



On modelling and control of a rotary sugar dryer

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Abstract

This paper deals with the problem of fitting a set of data collected on a rotary sugar dryer, by means of a first-principles mathematical model. Owing to the highly constrained structure of the model, it was discovered that the sugar dryer is characterised by two different working conditions: the “standard-mode” (characterised by a non-zero sugar moisture content), and the “overdried-mode” (namely a condition where the sugar moisture content is almost nil, and the evaporative phenomenon becomes negligible). On the basis of this two-stage behaviour, an accurate fit between the model and the measured data can be achieved, and an innovative control strategy can be drawn. © 2001 Elsevier Science Ltd. All rights reserved.

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1. Introduction

The aim of this paper is to analyse and discuss the problem of fitting a set of real data collected on a rotary sugar dryer (at the *CSR Ltd.* plant of *Plane Creek — Queensland, Australia*) with a first-principles model, and to propose an innovative regulation scheme, based on a posteriori model analysis.

The problem of regulating the characteristics (temperature and moisture content) of the sugar collected at the dryer outlet is a standard problem in sugar plants. Specifically, the goal of a sugar dryer (which is the last stage of the plant) is two-fold:

- *cooling the sugar* to a temperature of about 30–31°C or lower, in order to avoid the discolouring phenomena, which inherently reduces the commercial value of the sugar;
- *regulating the moisture content of the sugar*, in order to avoid sticking (caused by high-moisture content) and the dispersion of sugar dust in the environment (caused by low-moisture content). Sugar moisture content also has an immediate impact on price, with

excessive moisture attracting a penalty. The target moisture content is typically about 0.2%.

The regulation of the dryer is a crucial issue in a sugar plant; however, this problem is usually dealt with without resorting to mathematical modelling tools and automatic control: plant personnel manually change the control variables, according to accumulated experience of the plant behaviour. However, due to the comparatively large number of input/output variables and the complicated relationships among them, it is expected that replacing human operators with automatic controllers could remarkably improve the quality and the evenness of the output sugar characteristics.

It is interesting to point out that, at a first glance, the design of a feedback controller for the regulation of output sugar moisture and temperature seems to be rather simple and straightforward. As a matter of fact, the required bandwidth of the regulation loop is very small (if compared to the main time constants of the plant dynamics) since only the average value of the output variables really matters; moreover, many input variables are available for control purposes. However, unfortunately, additional constraints make this control problem difficult and tricky, namely:

- the on-line measurement of input/output variables is very difficult; this is especially true for moisture measurements;

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- even collecting an off-line set of input/output data of a plant is difficult and takes a lot of time: this makes the validation of physical models rather difficult, and black-box modelling hardly or not practicable;
- the plant is affected by large and unpredictable disturbances; therefore, an automatic control strategy must be simple and robust;
- the environmental conditions are extremely hard due to humidity, dust, and vibrations; as a consequence, only simple sensors and electronic equipment can be employed;

In the literature, the problem of controlling numerous types of dryers (rotary, solar, cross-flow, mixed-flow, fluidised or fixed bed, drum, etc.) used for the desiccation of many different materials (fertilisers, pharmaceuticals, mineral concentrates, agricultural products, cereals, cement, etc.) has been treated extensively, both from a theoretical and from a practical point of view (see Section 6 at the end of this paper for a brief overview of the literature on the topic). However, it is worth noticing that (to the best of the authors' knowledge) very few works have been published on the problem of modelling and control of rotary sugar dryers.

The purpose of building a dynamical model of a rotary sugar dryer is primarily for conducting "what if" simulation experiments. These must be sufficiently faithful to the real process to yield quantitative guidelines for confident decision-making on capital expenditure. Additionally, operating guidelines and feedback control principles can be derived from the model.

A relevant feature of the modelling proposed here is that it takes place using sampled data from an operating production dryer. This process operates in continuous mode, rather than in batch, with a counter-flow of hot/moist sugar and cool/dry air. The data collected on the dryer are of two types: highly aliased "full-state" samples of sugar temperature and moisture content from within the dryer at Macknade mill; corroborated and extended by more frequently sampled input–output data from the Plane Creek mill (the plant this paper mainly focuses on). The quantity of data is low, and the measurement of each sample's moisture content requires laboratory analysis. The variability of the data also is large.

Given the model's purpose, and the nature of the data, a discrete-time model was constructed that focuses on the determination of the dominant physical processes at play and the coarse quantification of their effects. What distinguishes this study from earlier ones is the material being dried, which does not exhibit a significant transition between drying phases, the continuous mode of operation using an industrial plant in use, and the coupling between model properties and its ultimate purpose.

Among the few recent papers on rotary sugar dryers, the most closely related to the present work is Douglas, Kwade, Lee, and Mallick (1993), which is mainly devoted

to the development of a first-principles dynamical model of a rotary sugar dryer. A sequence of data collected on two Australian sugar plants has been used for validation.

