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Performance of a double drum dryer for producing pregelatinized maize starches

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Abstract

The response of an industrial scale double drum dryer to variation of steam pressure, drums rotation speed and level (height) of the gelatinization pool between the drums is presented. To our knowledge, this is the first time that the gelatinization pool level is treated as an input variable. The output variables are the product's moisture content, mass flow rate and specific load (equivalent to the product's film thickness). The effect of the drum surface temperature and width of the gap between the drums on the behavior of the output variables is examined. A theoretical analysis is presented for the qualitative assessment of the basic process variables that control the film thickness of the product. The role of the thermal inertia of the drum wall to the response of the dryer is discussed. Changes in the thermal efficiency of the dryer are inferred from overall heat transfer coefficients.

Keywords: Drum drying; Double drum dryer; Pregelatinized starch; Instant starches; Film thickness

1. Introduction

Drum-drying is one of the main commercial hydrothermal treatments of starchy food products, i.e., cereals, tuber and legume seeds, flours (Collona, Buleo, & Mercier, 1987). This type of drying is suitable for products which are viscous in their natural state or after concentration, such as mashed potato, precooked starchy baby foods, casein, milk, maltodextrins, fruit pulps, etc. (Falagas, 1985).

Drum-drying results in specific physicochemical modifications of the native starch due to the gelatinization and further solubilization of the starch granules (Mercier, 1987). The delivered products, often referred to as pregelatinized or instant starches, are customarily in the form of thin solid sheets. These starches are prepared in two consecutive stages: complete gelatinization (to improve the nutritional value of starch) and drying. Both stages exploit the heat transferred from the surface of the steam-heated drums to the wet product. Enhanced by the boiling-type of drying (Vasseur, Abchir, & Trystram, 1991a) the obtained drum-dried sheets present excellent wettability and are easy to rehydrate, qualities very important for ready-to-use products (Bonazzi et al., 1996).

Specifically, in a double drum dryer gelatinization takes place inside a "pool" of material formed between the drums by the use of two spring-loaded end plates bearing on the flat ends of the drums. The actual drying starts only after the gelatinized material leaves the pool and forms a thin film upon the surface of the drums. Film thickness control is a result of adjusting the gap between the two drums where a limitation is set by the necessity for preventing fall-through of feedstock at the gap. This feature dictates somewhat different operational characteristics by double-drum dryers than for single drum dryers, i.e., single drum dryers usually use higher drum rotating speeds (Gardner, 1971).

Gardner (1971) presented a comprehensive review of a broad range of drum dryer operational characteristics for industrial applications. Fritze (1972, 1973a,b) compared the performance of four different types of drum dryers – including a double drum dryer – for drying maize starch slurries (15–40% w/w solids). Drum speed

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and steam temperature were varied simultaneously to achieve a constant final moisture content in order to meet specific industrial demands. Heat transfer phenomena were analyzed in terms of overall heat transfer coefficients, an approach which assumes a stationary steady conduction inside the drum metallic wall. Kozempel, Sullivan, Craig, and Heiland (1986) developed a simple model to describe the drum drying of flakes of different potatoes' strains in a single drum dryer. Their model was based on the assumption of steady-state heat fluxes through the drum wall. Therefore, the temperature of the drum was not only constant but also equal to the temperature of the condensing steam. All the above studies neglected the thermal inertia of the drum wall and attributed the large decrease of drying rate during the rotation to either a variation of product thermal conductivity (Fritze, 1972; Gardner, 1971) or heat transfer coefficient on the steam side (Kozempel et al., 1986), which are arguments difficult to justify.

Daud and Armstrong (1987) studied the combined heat and mass transport phenomena involved in the drying of rice flour slurry (35% w/w solids) in a single drum dryer. Steam pressure and speed of rotation were their input variables. Assuming that moisture transport within the drying starch was driven mainly by a moisture gradient, they achieved a reasonable match between predictions and experiments as regards the moisture content of the starch film and the temperature profile of the drum surface. In the employed range of values, film thickness was not found to play a key role in the moisture content profile. In another paper, Daud (1991) studied the thermal dynamics of a single drum dryer in an effort to design a control system that could ensure the desired moisture content and temperature of the end product but also optimize the energy and material utilization. Daud (1991) and Daud and Armstrong (1987) relaxed the earlier approximation of steady-state heat fluxes by taking the thermal resistance of the drum wall into account and divided the dryer into three angular zones with different heat and mass transfer characteristics. Drying phenomena were described by a composite mechanism that resembles drying in capillary porous media where no appreciable shrinkage occurs. Although this approach may work for slurries of relativily low initial moisture content, it is questionable if it also works for more dilute suspensions.

A more systematic investigation of drum-drying processes was presented by Vasseur and co-workers (Abchir, Vasseur, & Trystram, 1988; Bonazzi et al., 1996; Rodriguez, Vasseur, & Courtois, 1996a,Rodriguez, Vasseur, & Courtois,1996b; Trystram, Meot, Vasseur, Abchir, & Couvrat-Desvergnes, 1988; Trystram & Vasseur, 1992; Vasseur, 1982; Vasseur & Loncin, 1983; Vasseur et al., 1991a; Vasseur, Kabbert, & Lebert, 1991b) who employed suspensions of quite high moisture content (~80% w/w). For the first time it was mentioned that drum drying occurs through boiling at least during its initial stages (down to about 50% of film moisture content) where, also, most of the total heat is exchanged. It was further assumed that the mass flux is dependent only on the heat flux received by the product and which, in turn, depends on the temperature difference between the boiling product and the wall. Allowance was made for an increase in heat flux when the product film shrinks at the beginning of drying, due to evaporation of the free water in the material. During the later stages of drying, when the product becomes dry enough, boiling ceases and the heat fluxes are allowed to decrease as the temperature of the product rises. A complete rotation of the drum was divided into four regions where different phenomena prevail. The model equations were solved numerically in a non-stationary but cyclic state. Yet, the case of a single drum drier was solely examined.

A survey of the recent literature shows that studies dealing with double drum dryers are rather scarce and are mainly of technological orientation. Kitson and MacGregor (1982) used a double drum drier of novel design to dry fruit purees. These products because of their relatively high sugar content, present a thermoplastic behavior when they are hot. So, their removal from the dryer required modifications of the exit section of the equipment. Rosenthal and Sgarbieri (1992) gave the response curves of a double drum dryer in terms of process yield, moisture content and nutritional properties of dehydrated sweet corn pulp as a function of steam pressure and speed of rotation. Their intention was to identify a range of operational parameters that could lead to good quality of the product. While the interrelationships among the input variables were apparent no effort was made to interpret that complex behavior.

