Int. Agrophysics, 2002, 16, 191–195

INTERNATIONAL
Agrophysics
www.ipan.lublin.pl/int-agrophysics

Energy aspects in food extrusion-cooking

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Received February 25, 2002; accepted May 20, 2002

Abstract. Theoretical and practical energy balance considerations in food extrusion-cooking are presented in the paper. Based on the literature review as well as on own measurement results, the baro-thermal treatment of different vegetable raw materials is discussed together with the engineering aspects of the extruders' performance as a whole.

Keywords: extrusion-cooking, energy balance, specific mechanical energy

INTRODUCTION

Extrusion-cookers give the opportunity to combine pumping, mixing, kneading and heating operations in one machine (Fig. 1). As a consequence of this combination, however, the different operations interact with each other and can only be separated to a certain extent. An important objective in designing extruders is to define these interactions and to predict the influence of changes in parameters on the performance of the extruder as a whole [1,2].

Many of the parameters needed for food extrusion models are unknown, usually changed considerably during the process and are related to a wide variety of other parameters. Many of the numerical descriptions that can be of great benefit for the description of the extrusion of synthetic polymers may deviate considerably from the actual results in extrusion – the cooking of food, thus limiting their value in this field. Therefore, in evaluating the energy balance, it is of great importance to be acquainted with material properties like viscosity.

In plastics extrusion, viscosity is a unique function of temperature and shearing, but in extrusion-cooking, chemical and physical changes occur during the process. This immediately implies that viscosity is not only a function of

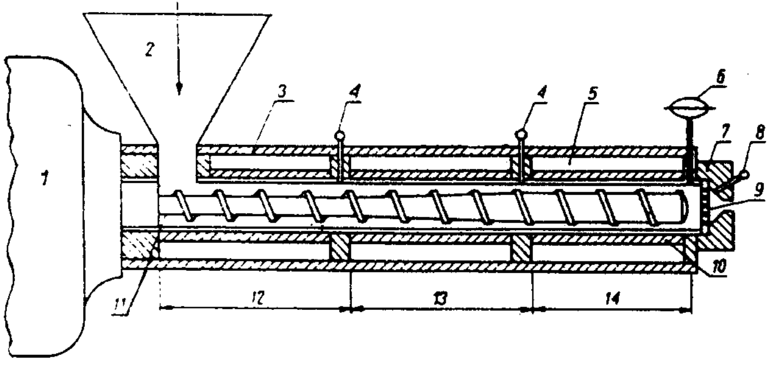

Fig. 1. Cross-section of a typical food extruder: 1 – drive, 2 – feed hopper, 3 – cooling water jacket, 4 – thermocouples, 5 – barrel steam jacket, 6 – pressure transducer, 7 – die, 8 – discharge thermocouple, 9 – breaker plate, 10 – barrel with hardened liner, 11 – screw with increasing root diameter, 12 – feed section, 13 – compression section, 14 – metering section [1].

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*Corresponding author's e-mail: moscicki@faunus.ar.lublin.pl instant temperature and shearing but also to a large extent a function of temperature history. Physical cross-linking and gelatination may modify the viscosity and contribute to a complex rheology that can hardly be interpreted as a mere change of state (melting). This can result in process instabilities and problems in control [2].

It is well known that within normal operating ranges, starches and protein-rich materials are shear thinning. This justifies the use of a power-law equation for viscosity. For changing temperatures, the power-law equation may be combined with temperature effects [3]:

$$\mu_a = \mu_0 |\dot{\gamma}|^{n-1} \exp\left[-\beta(T - T_0)\right], \tag{1}$$

where μ_a is the apparent viscosity (N s m^{-2}), μ_0 is the Newtonian viscosity (N s m^{-2}), n is the power-law index, $\dot{\gamma}$ is the shear rate (s^{-1}), β is a constant, T is the temperature (K), T_0 is a reference temperature (K).