The present work differs from Douglas et al. (1993) in that a purpose-directed model is fitted by determining the dominant physical processes capable of describing the data rather than developing a complete first-principles model. Like Douglas et al. (1993), the purpose is to provide a useful simulation system to aid capital decisions; here the use of the model is also extended to control strategy development. The models are constructed in discrete time and use internal state measurements in addition to input–output data to discriminate between phenomena and to fit parameters. Thus, this paper can be regarded as extending Douglas et al. (1993). The main contribution and novelty of this work can be summarised as follows:

- The first-principles model proposed in Douglas et al. (1993) has been slightly modified and refined. Specifically, in the model presented herein the time and spatial discretisation is optimised to deal with the two (clearly separated) dynamics of vapour/air and sugar/moisture velocities.
- The analysis of the model has been extended to low-moisture ranges. This analysis reveals that the model cannot be used for an accurate prediction of moisture content, when the sugar moisture is very low (below 0.1%); however, it can be effectively used to predict the sugar temperature, and the working region of the sugar dryer.
- The data paucity (in terms of number of available data and signal-to-noise ratio) and the use of only input/output data means that "grey-box" ideas have had to play a central function. Therefore, the validation-phase of a fixed first-principles model has been replaced by the selection of a small number of free parameters, and their subsequent tuning using a set of measured data.
- Using the grey-box model, it has been shown that the sugar dryer is characterised by two main working conditions, according to the sugar moisture content: the "standard-mode" (characterised by a non-zero sugar moisture content), and the "overdried-mode" (namely a condition where the sugar moisture content is almost nil, and the evaporative phenomenon becomes negligible). Even if this classification is simple, it captures the main non-linearity of the system.
- The estimated grey-box model and the "two-stage" behaviour of the dryer has been used to generate operating guidelines akin to feedback control solutions, making possible a considerable improvement in the plant control strategy.

The outline of this paper is the following: in Section 2 the sugar dryer is briefly described, and its

mathematical model is given, while in Section 3 the set of data collected on the *Plane Creek* dryer is presented and discussed. Section 4 is entirely devoted to the problem of fitting the model with the measured data; in Section 5 the problem of controlling the sugar dryer is considered, and an innovative control strategy is proposed. Finally, in Section 6 the literature related to this work is briefly outlined.

2. A rotary sugar dryer: model and description

The *CSR Ltd* Plane Creek sugar mill uses rotary drum dryers as the last stage in the sugar milling process. The dryer operates in continuous mode. A basic diagram of such a dryer is in Fig. 1.

It consists of a large drum (about 9 m long and with an internal diameter of about 2.5 m), set at a slight angle (about 3°) in order to let the sugar move slowly under gravity from the inlet to the outlet. The average residence time of the sugar in the dryer is about 7 min. The drum slowly rotates around its longitudinal axis (about 6 revolutions per minute), in order to expose the sugar to the air continuously, and to mix the sugar. To this end, the internal surface of the drum is entirely covered with small “hand-shaped” plates (usually called “flights” — see Fig. 2), which lift the sugar in the drum. The inlet “wet/hot” sugar passes through the dryer, moving counter-current to the air flow, and out the lower end where it is dropped onto a conveyor which takes the “dried/cooled” sugar to the sugar hopper, from where it is transported to the customer. A fan is enclosed at the end of the dryer, which draws the air in through a duct and air conditioner. A fan in the air conditioner also provides forcing. When the air exits the dryer it passes through a scrubber to remove sugar dust.

The Plane Creek Mill has actually two parallel dryers that operate all year round as the finishing stage of very low colour (VLC) sugar production.

A simple model of the dryer can be derived by first principles. Specifically, the following main phenomena have to be taken into account:

- the evaporation (removing both water and excess energy from sugar to the air);
- the convection (removing excess energy from sugar to the air);
- the transport of sugar and air along the drum in counter-flow.

The description of these main effects can be complemented — when necessary — with the following phenomena:

- the physical process of the wet sugar crystals sticking together;

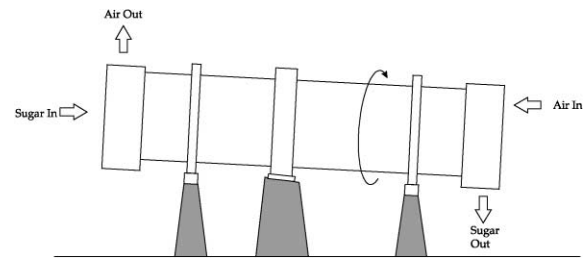


Fig. 1. Diagram of a standard rotary drum dryer.