Running a drum dryer is actually beset with many complexities so the control, regulation and optimization are subject to a good knowledge of the process (Rodriguez et al., 1996a,b). Hence, studying the performance of a small-scale industrial double drum dryer may be particularly beneficial in providing insight into the process. The purpose of the present work is to investigate the interactions between certain input variables (steam pressure, gelatinization pool level, rotation speed) and certain output variables (product mass flow rate, product moisture content and product specific load) for a pilot plant drum dryer. There is no published data treating the level of the pool as a control input variable. In order to obtain a preliminary estimation of the thermal performance of the drum dryer, overall heat transfer coefficients are calculated. Despite the limitations, this approach is still a common practice for industrial drum dryers (Fellows, 1988).

2. Materials and methods

2.1. Materials and equipment

Commercial native maize starch was purchased from GROUP AMYLUM S.A., Greece, with a moisture content of 13.5%. The mean diameter of the granules was 14.95 μm as determined by a laser diffraction particle size analyzer (Malvern Mastersizer, Malvern Instruments). The total amylose content was 26 ± 0.3%, determined by the method of Morrison and Laignelet (1983).

Native maize starch suspensions were modified in a double drum dryer (GOUDA; Fig. 1). The drums had a 0.5 m diameter and 0.5 m length and were synchronously rotated downwards towards the gap between the drums. The speed of rotation was adjusted by an electric motor with a mechanical built-in reducer. Starch/water suspensions with solids concentration of 10% w/w were employed as the dryer feed. The feed suspension was passed to the dryer from a large tank (150 l) by a rotary positive displacement pump. The suspension inside the tank was continuously agitated to ensure homogeneity. To prevent clogging of the pipes due to settling of solid particles, intense turbulent flow was maintained in all pipelines. A continuous uniform feed to the drums was achieved by letting the suspension pass through a perforated horizontal tube, extending over the length of the drums, and free fall into the space between the drums. There, the material created a pool, pre-heated and gelatinized. Due to pre-heating some preliminary water evaporation occurred there. Specific gravity and loss-in-weight tests showed that, for the range of our experimental parameters, moisture losses during gelatinization inside the pool were at most 3% of the total weight. A very small portion of the gelatinized slurry in the pool was continuously dragged by the rotating drums through the small gap between them and spread as a film over their surface. After traveling 3/5 of a revolution, the dried product was removed in the form of thin sheets by scraper ("doctor") knives extending the whole length of the drums.

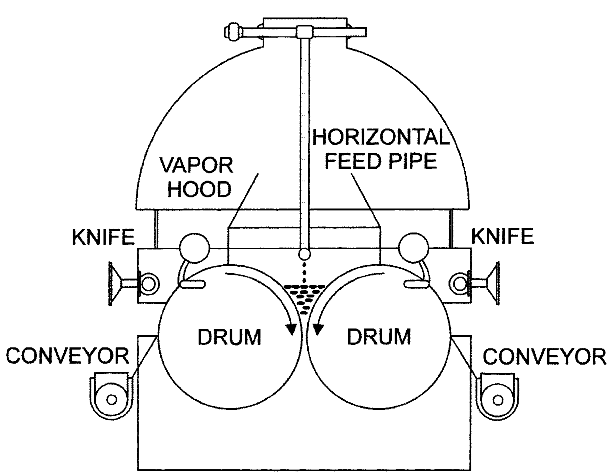

Fig. 1. Schematic representation of the employed double drum dryer (GOUDA).

The range of the rotation speeds – 2, 3, 4, 5 and 6 rpm – was selected so that the starch sheets neither become overdried and lifted off from the drums too early nor remained very humid and formed an amorphous cohesive mass at the doctor blades. The drums were internally heated by steam at 6, 7 or 8 bars. Smaller pressures could not sufficiently dry the product at the higher drum speeds whereas larger pressures caused excessive blistering of the starch sheets at the lower drum speeds. Before steam was supplied, the width of the gap between the cold drums was carefully adjusted to 0.90 ± 0.01 mm. This setting is based on recommendations of the manufacturer in order to achieve starch films that are not too thin and flimsy to be scraped away and also to increase the production rate. Of course, this gap setting is not maintained during the operation of the dryer because the drums expand when they heat up (see below). The free surface of the liquid pool between the two cylinders was kept at a level of 14, 18, or 22 cm above the gap.

Steam to heat the drums internally was coming from the building supply at a pressure of 10 bar. It was passed through a conditioning set-up involving a steam trap before it split in two lines, one for each drum. Two regulators with pressure gauges allowed adjustment of the line pressure to the desired value. The condensate from inside the drums was discharged to the sewage pipes through steam traps. Type K thermocouples, specially designed for rotating surfaces measured the temperature of the surface of the drum. Feeler gauges were used to measure the width of the gap between the drums. Mass flow rate was measured by collecting and timing large amounts of dry starch sheets as they came off the drums. Moisture content of the starch sheets was determined by drying samples to constant weight. A digital micrometer measured the thickness of the dry sheets with a resolution of 1 μm. The level of the slurry pool between the drums was monitored by an immersed graduated rod and was controlled by adjusting the feed flow rate.

2.2. Experimental procedure

The experiment started by rotating the drums at a required speed and then supplying steam to heatup the drums. The whole system was allowed to reach a thermal steady state (constant temperature at the surface of the drums). For all pressure levels, the time to reach thermal equilibration was independent of the rotation speed. Furthermore, the temperatures around the circumference of the drums were constant to within ±1°C but always less than the respective saturation values of the steam by 0.5–3°C. These observations combined indicate a satisfactory removal of condensate from the inside of the drums for all the rotation speeds examined.

The width of the gap was measured before the steam was supplied and also after the drums had warmed up to allow for expansion. Next, the starch/water suspension was introduced at a predetermined flow rate so as to maintain a constant pool level between the drums. Sufficient time was again provided to allow a new thermal steady state to develop. The steam pressure controllers always adjusted the steam flow rate to maintain a constant pressure level inside the drums despite the different thermal demands. Then, the product mass flow rate was measured and samples were taken to determine their moisture content and film thickness.

The experimental design included $3 \times 3 \times 5$ permutations for the studied levels of the input variables. Two independent series of experiments were conducted, each one for a different lot of native maize starch from the same provider. Each experiment included at least three measurements at every particular set of conditions. No significant differences were found between lots with respect to the parameters presented herein. The final selection of the data was based on a systematic scrutiny of experimental errors and further rejection of extreme outliers. The average variance V (= S.D./mean) calculated from all measurements was less than 0.1. The (x–x–x) format in the legends of the plots stands for (bars–cm (level)–rpm). Results from the two drums were close to each other so no differentiation is made between them but instead single mean values are presented.

3. Results and discussion

3.1. Material application on drum surface

There is enough evidence in the literature that the rotation speed and the width of the gap together with the rheological properties of the material control the deposition of the product layer (Fritze, 1973a; Trystram & Vasseur, 1992). This is easily realized since the gelatinized material leaving the pool occupies the entire space of the gap as it is dragged by the rotation of the drums in a *lubrication* type of flow (Papanastasiou, 1994). Just below the gap, the stream of the exiting material splits into two parts (layers). These layers have one side in contact with the metallic surface of the drum and the other side in contact with the ambient air. The flow of the layers now transforms to a so-called *stretching* type of flow (Papanastasiou, 1994). The transition between the two kinds of flows is accompanied by the formation of a meniscus at the split point. The stronger the adherence of the material to the drums the shorter distance below the gap where the meniscus appears. These adherence forces increase with product viscosity which, in turn, increases with solids content (Fritze, 1973a,b). For the employed gap widths ($\sim$0.3 mm), drum curvature ($R = 0.25$ m), drum speeds ($<$6 rpm) and viscosity values of the gelatinized product ($\sim$10,000 cp), order of magnitude analysis indicates that capillary and inertia phenomena have only a limited influence on meniscus formation and viscous forces prevail.

