopping the discharge screw.

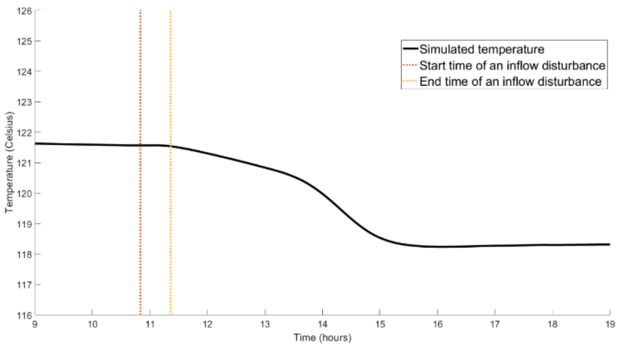

Figure 8. The effect of disturbing the inflow of the material.

5. DISCUSSION

The model guides the design of a control strategy for the process, revealing the underlying dynamics and assisting in choosing the right input-output combinations. The model supported the change from the traditional manual operating culture to a better performing feedback control strategy. The mental model of the plant personnel was also transformed, and their understanding of the plant process was improved.

The plant is traditionally operated by manipulating both the control variables, the steam pressure and the discharge screw speed. However, due to the long-lasting implications of manipulating the input variables too quickly, the control strategy proposed here is to let the steam valve control the steam pressure and to make the feedback loop fast. The steam pressure is set to a constant setpoint and it is not used to control the material temperature. Instead, the discharge screw is set to control the temperature of the material and the feedback loop is set to be deliberately slow. This way the discharge screw keeps the temperature close to the wanted setpoint and does not react unnecessarily fast to disturbances. This way, the effect of the two control variables interacting with each other is decreased, leading to a more stable control. The proposed input-output combinations are shown below in Figure 9.

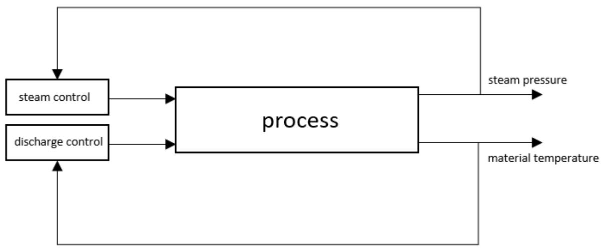

Figure 9. The proposed input-output combinations.

6. CONCLUSIONS

A model of a drying process was created in this paper.
The model was obtained through first principles modelling and was complemented with experimental data obtained from a poultry by-product processing plant. The model correctly predicted the temperature of the material during multiple control actions. The model is used to demonstrate the underlying dynamics of the drying process and to help the process operators in controlling the process more accurately. The model is also used to design a control strategy for the process, relieving the operators of manual control. The design of the control structure is safe with the model, as no risk for the process equipment or the material quality is involved. Further research is warranted for the parameters of the model, e.g. the drying rate and the overall heat transfer coefficient.

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