Both starch-rich and protein-rich materials show an increase in viscosity when extrusion-cooked. This may be attributed to network formation of the molecules, in protein-rich material by cross-linking and in starch-rich materials by entanglement of the amylose and amylopectin chains. One may assume that the network formation may roughly be described as a first-order reaction and that the viscosity increase is roughly proportional to the thus – formed concentration of cross-links or entanglements. Therefore, the viscosity of a fluid element with a residence time θ in the extruder may be expressed as [2]:

$$\mu_a = \mu_0 |\dot{\gamma}|^{n-1} \exp\left\{(-\beta T)\left[1 - \exp\left(\int_0^{\theta} K \exp\frac{\Delta E}{\mathrm{R}T(t)} Dt\right)\right]\right\}, \tag{2}$$

where ΔE is the activation energy (J), R is gas constant (8.3143 J mol^{-1}K^{-1}), t is time (s). Dt denotes that the integration must be performed in a co-ordinate system travelling with the fluid element; as a result $T(t)$ is the temperature as a function of the time that the fluid particle experiences while travelling through the extruder.

Depending on the actual values of the constants β and ΔE and the temperature profile, the viscosity of the material may increase or decrease during extrusion [2,9]. A small decrease in throughput (e.g., by a small increase of die resistance) may change the hold up, increase the residence time, and therefore increase viscosity. If this viscosity change strongly affects the pressure built up at the die, the pressure flow increases and the throughput may decrease further, especially in an extruder with soft material or insufficient grooves. This gives rise to instabilities. If, on the other hand, the increase of viscosity affects the back flow most strongly, the process becomes more stable [2]. Further investigation into the dependence of the stability on the actual value of the parameters β and ΔE in connection with the temperature profile is needed.

ENERGY BALANCE

The extrusion-cooker is a thermodynamic unit: for steady-state operation, it means all energy that is introduced into the apparatus must also come out again. An energy balance consists of four terms [2,8,9]:

- mechanical energy added by the rotation of the screw,
- heat transferred through the barrel wall,
- mechanical energy partly used to increase the pressure of the material,
- mechanical energy partly converted into heat by viscous dissipation.

The thermal energy (generated by viscous dissipation or transferred through the wall) results in an increase in the temperature and phase changes (e.g., melting of solid material or evaporation of moisture). The energy balance can be described as [2]:

$$E + E_h = Q_v\left(P + \rho c_p \Delta T + \rho e\right), \tag{3}$$

where E is the effective motor power that is transferred to the screw(s) (W), E_h is the net heat added through the wall (W), Q_v is the volumetric throughput (m^{3}s^{-1}), P is pressure at the die opening (Pa), ρ is the density (kg m^{-3}), c_p is the specific heat of the food material (J kg^{-1}K^{-1}), e is the phase change enthalpy per unit weight includes the energy needed for chain splitting (J kg^{-1}), for cross-linking in protein rich materials and for the generation of new surfaces when the material expands.

If the product temperature is measured after the material has left the die instead of before, the pressure energy has also been transferred into heat and the energy balance reduces to [2]:

$$E + E_h = Q_v\left(\rho c_p \Delta T + \rho e\right), \tag{4}$$

where ΔT is now the temperature change measured just after the die, before any cooling by convective and radiative losses.

To establish the ratio between energy added by the drive unit and energy transferred through the wall, the Brinkman number must be used defined as [2]:

$$\mathrm{Br} = \mu v^2 / \lambda \Delta T, \tag{5}$$

where μ is the viscosity (N s m^{-2}), v is a representative velocity (m s^{-1}), λ is the thermal conductivity (W m^{-1}K^{-1}), ΔT is the temperature difference between the food material and the barrel wall (K).

Assuming that the heat needed to melt solid fractions of the material is much smaller than that needed for heating the material and evaporation of the moisture, two extremely simple and useful equations can be used. If no evaporation of moisture at the die end occurs, the final material temperature (T_f) is given by:

$$T_f = T_0 + \frac{E}{Q_v \rho c_p}. \qquad (6)$$

When a moisture content of fraction *f* evaporates at the die, Eq. (6) can be modified as [2]:

$$T_f = T_0 + \frac{E - f e_w Q_v \rho}{Q_v \rho c_p}. \qquad (7)$$

Here e_w denotes the phase change enthalpy of the vaporising component (for water, $e_w = 2257.8$ kJ kg^{-1} at 100°C). Taking into account that the temperature rise of the material during extrusion ($T_f - T_0$) is a unique function of motor power, throughput, and material properties, the analysis above is particularly useful since its application is not restricted to one particular type of machine. With simple modifications, various other effects (cross-linking, chain splitting, surface generation) can be taken into account [2].