The present problem is customarily called roll (drum) coating and it is quite different from roll calendering, where the film is detached from the drums; a problem which has been studied extensively in the polymer literature (Alston & Astill, 1973). Daud (1986) studied a problem relevant to drum drying where the film is attached to one drum but detached from the other. This is actually a calendering problem and his methodology cannot be used in a double drum dryer.

The roll coating problem for Newtonian fluids and isothermal conditions has been solved in a fundamental fluid mechanics context both approximately, using asymptotic analysis (Pitts & Greiller, 1961), and numerically, using finite element analysis (Coyle, Macosko, & Scriven, 1986). The direct solution of the non-isothermal, non-Newtonian problem, as is the case in a double drum dryer, is extremely difficult to derive. However, the solution of the isothermal Newtonian case can be used for an (at least qualitative) assessment of the process. Pitts and Greiller (1961) used the lubrication approximation for the gap region, while for the meniscus region they assumed a parabolic shape for the meniscus and solved the Stokes equation using the separation of variables technique. The solutions for the two regions were matched appropriately to give the solution for the whole domain. The final result is a system of six algebraic and integral equations which implicitly relates the outflow from the system $2Q/L$ (Q is the volumetric flow rate through the gap and L is the length of the drums), with the gap thickness $2d$, drum radius R, viscosity μ, surface tension, σ and drum velocity U ($U = 2\pi Rf$, where f is the rotation frequency).

Performing an analysis similar to that of Pitts and Greiller (1961) but taking into account the finite depth of the pool as a boundary condition for the lubrication approximation, the following system of equations results:

$$3\kappa = 1 - \frac{1}{29.3A} - 2.44\kappa\beta + 3.28\alpha\beta(1 - 1.5\kappa), \tag{1a}$$

$$\alpha = \left(\frac{1}{56.4A} + 0.59\kappa\beta\right) \Big/ [(1 - \kappa + 3.2\beta(1 - 1.5\kappa)], \tag{1b}$$

$$-2\sigma_1 \int_{-h}^{(2\sigma_1-2\beta\sigma_1-2)^{0.5}} \frac{1+\xi^2/2-\lambda}{(1+\xi^2/2)^3}\mathrm{d}\xi = \left(\frac{2d}{AR}\right)^{0.5}\left[\frac{0.148}{A}+0.3\kappa\beta+1.65\alpha\beta(1-1.5\kappa)\right], \tag{1c}$$

$$\kappa = \lambda/\sigma_1, \tag{1d}$$

$$\beta = \left(\frac{2\alpha d(\sigma_1-1)}{R}\right)^{0.5}, \tag{1e}$$

where λ is the dimensionless outflow through the gap which is equal the ratio of the film thickness to the gap width, $(=Q/ULd)$, whereas h is the height of the liquid pool non-dimensionalized by $(Rd)^{0.5}$. The system (1a)–(1e) consists of five equations and must be solved for the five unknowns $\kappa, \lambda, \alpha, \beta$ and σ_1 in order to get their values as function of the system parameters d, R, A and h. The dimensionless parameter $A = \mu U/\sigma$ is the ratio of the viscosity versus the surface tension influence. For the present work the value of this parameter is typically higher than 10, which denotes that viscosity dominates the process. Also the parameter β in the present work is of the order of 10^{-2}. It can be shown that for $A > 10$ and $\beta < 0.05$ the above system (1a)–(1e) can be further simplified by taking the asymptotic result for large A and small β. So, by taking the limit $A \to \infty$ and $\beta \to 0$ and performing the integration in Eq. (1c) analytically, the system (1a)–(1e) can be replaced by the following single equation:

$$\lambda\left[F\left(2(3\lambda-1)^{0.5}\right)-F(-h)\right] - \lambda^2\left[G\left(2(3\lambda-1)^{0.5}\right)-G(-h)\right] = -0.32\left(\frac{d}{AR}\right)^{0.5}, \tag{2}$$

where

$$F(x) = \frac{x}{x^2+2} + 2^{-0.5}\tan^{-1}(x/2^{0.5}), \tag{3a}$$

$$G(x) = \frac{x}{(x^2+2)^2} + \frac{3}{4}F(x). \tag{3b}$$

Eq. (2) must be solved numerically to give λ as a function of A, d, R and h. The Newton–Raphson method was used with a starting value $\lambda = 4/3$ which corresponds to the case of drums being completely submerged in the fluid of the pool. For the experiments of the present work the right-hand side of (2) is smaller than 0.01 and also $10 < h < 30$. In this range of parameters the solution of Eq. (2) gives that λ is essentially constant, $\lambda = 1.3$ with a deviation of less than 0.2%. This, in turn, means that the film thickness is simply proportional to the width of the gap without any direct influence from the other parameters. The direct proportionality between the film thickness and the gap is expected to hold for every viscous fluid as long as the parameter A is large. This is the case for the dilute starch slurries used in this work which despite their viscoelastic character (e.g., Evans & Haisman, 1979) exhibit mainly a viscous behavior. It is worth noticing that the end result of the present analysis is not dissimilar to the work done by Daud (1986) for single drum dryers since in both cases the dimensionless film thickness is constant for a practical range of pool levels and depends only on the width of the gap. This is interesting if one notes that in single drum dryers the height of the pool is customary of the order of an inch, e.g. Daud and Armstrong (1987).

The above results are very significant since they show that the solution of the complicated hydrodynamic and heat transfer problem in the pool and between the drums is not necessary for the prediction of the film thickness. The only important information is the shape of the gap between the drums or, alternatively, the width of the gap if it is assumed that the drums retain the cylindrical shape during their thermal expansion. To find the width of the gap the temperature distribution on the drums must be known. Here is where the pool level acts on film thickness, that is, by reducing the drum temperature and not directly on the hydrodynamics.

For dense feed suspensions most of the boiling takes place in the pool and at the moment the material passes through the gap it is already dry enough with most of its moisture captured in the gelatinized starch matrix (Daud & Armstrong, 1987). However, for the present dilute suspensions, temperatures close to 100°C were often measured on the surface of the film, especially with speeds of 5 and 6 rpm. Therefore, in some experiments of this study boiling may continue also after the gap. Evidently, if this is happening the above theoretical analysis is not longer valid since there is not anymore a pure viscous fluid in contact with the drums. Also, the material pick-up by the drums may be largely influenced by the boiling on the surface of the drums and not by the speed of the drums and this is probably the reason for the leveling-off of film thickness at higher rotation speeds (see Fig. 6(c)).