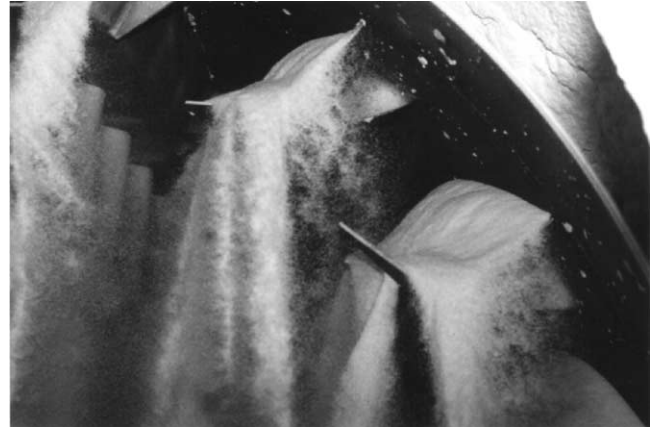


Fig. 2. Internal surface of a sugar dryer in working condition (detail of two “flights”).

- diffusion effects affecting the retention of water by the drying sugar due to the hygroscopic properties of the dry sugar;
- impurity levels in the water layer, raising the boiling point;
- spontaneous crystallisation in the water layer.

The model presented herein (originally described in Bitmead, Kammer, Jones, & Connolly, 1997; Tait, Schinkel, & Grieg, 1997; Wright & Shardlow, 1997, and recalled here for the sake of the consistency of this work) represents a slightly modified version of the model proposed and validated in Douglas et al. (1993). Specifically, the time and spatial discretisation is optimised here to deal with the two clearly separated dynamics of vapour/air and sugar/moisture velocities.

Following a typical approach used to model drums characterised by high length-to-diameter ratio, a “finite-element” approach is used: the length of the dryer is subdivided into a number N of equal-length “slices”, so reducing an infinite-dimensional model (the dynamic behaviour of the dryer is rigorously described by a set of partial differential equations) into a finite-dimensional one. The dryer is therefore assumed to be completely described by $4N$ state variables, given by (The subscript

index “ i ” indicates the actual slice.):

$$\begin{bmatrix} M_1^m(t) & M_2^m(t) & \cdots & M_i^m(t) & \cdots & M_N^m(t) \\ M_1^v(t) & M_2^v(t) & \cdots & M_i^v(t) & \cdots & M_N^v(t) \\ T_1^s(t) & T_2^s(t) & \cdots & T_i^s(t) & \cdots & T_N^s(t) \\ T_1^a(t) & T_2^a(t) & \cdots & T_i^a(t) & \cdots & T_N^a(t) \end{bmatrix},$$

where $M_i^m(t)$ is the mass of moisture in the sugar (liquid water), $M_i^v(t)$ is the mass of vapour (gaseous water) in the air, $T_i^s(t)$ and $T_i^a(t)$ are the temperatures of sugar/moisture and air/vapour, respectively. It is assumed that the moisture has the same temperature as the sugar, and the vapour has the same temperature as the air. Notice that the sugar and air masses (which will be indicated with the symbols $M_i^s(t)$ and $M_i^a(t)$) are modelled as being constant and the same for each slice.

Discretisation in time is a somewhat more difficult task than spatial discretisation, since the air/vapour velocity is dramatically higher (around 100 times) than that of the sugar/moisture. A suitable choice for the sampling rate τ is given by the *residence time of air in each slice*, namely:

$$\tau = \frac{\text{length of dryer (m)}}{\text{number of slices}} \cdot \frac{1}{\text{velocity of air (m s}^{-1}\text{)}}.$$

Given the above definitions, the model equations are the following:

Evaporative mass transfer: This paper concentrates only on evaporation as the vehicle of mass transfer. The extent of evaporation depends upon the difference between the partial pressure of moisture in the film surrounding the sugar crystal and the partial pressure of vapour in the air. It is given by

$$E_i(k\tau) = m_x A_i \tau (P_{p-sugar_i}((k-1)\tau) - P_{p-air_i}((k-1)\tau)), \quad (1a)$$

where m_x is the mass transfer coefficient, A_i is the available sugar surface area, while $P_{p-sugar_i}$ and P_{p-air_i} are the partial pressure of water in the sugar and in the air, respectively, within slice i . They are given by

$$P_{p-sugar_i}(k\tau) = \exp\left(16.31 - \frac{3829.48}{T_i^s(k\tau) + 227.51}\right),$$

$$P_{p-air_i}(k\tau) = 101.3 \frac{1}{1 + 18M_i^a(k\tau)/28.818M_i^v(k\tau)}. \quad (1b)$$

The Eq. (1b) used to compute the partial pressure of water in the sugar is the Antoine formula, which assumes that the water film surrounding the sugar crystals behaves as free liquid water (see e.g., Incropera & de Witt, 1990; Perry & Green, 1997). Note that this assumption (used also in Douglas et al., 1993) is realistic only when the moisture content is comparatively large (more than 0.1–0.2%).

The time updates for the moisture and vapour masses in slice i at time $k\tau$ are given by

$$\begin{aligned} \bar{M}_i^m(k\tau) &= M_i^m((k-1)\tau) - E_i(k\tau), \\ \bar{M}_i^v(k\tau) &= M_i^v((k-1)\tau) + E_i(k\tau). \end{aligned} \quad (1c)$$

We denote intermediate variables, before spatial updates, by using an overbar.