Knowledge of moisture flash at the extrusion-cooker outlet is needed to perform material and energy balances around the extrusion system. The most common approach to obtaining this information is to attempt a sample of the product as it leaves the extrusion die, however that gives very imprecise moisture analyses. The moisture of the extrudate changes so rapidly that any variation in the collection of the sample and the sealing of the sample container results in a sample whose moisture varies appreciably [5].

The energy lost by the extrudate as it passes through the die is equal to the energy available for the evaporation of steam on the law-pressure side of the die. From a thermodynamics perspective, the enthalpy of the steam entering the die is equal to the enthalpy of the exiting steam.

As previously indicated, the extruder imparts energy into the extrudate via the dissipation of mechanical energy and/or the transfer of thermal energy. This energy results in heating the extrudate. In fact, it is the storage of the viscous and thermal energy inputs as thermal energy in the extrudate. Since the temperature associated with the equilibrium water vapour pressure on the law-pressure side of the die, some of the water will be converted into steam until equilibrium is attained. The energy 'stored' in the extrudate must be conserved during this process. That is, the stored energy that entered the die is equal to the total energy in the two exit streams (puffed extrudate and steam). This can be described by a very simple heat balance. One must know the temperature on the high-pressure side of the die and the equilibrium vapour pressure of water on the discharge side of the die. This pressure defines the temperature at which the flash occurs. There is always fresh air mixed with the steam in the discharge area, so appreciable vaporisation occurs even though the extrudate has cooled to below 100°C. The observations of many authors suggest that the water flashes at about 85°C. Using the assumption of an 85°C flash, the heat balance to estimate the mass of steam G_s released per unit mass of wet extrudate is [5]:

$$G_s = c_p \left(T_{die} - 85\right) / H_{vap}, \qquad (8)$$

where G_s is steam formed (kg kg^{-1}), c_p is the heat capacity of the wet extrudate (J kg^{-1}K^{-1}), H_{vap} is the heat of vaporisation of water (2.26 MJ kg^{-1}), T_{die} is the temperature behind the die (K).

To estimate the c_p of the extrudate requires a proximate analysis for water, protein, carbohydrate, fat and the ash of the extrudate, and the data summarised in the literature [5,10]. The heat capacity of the extrudate is simply the weighted average of the heat capacities of the individual components. For example an extrudate which is 25% water, 10% protein, 59% starch, 5% fat and 1% ash, the heat capacity $c_p = 2.61$ kJ kg^{-1}K^{-1}. If the temperature behind the die is 150°C, 75 g steam per kilogram of wet extrudate will be liberated. A simple water balance reveals that the final moisture of the product would be 18.9%.

In the absence of any heat inputs or losses, such as steam injection or venting, barrel heating or cooling, or convective or radiative losses, that is, in the adiabatic condition, the temperature rise of the extrudate can be found from the following relationship [6]:

$$c_p \Delta T + m_t H_t = \text{SME}, \qquad (9)$$

where ΔT is the temperature change of the extrudate (K), m_t is the mass of extrudate which can undergo phase transformation per unit mass of extrudate) (kg kg^{-1}), H_t is the energy associated with the phase transformation (J kg^{-1}), SME is the specific mechanical energy input of the extruder's motor (kWh kg^{-1}).

During phase transformation (e.g., gelatinization of starch and denaturation of protein) a reasonable estimation of the energy required is calculated for 17 J g^{-1} [8]. The SME is usually calculated from the percent torque of the extruder motor and its speed or by direct measurement with a watt meter [1,6]. If the raw materials enter the extruder at 30°C and the SME is 0.1 kWh kg^{-1} (a typical value), the 25% moisture extrudate described above exhibits a temperature rise of approximately 133°C, or a die temperature of approximately 163°C. Any value significantly different from this, assuming that the die temperature measurement is correct, indicates that an appreciable heat transfer from other sources is taking place.

The heat transfer from all sources may be included in the analysis with a simple modification of the following equation [6]:

$$c_p \Delta T + m_t H_t = \text{SME} + \text{STE}, \qquad (10)$$

where STE is the specific thermal energy (kWh kg^{-1}) from other heat sources or sinks.

A negative value of STE represents a heat loss, a positive value is a heat input. For large industrial extruders, in the absence of steam injection or venting, the magnitude of STE is about 0.03 kWh kg^{-1}, or less, because large extruders have little surface area per unit volume. Small laboratory extruders may exhibit much larger values of STE. The key point to remember is that any analysis of extrusion behaviour must include an estimate of both SME and STE since ultimately the quality of the product is controlled by both of these parameters [6].