As soon as the material splits into the two layers running over the drums, two phenomena may occur. Stretching which tends to diminish film thickness and boiling which has a dual counteracting effect: it shrinks the product sheet because of moisture losses but also it inflates it due to a considerable inclusion of gas bubbles (porosity). Stretching is governed by the shear stresses created by the motion of the drums as well as the rheology of the material. For high consistency viscoelastic materials like the present one (Evans & Haisman, 1979; Karapantsios, Sakonidou, & Raphaelides, 2000; Wong & Lelievre, 1981), shrinkage is dictated not only by moisture removal but also by the material's rheology which in the present case is not even stable but changes dramatically with solids content upon drying. Inflation is controlled by the intensity of boiling, driven by the temperature difference between the drums and the product (Bonazzi et al., 1996; Fritze, 1973a), and, once more, by the rheology of the product. The interplay among these factors determines the instantaneous film thickness around the drum circumference. In the present study, the feed slurry is very dilute and, therefore, film shrinkage may prevail over the entire drying cycle giving an end-product with a small thickness, <0.1 mm.

3.2. Temperature of drums surface

During the operation of the dryer – with material covering the drums surface – the temperature is not the same at every angle around the drums but exhibits large deviations, (Daud & Armstrong, 1987; Vasseur et al., 1991a). These deviations vary a lot with operating conditions and are not readily predictable (Trystram & Vasseur, 1992). Fig. 2 shows the effect of the three input variables (steam pressure, gelatinization pool level, rotation speed) on a characteristic local surface temperature of the drums; readings are taken on the bare metallic surface of the drums just after the doctor blades position. These values are lower than the values of the respective saturated steam by as much as 30°C, depending strongly on the other operating conditions. According to Daud and Armstrong (1987) and Vasseur et al. (1991a), the temperature in this section of the drums has the tendency to climb fast because there is no material on it. Temperature data recorded in this section are very stable with respect to both time and drum length. Measurements on other locations around the circumference where material is present – not shown here – exhibit large fluctuations brought about by the random occurrence of wetter zones over the drum surface. These zones have lower temperatures. Rodriguez et al. (1996a) reported fluctuations higher than 10°C with respect to time and higher than 20°C with respect to the drum length, which were caused by such wet zones. Rodriguez et al. (1996a) and Trystram and Vasseur (1992) attributed the existence of wet zones to possible uneven initial deposition of the material onto the drum surface, a fact which was considered responsible for a subsequent irregular drying.

As can be seen in Fig. 2(a), higher steam pressures heat up more the drums. The effect of the pool level becomes more evident at high rotation speeds, Fig. 2(b). The higher the level the lower the temperature. There is experimental evidence (Abchir et al., 1988) that the internal temperature of the drums changes very little along the rotation as long as a constant steam pressure is maintained which, in turn, means that the internal heat flux (steam → drum) is pretty constant. On the other hand, the external temperature of the drums was measured to fall quickly at the start of the cycle (when the drum is in contact with the pool material) and then,

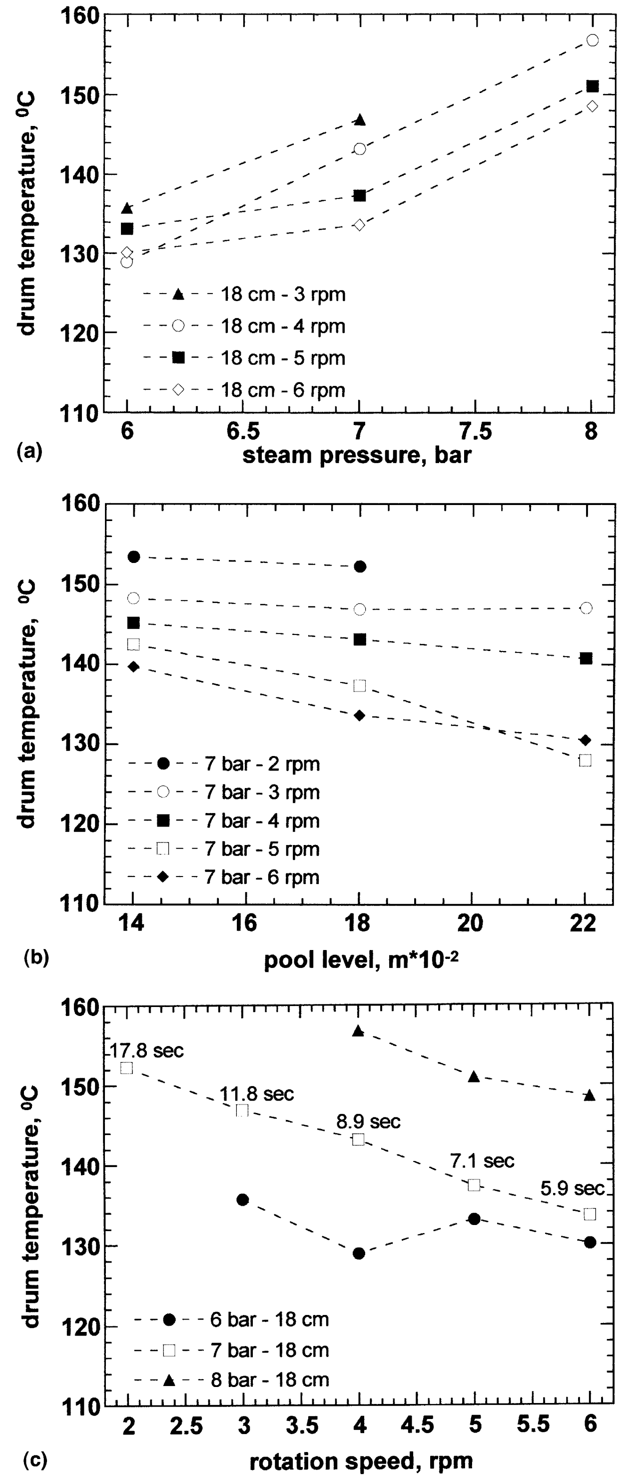

Fig. 2. The effect of (a) steam pressure, (b) pool level between the drums and (c) drums rotation speed on the drums temperature measured at the bare surface just after the position of the scraper blades. In graph (c) the respective times from the gap to the scraper blades are also presented.

during the drying of the thin film, to rise slowly (Abchir et al., 1988). The heat flux through the drum metallic wall varies accordingly. Hence, when more material accumulates in the pool the boiling load increases and this causes the temperature of this section of the drums to decrease in order to favor a higher heat flux from the wall. The temperature of the drums eventually stabilizes at a value where the boiling load is compensated by the increased heat flux from the wall. As soon as the drum surface departs from the section where boiling takes place then heat is accumulated again in the metallic wall of the drum, which now acts as a heat reservoir.