Energy transfer: As already pointed out, energy transfer occurs due to evaporation and convection; accordingly, the temperature changes are computed by enthalpy balance as follows:

$$\begin{aligned} \bar{T}_i^a(k\tau) &= T_i^a((k-1)\tau) \\ &+ \frac{(h_x A_i \tau + C_{pv} E_i(k\tau))(T_i^s((k-1)\tau) - T_i^a((k-1)\tau))}{C_{pa} M_i^a + C_{pv} \bar{M}_i^v(k\tau)}, \\ \bar{T}_i^s(k\tau) &= T_i^s((k-1)\tau) - \frac{L_{H_2O} E_i(k\tau) + h_x A_i \tau (T_i^s((k-1)\tau) - T_i^a((k-1)\tau))}{C_{ps} M_i^s + C_{pm} \bar{M}_i^m(k\tau)}, \end{aligned} \quad (1d)$$

where C_{pa} , C_{pv} , C_{ps} , and C_{pm} are the heat capacities of air, vapour, sugar and moisture, respectively, h_x is the convection heat transfer coefficient, and L_{H_2O} is the latent heat of vaporisation of water. As before, A_i is the available surface area within slice i . In the above equations it is assumed that the moisture and vapour mass changes within a slice are small enough that they may be considered as being negligible for enthalpy balance purposes.

Transport: The dynamical model is realised by calculating the heat and mass transfers for each slice, and then shifting the appropriate values along to the next slice. Notice that the air temperature and vapour masses in each slice are simply shifted along to the next slice after each time update. However, the air/vapour mixture moves through the dryer much more quickly than the sugar/moisture mixture, so the sugar temperatures and moisture masses are only partially shifted after each time update. So, the shifts of the four state variables of each slice (moisture and vapour masses, sugar and air temperatures) are given by

$$\begin{aligned} T_i^a(k\tau) &= \bar{T}_i^a(k\tau), \\ M_i^v(k\tau) &= \bar{M}_i^v(k\tau), \\ T_i^s(k\tau) &= (1 - \alpha) \bar{T}_i^s(k\tau) \\ &+ \frac{\alpha \bar{T}_{i-1}^s(k\tau) (C_{ps} M_{i-1}^s + C_{pm} \bar{M}_{i-1}^m(k\tau))}{C_{ps} M_i^s + C_{pm} \bar{M}_i^m(k\tau)}, \\ M_i^m(k\tau) &= (1 - \alpha) \bar{M}_i^m(k\tau) + \alpha \bar{M}_{i-1}^m(k\tau), \end{aligned} \quad (1e)$$

where $\alpha = (\text{velocity of sugar})/(\text{velocity of air})$.

The model (1) is based on first-principles, and it is characterised by many physical constants and

parameters. This model can now be used in two slightly different ways: as a *fixed* physical model, or as a *grey-box* model. In the former case, the numerical value of the whole set of parameters is calculated from a priori knowledge of the physical characteristics of the system; this computation is typically followed by a validation procedure, which is made using a set of data collected on the real plant. In the latter case, a few parameters (typically the parameters whose physical-principle-based estimation is difficult or unclear) are assumed to be unknown, and a posteriori estimated from real data.

Apparently, the difference between these two approaches is subtle and somewhat fuzzy: they both make use of a first-principles model, and a set of measured data. Grey-box modelling typically has the advantage of providing a better fit between model and data; on the other hand, the estimated value of the free parameters in a grey-box model may lose a clear physical meaning. This is due to the intuitive fact that the uncertainties, noise, and modelling errors are condensed and taken into account by a small subset of parameters only.

In this work a grey-box point of view is adopted, because of the paucity of data required to fit many parameters. Specifically, the model parameters which are assumed to be unknown and used as “tuning knobs” are given by:

- the free surface area per slice of the sugar in the drum, A_i (m^2);
- the heat transfer coefficient, h_x ($\text{kW}/\text{m}^2 \text{K}$);
- the mass transfer coefficient, m_x ($\text{kg}/\text{m}^2 \text{s kPa}$);

It is important to notice that, due to the model structure (see (1a) and (1d)), the number of parameters to be estimated is in fact two for each slice: $A_i h_x$ and $A_i m_x$. In the rest of the paper the total sugar-free surface $A = \sum_{i=1}^N A_i$ will be considered fixed at the value of 5000 m^2 (this value is a re-scaled version — for a different sugar flow-rate — of a value empirically estimated on a similar rotary drum dryer — see Bitmead et al., 1997),

simply assuming that $A_i = A/N$. This is an acceptable approximation when the sugar is very dry; on the contrary, if the sugar is wet and important sticking effects occur, A_i is a function of i .

Finally, it is important to remark that the structure of the above model (1) has been already extensively validated using real data in Douglas et al. (1993) on two sugar plants located in *Victoria* and *Kalamia*, and in Bitmead et al. (1997) on a sugar plant located in *Macknade*. The validation, however, has been done only in operating conditions characterised by sugar moisture content higher than 0.2%.