Taking into account the previous example, we can easily estimate the STE from the die temperature measurement. If the exit temperature was measured as 143°C, instead of the predicted adiabatic value of 163°C, the STE would be – 0.015 kWh kg^{-1} (a heat loss). Conversely, if the exit temperature was measured as 183°C, the STE would be 0.015 kWh kg^{-1}.

The direct measurement of STE is not simple, due to convective and radiative losses to the environment and the heat inputs or losses from electrical coils or jackets. Moreover to estimate the heat transferred by direct steam injection or venting is not so easy.

PRACTICAL REMARKS

Estimation of the SME is usually accomplished by electrical measurement. The mechanical energy input is readily estimated for a direct current motor drive by [7]:

$$\mathrm{SME} = (PN\tau)\, t / N_m m, \tag{11}$$

where P is rated motor power (kW), N is motor speed (rots^{-1}), τ is % of torque, N_m is maximum motor speed (rots^{-1}), t is time (s), m is mass of the extrudate.

The actual value should be reduced by the power consumption of the extruder when it is running empty. The input power for an AC motor is given by [7]:

$$\mathrm{SME} = \eta P_r t / m, \tag{12}$$

where η is efficiency, P_r is watt meter reading (kW).

The efficiency of an AC motor is the strong function of the load. The value of the efficiency can be obtained only from motor curves (easy to obtain from the producer). In case of DC motors, the input power is a gross measurement and needs to be reduced by the power consumption of the empty extrusion-cooker.

Estimating thermal energy inputs can sometimes be difficult. There are a number of thermal energy sources and sinks. Steam injection (a source) and venting (a sink) are calculated by measuring how much water is added as steam or removed as water vapour. The heat added, or removed, by these actions is given by [7]:

$$Q = m_s c_v, \tag{13}$$

where Q is heat (J), m_s is mass of steam (kg), c_v is heat of vaporisation (J kg^{-1}).

The heat of vaporisation of water is obtained from steam tables and is a function of the steam injection pressure or the vent pressure. At 0.1 MPa pressure it has a value of 2.26 MJ kg^{-1} of steam. The water added can be measured with flow meters or can be estimated by a mass balance.

The other heat sources and sinks of thermal energy are through barrel heat transfer by using steam, water, or other heat transfer fluids or by electrical heaters. Moreover, significant heat losses occur via convection of heat from the barrel surfaces to the environment (important when processing at high temperatures).

The energy obtained from electrical heaters will be in the form of units of watts (J s^{-1}). If the jackets are heated with steam, the energy input is the same as that given above for steam injection, or venting, except where the mass of steam is the quantity being condensed in the condensate leaving the steam traps with a bucket.

If a thermal fluid, or water, is begin used, the thermal energy being transferred is obtained by a simple balance [7]:

$$E_t = W c_p \Delta T, \tag{14}$$

where E_t is the thermal energy (W), W is flow of thermal fluid (kg s^{-1}).

The flow rate of thermal fluid, or water can be measured by flow meters, c_p of the fluid is given in heat tables (4.184 kJ kg^{-1}K^{-1} for water), temperature change can be measured on thermal fluid.

Measuring the thermal losses that occur by convection of heat from the jacket to the surrounding air can be difficult. The losses mentioned can be measured directly with heat flux sensors [4]. A number of sensor simultaneously, must be used because the heat losses are different all over the extruder's surfaces. These sensors can correlate heat losses as a function of the extruder barrels' external surface temperatures, the environmental temperature, and location. The heat losses can be calculated as [7]:

$$h = \frac{l_h}{St\left(T_s - T_a\right)}, \tag{15}$$

where h is the heat transfer coefficient (Wm^{-1}K^{-1}), l_h is the heat loses (J), S is a barrel surface (m^2), T_s is a surface temperature (K), T_a is the air temperature (K).

The barrel surface temperature can be measured by the attachment of surface thermocouples at a number of places on the barrel surface. The heat transfer coefficient is a function of the position (top, bottom or sides) and the temperature difference between the extruder barrel's surface and the environment. Karwe and Godavri [4] provide a number of equations for heat transfer coefficients, h may be assumed to be about 10–15 W m^{-2} K^{-1}.

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