When the rotation speed increases, the surface temperature decreases, Fig. 2(c). One might think that this is only due to the increase in product throughput rate (see below) which absorbs now more heat from the drums to remove the bigger moisture load. A better explanation of the behavior in Fig. 2(c) may also include the reduction in both the film thickness (see below) and the drying period as the rotation speed increases. The drying times (from the gap to the doctor blades) are estimated as 17.8, 11.8, 8.9, 7.1 and 5.9 seconds, for 2, 3, 4, 5 and 6 rpm, respectively. Since the external heat flux is higher in thinner films and the time for drying is shorter, the drums must lower their temperature to accommodate the bigger thermal demands. This is also what other researchers observed for single drum dryers, e.g., Daud and Armstrong (1987); Trystram and Vasseur (1992); Vasseur et al. (1991a). As a further support to this notion, it must be added that for speeds 2, 3 and 4 rpm the temperatures measured on the material surface just before the product sheet is scraped off – not shown here – are at most 5°C less than the corresponding bare surface values. On the other hand, for 5 and 6 rpm the material temperatures are much lower, in the range 101–109°C.

3.3. Width of gap between drums

During the operation of the dryer the gap attains a width value different from the initial (cold) setting since the drums expand or contract in compliance with their thermal condition. Fig. 3 displays gap widths measured over a realistic range of drum surface temperatures after thermal equilibration is reached and with no material on the dryer. Gap width decreases with temperature in a smooth but no linear fashion. In a drying experiment, where the temperature is no longer uniform around the drums because of the presence of material, it is the overall mean temperature of the drums that dictates the eventual gap width. This quantity cannot be measured easily due to interference with the continuously flowing product. So, at present, only a rough estimation can be made of the actual operational gap width. If one combines Figs. 2 and 3 it seems that the gap gets narrower when the steam pressure increases and the pool level and drum speed decreases. Apparently, the operational gap has to be greater than the value corresponding to the local temperature of the bare drum surface in Fig. 2 and much less than the initial gap setting.

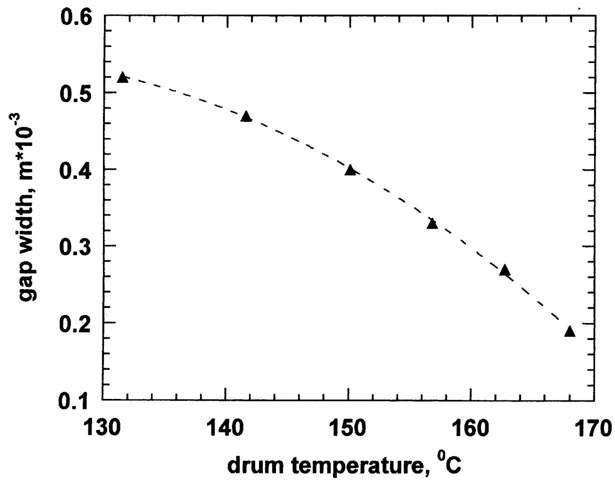

Fig. 3. Variation of gap width versus surface temperature of the drums. Measurements are conducted with no material over the drums. The initial gap setting is 0.9 mm at 15°C.

3.4. Moisture content of dry product

The determination of the end-product moisture content is hindered by non-uniformities caused by wetter zones across the product sheets. To reduce fluctuations in the final moisture values copious quantities of the product sheets are mixed together before moisture determination. Fig. 4 displays the effect of the three input variables on the moisture content (wet basis) of the produced dry sheets. Increasing the steam pressure usually makes the moisture decrease, Fig. 4(a). This is rather expected due to the higher drum temperatures (Fig. 2(a)). Similar observations were made in all previous studies for either single or double drum dryers (i.e., Daud & Armstrong, 1987; Fritze, 1972; Rosenthal & Sgarbieri, 1992; Trystram et al., 1988; Trystram & Vasseur, 1992; Vasseur et al., 1991a). Interestingly though, for an 8 bar steam pressure the final moisture is approximately constant for all rotation speeds and this holds also for the other pool levels (not shown here). Apparently, the temperature attained by the drums at this pressure and the heat flux they supply are high enough to dry the material completely regardless of the other conditions. For a rotation speed of 6 rpm the product moistures for 6 and 7 bars are comparable. The large throughput rate and the short drying time in this occasion are rather responsible for this. On the other hand, when the moisture is already small (~5% for 3 rpm at 6 bars) and mostly bound water is present in the starch sheets, any further increase in steam pressure does not have a marked effect. A decreasing trend of moisture as steam pressure increases in the range 3.5–5 bars, is reported by Rosenthal and Sgarbieri (1992) for speeds between 1.5 and 3 rpm. What is also inferred from their data is a gradual leveling-off of moisture towards higher pressures, which is in line with the present results for 8 bars.