3. Analysis of the measured data

Four dryer performance trials were undertaken at the Plane Creek plant during July 1998 (Griffith, 1998). Each trial lasted about 1 h. Table 1 describes the measured input/output variables. Sugar moisture content is given by the mass ratio between liquid water and dry sugar.

Notice that the sugar inlet flow-rate is measured simply by manually counting the number of drops from the centrifuges (usually called “*fugals*”), that represent the last stage of the plant before the dryer. Since only the average amount of sugar contained in each fugal is known, the measured inlet flow-rate represents only a rough estimate of the actual one. Also note that the outlet air temperature and humidity are unavailable for the measurement because of the use of water scrubbers to remove sugar dust.

The four trials were performed under slightly different conditions to ascertain the effects of these conditions upon dryer performance. Trial #1 was performed in the mid-afternoon under standard operating conditions. Trial #2 was undertaken in the early morning to determine the effects of the cooler air and humidity variations. Trials #3 and #4 were performed during the afternoon of the same day, but the dryer load was changed, in order

Table 1
Measured input/output variables

Measured variable	Measurement unit	Method of measurement
<i>Inputs</i>		
Sugar temperature	$^{\circ}\text{C}$	Measured with a thermometer
Sugar inlet flow-rate	ton/h	Manually counted number of fugal drops
Sugar inlet moisture content	%	Samples taken and analysed in laboratory
Air temperature	$^{\circ}\text{C}$	Measured with a thermometer (dry-bulb measurement)
Air inlet flow-rate	ton/h	Turbine anemometer used over a 1 min period
Air absolute humidity	%	Computed from wet-bulb temperature measurements
<i>Outputs</i>		
Sugar temperature	$^{\circ}\text{C}$	Measured with a thermometer
Sugar exit moisture content	%	Samples taken and analysed in laboratory

Table 2
Distillation of measured data

	Trial #1			Trial #2			Trial #3			Trial #4		
	Lower bound	Central value	Upper bound	Lower bound	Central value	Upper bound	Lower bound	Central value	Upper bound	Lower bound	Central value	Upper bound
<i>Inputs</i>												
Sugar temperature (°C)	53.6	54.3	55.1	53.5	53.8	54.2	55.0	55.6	56.2	55.0	55.6	56.2
Sugar flow-rate (ton/h)	37.6	39.1	40.7	37.8	39.1	40.5	50.0	51.5	52.9	31.4	32.3	33.3
Sugar moisture content (%)	0.606	0.665	0.725	0.749	0.825	0.901	0.799	0.843	0.888	0.799	0.843	0.888
Air temperature (°C)	27.7	27.8	27.9	20.2	20.4	20.6	22.6	22.7	22.8	22.6	22.7	22.8
Air flow-rate (ton/h)	19.2	19.7	20.1	19.1	19.3	19.6	18.9	19.2	19.5	18.9	19.2	19.5
Air absolute humidity (%)	0.844	0.850	0.856	0.580	0.590	0.600	0.675	0.680	0.685	0.675	0.680	0.685
<i>Outputs</i>												
Sugar temperature (°C)	37.2	37.5	37.8	30.3	30.7	31.1	31.3	31.8	32.2	28.8	29.1	29.4
Sugar moisture content (%)	0.046	0.048	0.050	0.061	0.064	0.066	0.040	0.042	0.045	0.026	0.029	0.031

to understand better the effect of load variations on the output variables.

The sampling time was 5 min. Therefore, for each variable, only 10–13 samples per-trial are available.

During each trial, the six inputs and the two outputs of the dryer are assumed to be constant (even though obviously affected by noise). According to this assumption, the average value of each measured variable and their “range of indetermination” are numerically computed; it is assumed that, within such a range, every value is equally likely and good for data-fitting purposes.

Specifically, the mean value of each measured variable — say $y(t)$ — is computed using the standard sample mean-value estimation

$$\hat{y} = \frac{1}{n} \sum_{t=1}^n y(t),$$

where n is the number of available data-points while the lower and upper bounds (say L_y and U_y) of the “range of indetermination” are computed as

$$L_y = \hat{y} - \frac{\hat{\lambda}_y}{\sqrt{N}},$$

$$U_y = \hat{y} + \frac{\hat{\lambda}_y}{\sqrt{N}},$$

where $\hat{\lambda}_y$ is the estimated standard deviation of the measured variable, computed as

$$\hat{\lambda}_y = \sqrt{\frac{1}{n} \sum_{t=1}^n (y(t) - \hat{y})^2}.$$

Rigorously speaking, the definition of “range of indetermination” as $[L_y, U_y]$ has a simple but precise mathematical meaning: it is the interval $[\mu - \sigma, \mu + \sigma]$, μ and σ being the expected value and the standard deviation of

$y(t)$, respectively. This is a sound definition, if the following set of assumptions holds:

- the measured variables can be modelled as a constant value plus white noise: $y(t) = \bar{y} + \varepsilon(t)$ $\varepsilon \approx WN(0, \lambda^2)$; namely it is assumed that the process $y(t)$ is at steady state, and the additional noise has no residual dynamics;
- $y(t)$ is a stationary process, namely the probability distribution of $y(t)$ is independent of the time index;
- the probability distribution of $y(t)$ in the interval $[\mu - \sigma, \mu + \sigma]$ is approximately flat. Note that this assumption is roughly fulfilled if we make the standard assumption of Gaussian, uniform, or similar distribution of $y(t)$.