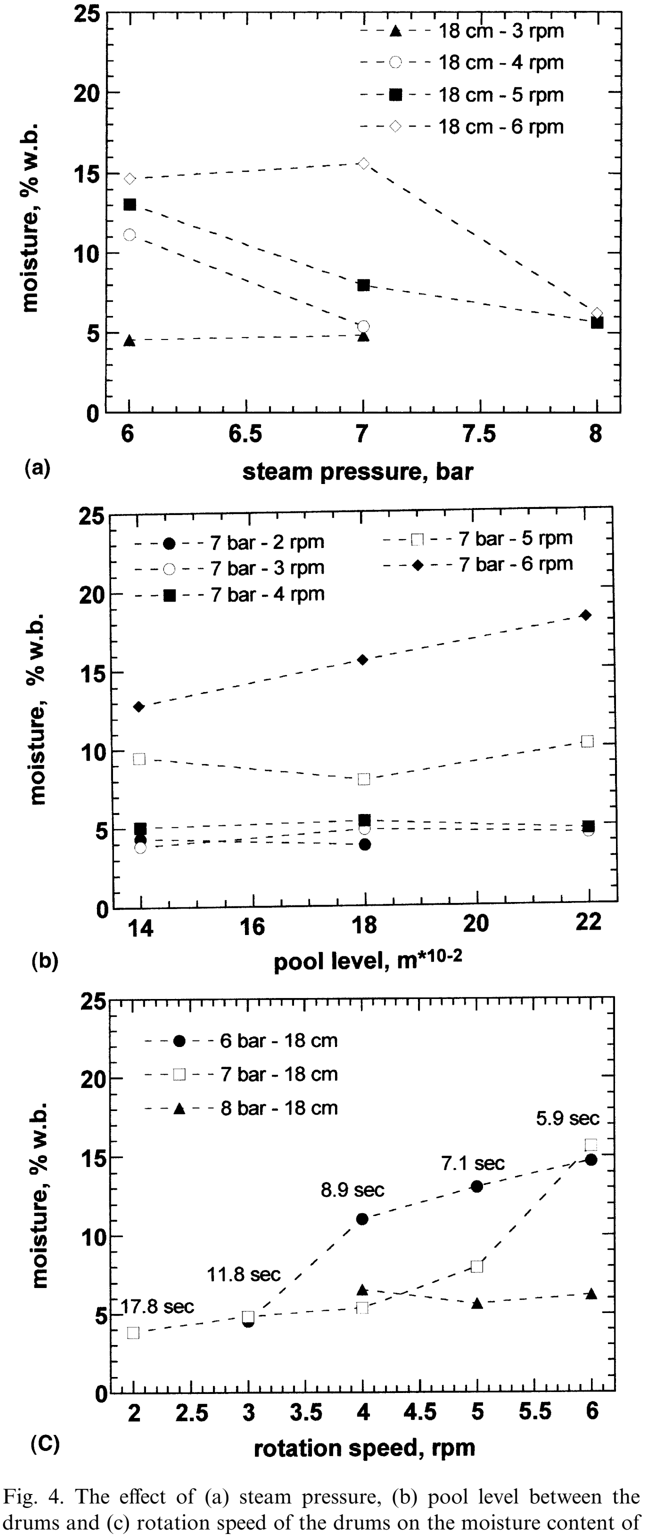

Fig. 4. The effect of (a) steam pressure, (b) pool level between the drums and (c) rotation speed of the drums on the moisture content of the end product. In graph (c) the respective times from the gap to the scraper blades (drying times) are also presented.

Product final moisture is hardly affected by the level of the pool, as shown in Fig. 4(b) for a steam pressure of 7 bars. This conforms to the picture described in Fig. 2(b) regarding the surface temperature of the drums. Only for 6 rpm a variation is observed, most likely because then the drum temperature gets considerably lower as level increases. For a steam pressure of 6 bars this variation persists down to 4 rpm whereas at 8 bars no significant variation is found for all speeds. Rosenthal and Sgarbieri (1992), used only a fixed pool level of 10 cm ($D = 45.7$ cm; gap width $= 0.15$ mm) so no direct comparison can be made with the only other available detailed study of a double dryer.

Fig. 4(c) presents the variation of moisture with rotation speed at a pool level of 18 cm. As already discussed, no significant variation is measured at a steam pressure of 8 bars. For 6 and 7 bars, it appears that the product moisture increases continuously with speed. However, the shapes of the curves for 6 and 7 bars are distinctly different (also valid for 14 and 22 cm pool levels). For a 6 bar pressure, moisture loss is larger at lower speeds whereas the reverse is true for 7 bars. Rosenthal and Sgarbieri (1992) working between 1.5 and 3 rpm found an almost linear increase in final moisture with speed in the range 3.5–5 bars.

The role of the rotation speed on the final moisture content is quite intriguing. When speed increases both drying time and drum surface temperature decreases (Fig. 2(c)). So, for a shorter period of drying and a lower surface temperature a moister product would be expected. However, when speed increases, the film thickness over the drums decreases (see below) resulting in much higher heat fluxes and, therefore, higher evaporation losses. Yet, the latter phenomenon is prominent only at the start of drying when the material has still appreciable moisture and fluidity (Vasseur, 1982; Vasseur & Loncin, 1983). As the film dries beyond the point where free water is present, the product layer heats up since its boiling temperature goes over 100°C and, consequently, the heat (and mass) flux decreases (Daud & Armstrong, 1987; Vasseur et al., 1991a,b). On this account, the literature offers two different explanations. Vasseur et al. (1991a,b) claim that heat flux decreases because the temperature difference between the drum surface and the product layer gradually diminishes so the driving force for drying gets smaller. On the other hand, Daud and Armstrong (1988), Vasseur (1982) and Vasseur and Loncin (1983) consider that any further moisture removal occurs mainly by diffusion which means that the heat transfer coefficient becomes smaller. This last argument is in line with earlier phenomenological studies (Fritze, 1972, 1973a; Gardner, 1971). The present work cannot offer direct evidence in this respect. Nevertheless, whatever the dominant mechanism, it appears that as speed increases the cumulative heat flow over the entire drying cycle gets less, producing, thus, more humid products, Fig. 4(c).

In order to elucidate the complex interaction among input variables on the product final moisture for all the examined conditions, a multiple linear correlation was performed. The statistically most significant fit was given by the following equation with a multiple correlation coefficient of $R = 0.82$:

$$M = 16.98(\pm 2.3) + 2.50(\pm 0.21) * V - 3.28(\pm 0.38) * P, \quad (4)$$

where M is the moisture content, (% wet basis), V is the speed of rotation (rpm), and P is the steam pressure (bar). The quantities in the parentheses are the standard deviations of the coefficients. It must be noted that the effect of the pool level was shown to be overall statistically insignificant ($p > 0.95$). According to the partial correlation coefficients and the c_p criterion (Petridis, 1997), the speed of rotation is more significant than steam pressure to the outcome moisture. The above correlation is strictly applicable to the specific drum dryer of the present work and cannot be used in other designs. The generalization or scale-up of the problem requires a better understanding of the process and development of an appropriate mathematical model, which would allow the knowledge gained in one system to be transferred to another system.

Recently, Rodriguez et al. (1996a) developed a technique which allowed the evaluation of the final moisture of the dry product with good precision by measuring its temperature at the scraper blade. The relationship between the temperature and the moisture content was given by the characteristic *equilibrium* desorption isobar of the product (Bassal & Vasseur, 1992; Bassal, Vasseur, & Loncin, 1993). This technique was next employed to the automatic control of a drum dryer (Rodriguez et al., 1996b). A similar effort was also undertaken in the present study. However, the temperatures measured on the material surface were found to fluctuate so intensively that they could not be used for automatic control purposes. Besides, although the product's temperature can be uniquely linked to its moisture content through an equilibrium desorption isobar, it is questionable whether such an equilibrium is ever reached in the short times encountered in drum drying.

3.5. Mass flow rate of dry product

Fig. 5 shows the effect of the three input variables on the exit flow rate of the mass removed by the scraper knifes. The respective graphs of the exit flow rate of the dry mass (not shown here) present the same qualitative behavior. An increase in steam pressure makes the exit flow rate decrease, Fig. 5(a). This is opposite to what past literature reports about single drum dryers with spreading rollers (Gardner, 1971; Kozempel et al., 1986). On the other hand, Rosenthal and Sgarbieri (1992) in their double drum dryer observed the exit flow rate to decrease first and then increase with steam pressure in the range 3.5–5 bars.

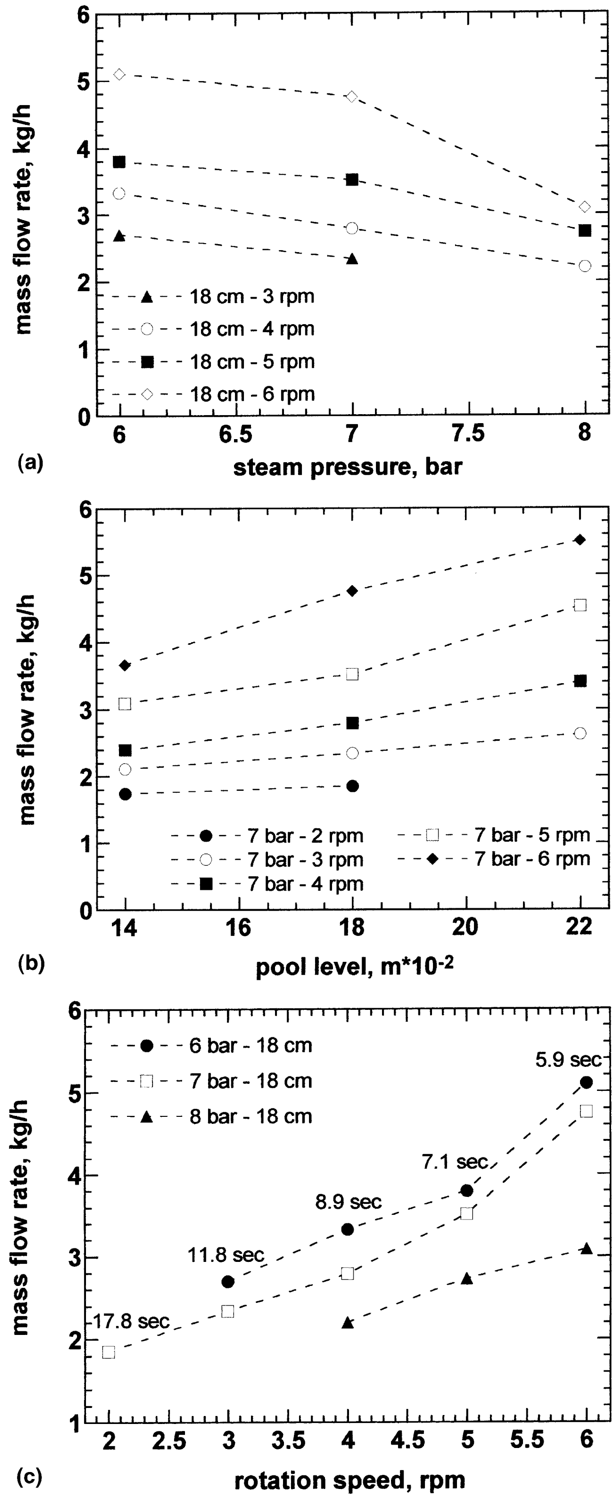

Fig. 5. The effect of (a) steam pressure, (b) pool level between the drums and (c) rotation speed of the drums on the mass flow rate (wet basis) of the end product. In graph (c) the respective times from the gap to the scraper blades (drying times) are also presented.

A higher steam pressure leads to a higher drums temperature (Fig. 2(a)) which apparently narrows the gap between the drums, so less material can go through. In addition, a higher drum temperature means a higher moisture removal and consequently a lower flow rate of end product. However, the graphs for the exit flow rate of the dry mass (mentioned above), show also a significant decreasing trend with pressure, so the impact of the moisture removal must be less compared with the effect of gap narrowing. In addition, when the temperature increases the viscosity of the material inside the pool decreases and it is now easier to be sheared off by the motion of the drum thus increasing the flow rate. Such opposing phenomena might possibly explain the non-continuous variation observed by Rosenthal and Sgarbieri (1992). In our case, it seems that the narrowing of the gap prevails.

An increase in the level of the pool results in larger flow rates, Fig. 5(b). This becomes progressively more evident as the rotation speed increases. Such a behavior may not be ascribed only to the small change of the drums temperature, shown in Fig. 2(b), which affects the gap width. Another concomitant effect involves the longer residence times of the feed slurry in the pool when the pool level increases. It is then possible that the more severe thermal treatment of the material in the pool due to the prolonged preheating can give lower viscosities and, therefore, easier pick-up by the drums.

The rotating speed of the drums influences the exit mass flow rate (Fig. 5(c)), apparently, by dragging the material through the gap. Earlier studies claimed that this is true only with drum speeds over the lower ranges (e.g., Fritze, 1973a; Gardner, 1971). Trystram and Vasseur (1992), though, predicted for a single drum dryer that the product flow rate would steadily rise with rotation speed despite the reduction in film thickness. Rosenthal and Sgarbieri (1992) also observed a positive dependence of product yield on drum speed in their double drum dryer. However, unlike in other studies, there was a tendency to gradually reach a constant value (plateau) at speeds a little higher than those employed (>3 rpm). Of course, when the rotation speed goes up, the gap width also goes up (Fig. 2(c)) and the part played by this factor must not be overlooked.

The most significant correlation equation for the exit mass flow rate was

$$m = 1.44(\pm 0.43) + 0.60(\pm 0.04) * V - 0.55(\pm 0.06) * P + 0.14(\pm 0.01) * H \qquad (5)$$

with a multiple correlation coefficient $R = 0.93$. The symbols are the same as in Eq. (4) with the addition of m for the exit mass flow rate, (kg/h), and H for the pool level (cm). Seemingly, the significance of this correlation is higher than for moisture in Eq. (4). According to the partial correlation coefficients and the c_p criterion (Petridis, 1997), the order of significance among the input variables for the determination of the flow rate is: speed of rotation, pool level and steam pressure. Thus, pool level contributes more than steam pressure to the dryer's production. Finally, it must be added that the mass flow rate data present a good correlation with the moisture data, $R = 0.82$.

3.6. Specific load of dry product

The variation of film thickness with respect to operating conditions was overlooked by many earlier studies as regards the performance of the drying process. Vasseur et al. (1991a,b) were the first to recognize its importance and investigated the influence of the so-called specific load (kg dry mass/m^2) on the efficiency of the dryer. This variable does not change during the drying cycle and was found to be strongly associated with the drying kinetics of the product layer (Vasseur et al., 1991b). Inasmuch as this variable describes the quantity of material applied over the area of the drum, it was further considered as equivalent to the film thickness. Strictly speaking this is not valid since the bulk density of the product layer may vary considerably during drying according to the severity and duration of the treatment (Fritze, 1973a).

Figs. 6(a) and (b) display the variation of the specific load versus steam pressure and pool level, respectively. It appears that the specific load had the same trend as the exit mass flow rate presented in Fig. 5(a). Arguments advanced there may hold here, too. A good correspondence with the exit mass flow rate is found also as regards the dependence on pool level, Fig. 5(b). What is perhaps of greater significance is the decay pattern with respect to the rotation speed in the curves for 6 and 7 bars, Fig. 6(c). For the sake of comparison, the micrometrically determined film thickness of the exit dry product for 7 bars is also included in Fig. 6(c); error bars represent the fluctuation in measurements, $\sim \pm 5\%$. Apparently, there is a considerable qualitative correspondence between specific load and actual film thickness. Contrary to what is observed for the exit mass flow rate for 6 and 7 bars (Fig. 5(c)), the thickness of the film gradually decreases when speed increases and ultimately reaches a constant value (plateau); for 8 bars the film thickness is pretty constant throughout. At first sight this behavior may look contradictory. However, it is not so if one realizes that the exit mass flow rate reflects the product of the specific load multiplied by the speed of rotation, as also noted elsewhere (Trystram & Vasseur, 1992). The thinner films applied over the drums at the higher speeds are responsible for enhanced heat transfer and higher drying rates. However, the cumulative moisture removal from the product upon drying is dictated by the total mass flow rate and that's why moisture content eventually increases at higher speeds. Having in mind that the specific load is directly related to the initial film thickness since the water content in the pool is more or less constant (Section 3.1), another useful observation can be made: the variation of the specific load among