These assumptions fit reasonably with the true data-generation process. In Table 2, the lower bound, the mean/central value and the upper bound of the “range of indetermination” is reported for each variable. Such values will be used in the rest of the work, for model-fitting purposes.

Remark. The main remark to be made about the data of Table 2 concerns the output sugar moisture content. Notice that the output sugar moisture content is always very low (about $[0.03, 0.06]\%$, which is much smaller than the 0.2% “target” moisture content). This is due both to the efficiency of the Plane Creek dryer, which is characterised by a comparatively high length-to-diameter ratio, and to the fact that all the trials were made in winter-time, when the outside air is comparatively dry and cool.

It is important to notice that, in these conditions (namely when the moisture content is lower than 0.1%), discriminating between two values of moisture contents has little or no meaning. A value in this region can be “conventionally” approximated by zero. A major implication of this fact is that the output sugar moisture

content cannot — in practice — be used for model fitting purposes. A model capable of accurately predicting the sugar moisture content within the $[0, 0.1]\%$ region must necessarily be much more complicated than (1), since, in this region, the Antoine formula for the computation of the partial pressure of moisture in the sugar is no longer valid, and diffusion effects should be taken into account (hygroscopic properties, impurity levels, spontaneous crystallisation, etc.). The problem of finely modelling the drying rate at very low moisture content has been already considered in the literature (see e.g. Farkas, Remenyi, & Biro, 1998b; Van Boxtel, & Knol, 1996; Toyoda, 1992). However, as already pointed out, the relevance of developing a model capable of accurately predicting the moisture content in this region is — for control purposes — negligible; as a consequence, in the rest of the paper the simple model (1) will continue to be used, without discriminating between different values of moisture content in the range $[0, 0.1]\%$.

4. Fitting the data with the model: optimisation and analysis

As already pointed out, the model (1) was first developed for the interpretation of the *CSR Ltd. Macknade* plant data collected during the crushing season 1997 (Bitmead et al., 1997), and it is characterised by two parameters (m_x and h_x) which are unknown. The rest of this section is devoted to the problem of tuning the parameters m_x and h_x in order to obtain the “best” fit between the output values predicted by the model and those measured on the real plant. Specifically, in Section 4.1 it will be shown that the model in “standard-mode” cannot fit the measured data, whatever the values of m_x and h_x are; in Section 4.2 it will be shown that, with a suitable choice of m_x and h_x , the measured output sugar temperature can be accurately predicted, if the model is allowed to operate both in “standard-mode” and in “overdried-mode”. In both cases, the optimal values of m_x and h_x are obtained by minimisation of the summed squared error between measured and predicted outputs.

4.1. Model operating in “standard-mode”: analysis and discussion

The most intuitive and natural approach to the problem of fitting the measured data (Table 2) with the model (1) is to use the model in “standard-mode”, namely the operating mode characterised by a sugar moisture content higher than zero at any point of the drum. In Fig. 3 an example of the computed profile of the sugar temperature and moisture content along the drum in standard mode is displayed; apparently, in this operating mode both sugar temperature and moisture content have a regular (almost linear) behaviour along the drum.

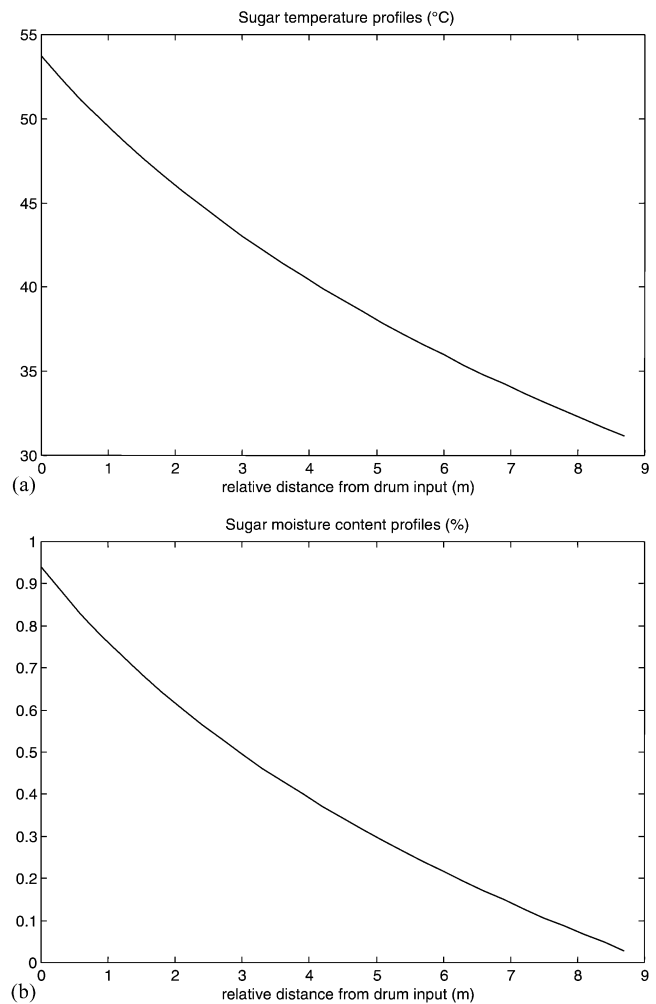


Fig. 3. (a) Example of sugar temperature profiles, in “standard-mode”. (b) Example of sugar moisture content profiles, in “standard-mode”.

Owing to the fact that there are just two parameters involved in the optimisation/fitting process, the problem of searching for the “optimal” values of parameters m_x and h_x is comparatively easy, and — in practice — can be solved by exhaustive exploration, namely by finely gridding the ranges of admissible values for such parameters. The result obtained by running the optimisation process is somewhat surprising: the input/output data of Table 2 cannot be fitted by a model constrained to operate in “standard-mode”.

With the aim of gaining some insight into the main reasons of this behaviour of the model, a local linear model was computed about a nominal operating condition of the model working in “standard-mode”. If there is no switching from one mode to another, a linearised model just introduces small approximation errors, while providing a useful and handy analysis tool.

In order to compute a local linear model, first a set of nominal inputs for the non-linear model, and two nominal values for m_x or h_x has to be selected. It is easy to see

Table 3
Nominal point about which the local linear model is computed

	Nominal value
<i>Inputs</i>	
Sugar temperature (°C)	53.8
Sugar flow-rate (ton/h)	39.1
Sugar moisture content (%)	0.825
Air temperature (°C)	20.4
Air flow-rate (ton/h)	19.3
Air absolute humidity (%)	0.59
<i>Outputs</i>	
Sugar temperature (°C)	30.877
Sugar moisture content (%)	0.062
<i>Parameters</i>	
Heat transfer coefficient h_x (kW/m ² K)	0.003
Mass transfer coefficient m_x (kg/m ² s kPa)	2.7×10^{-6}

that if $h_x = 0.003$ kW/m²K and $m_x = 2.7 \times 10^{-6}$ kg/m²s kPa (the values used for the data collected on the CSR Ltd. Macknade plant in 1997 — see Bitmead et al., 1997), the I/O variables of Trial #2 are fitted quite accurately. Henceforth, we take the central values of the inputs in Trial #2 as “nominal inputs”. In Table 3 the nominal inputs, outputs and parameters are summarised. Since the output moisture content is strictly larger than zero, the nominal operating condition is a “standard-mode” operating condition.

It is important to stress that the validity of the following analysis only marginally depends on the nominal inputs and parameters taken. This fact has been empirically tested by computing local linear models about different sets of inputs and parameters (those of Trials #1, #3, and #4). The resulting local linear models are close to each other. This reveals that, if we do not switch operating mode, the model is in fact only slightly non-linear.

The local linear model is numerically computed by giving an independent +1% variation to each input, about the nominal equilibrium point. Note that this is the easiest way of computing the linear approximation, even if, in principle, it could be computed directly from the steady-state equations of (1), or even by direct small-signals excitation of the real plant; clearly, due to the inherent difficulty of collecting real data, the latter approach is not practicable. The equations of the obtained linear model are given by:

$$\begin{aligned}
 & (\text{SugarTempOut}-30.877) \\
 & = 0.2593(\text{SugarTempIn}-53.8) + 0.3427(\text{SugarFlow}-39.1) \\
 & + 0.3030(\text{SugarMoistIn}-0.825) + 0.2971(\text{AirTemp}-20.4) \\
 & - 0.2176(\text{AirFlow}-19.3) + 1.8136(\text{AirHumid}-0.59), \quad (2a)
 \end{aligned}$$

$$(\text{SugarMoistOut}-0.062)$$

$$\begin{aligned}
 & = -0.0281(\text{SugarTempIn}-53.8) + 0.0095(\text{SugarFlow}-39.1) \\
 & + 1.0061(\text{SugarMoistIn}-0.825) - 0.0034(\text{AirTemp}-20.4) \\
 & + 0.0031(\text{AirFlow}-19.3) + 0.1356(\text{AirHumid}-0.59) \quad (2b)
 \end{aligned}$$

In order to understand better and to compare the effects of input variations on the outputs, in Fig. 4a and b the absolute variations of output sugar temperature and moisture content to +1% input variations are displayed. The following observations can be drawn:

- The sugar temperature, the sugar flow-rate and the air temperature are the input variables that mainly affect the output sugar temperature, whereas the sugar moisture content and air humidity seem to play a negligible role in determining output sugar temperature.
- The sugar temperature, the sugar flow-rate and the sugar moisture content are the input variables that mainly affect the output sugar moisture content. Although, all the characteristics of input air (temperature, humidity and flow-rate) seem to have a negligible effect on the sugar moisture content.
- It is interesting to notice (see Eq. (2b)) that the gain from input moisture content to output moisture content is almost exactly 1. This suggests that model (1) cannot accurately explain the output sugar moisture content when the sugar is overdried, since the gain from input to output moisture content abruptly switches from 1 to 0, while an accurate moisture content prediction when the sugar is overdried would require a smoother transition from gain 1 to gain 0. As already pointed out, this is due to the fact that the Antoine formula for the computation of the partial pressure of the moisture in the sugar is no longer valid when the sugar moisture content is very low (in the range [0, 0.1]%).