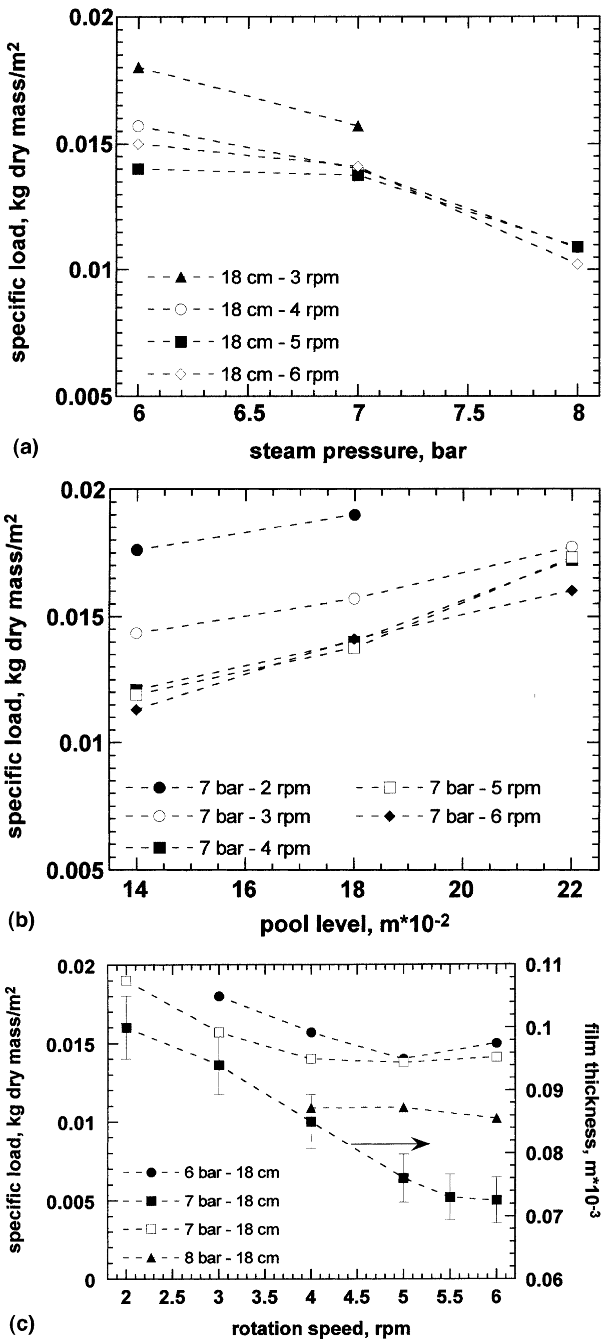

Fig. 6. The effect of (a) steam pressure, (b) pool level between the drums and (c) rotation speed of the drums on the specific load (dry basis) of the end product. In graph (c) the respective times from the gap to the scraper blades (drying times) are also presented.

the several experiments in Fig. 6 is of the same order with the variation of the gap width in Fig. 3.

3.7. Thermal efficiency of drum dryer

When it comes to an industrial situation the operating cost of a drum dryer is of high significance. Therefore, it is a top priority not only to achieve a good quality product but also run the dryer with a high energy exploitation. A correct definition of thermal efficiency would be to take the ratio of the heat used in removing the material moisture over the total heat supplied by the steam. However, a gross estimate of this, can also be made by calculating the overall heat transfer coefficients. This approach assumes that stationary heat transfer takes place around the drums which is by far not the case. Interestingly, it found a broad utilization in the past (Fritze, 1972, 1973a; Gardner, 1971) and it is still in use for industrial applications (Fellows, 1988). Moreover, the phenomena occurring in drum drying are very complex and as yet a simple comprehensive model suitable for industrial applications is not available (Abchir et al., 1988; Trystram & Vasseur, 1992). Thus, despite the shortcomings, it was decided to calculate overall heat transfer coefficients at the various experimental conditions for the sake of comparing them to each other and also to other published values.

Average around-the-drums coefficients obtained from our data lie in the range 300–1000 W/m^{2}C. These coefficients show a dependence on all input variables (graphs not shown). In particular, higher heat transfer coefficients are calculated at low steam pressures, high pool levels and high rotation speeds. Fritze (1973a) communicated that under optimum conditions the coefficient may vary approximately from 7000 W/m^{2}C at the gap to 600 W/m^{2}C in the area next to the knifes. Gardner (1971) for the same angular locations around the drums gave the values 2050 W/m^{2}C and 1250 W/m^{2}C, respectively. McCabe and Smith (1971) reported that under adverse conditions the average around-the-drums coefficient can be as low as 170 W/m^{2}C.

4. Conclusions

A complex interaction was found between steam pressure, drums rotation speed and pregelatinized pool level during the operation of a double drum dryer. Evidence was provided that all input variables seriously affect the drum temperature and, subsequently, the width of the gap between the drums. When the steam pressure increased, moisture content, mass flow rate and specific load of the dry product decreased. When the drum speed increased, the response of the dryer was different for the different steam pressures employed. Thus, at 6 and 7 bars the product moisture content and mass flow rate went up whereas the specific load went down. Yet, for the highest rotation speeds (5 and 6 rpm), the specific load gradually leveled-off to a rather constant value. On the contrary, at 8 bars the response variables were virtually invariant for all the employed operating conditions.

For the first time, the role played by the gelatinization pool in a double drum dryer was explicitly investigated.

When the gelatinization pool level increased, mass flow rate and specific load of the product increased, too. However, as regards the moisture content only a small increasing effect was observed at the highest rotation speed (6 rpm). Among all parameters, the one with the most profound effect on the overall performance of the dryer was the rotation speed of the drums. To this end, if the objective were to obtain pregelatinized starch sheets with just 5% moisture at the higher possible throughput rate and the more efficient energy utilization, then the present drum dryer should be operated at 7 bar steam pressure, 22 cm pool level and 4 rpm rotation speed.

A theoretical analysis was developed to qualitatively assess the influence of the basic process parameters on the product's film thickness. It was shown that for the present case the thickness of the material spread on the drums was practically determined only by the width of the gap. The experimental observations that the increase of pool level leads to a decrease in drum temperature, increased gap width and increased of specific load were in line with the theoretical predictions. Despite the reservations for the suitability of the overall heat transfer coefficients to describe the phenomena occurring in a drum dryer, these coefficients can serve as preliminary indicators among different operating conditions regarding energy utilization. In the present study, the calculated coefficients were in a range of values similar to those reported in literature for efficient operation of the dryer.

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