An intuitive yet simple analysis to understand roughly the extent to which the model in “standard-mode” is unable to model the measured I/O behaviour of the Plane Creek sugar dryer is to predict the output sugar temperature of Trials #1, #3, and #4 using the local linear model (2).

In Table 4 the results obtained by feeding the linear model (2) with the input measurements of Trials #1, #3, and #4 are summarised.

It is important to notice that, in order to be consistent with the approach based upon the definition of “ranges of indetermination” for each measured variable, the input values which provide the best fit with the measured output temperatures were selected, within their ranges of indetermination.

The large prediction error made by the linear model (2) in the case of Trials #1 and #3 clearly denotes that the model in “standard-mode” is inadequate to provide an

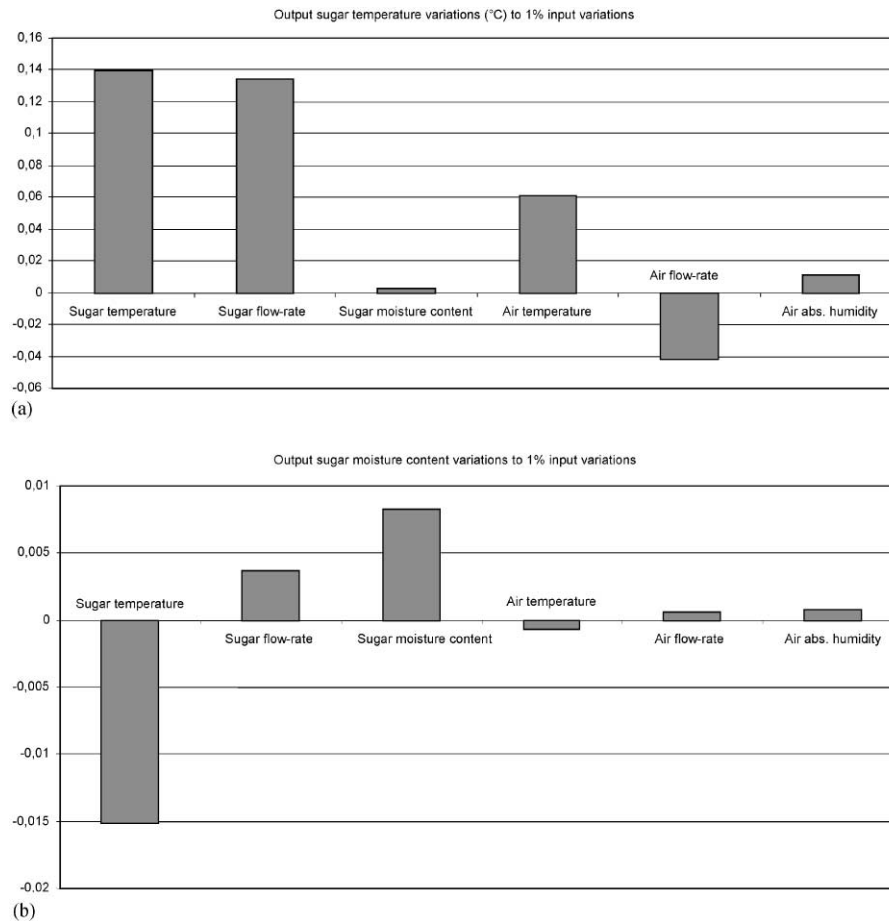


Fig. 4. (a) Effect of + 1% input variations on the output sugar temperature. (b) Effect of + 1% input variations on the output sugar moisture content.

Table 4
Fitting the output sugar temperature using the linear model

Trial #	Measured output sugar temperature (°C)	Output sugar temperature predicted by the linear model (°C)	Error (°C)
1	37.5	34.5	- 2.0°C
2 (nominal)	30.9	30.9	—
3	31.8	35.7	+ 3.9
4	29.1	29.3	+ 0.2

accurate description of the dryer behaviour. Specifically, notice that, since the linear model strongly underestimates the output temperature in Trial # 1 and strongly overestimates the output temperature in Trial # 3, the sugar drier behaviour in Trials # 1 and # 3 must be explained using two different models or two different “working conditions”.

4.2. “Beyond the knee”: fitting the model with measured data

In the previous subsection it was shown that the model in “standard-mode” is unable to fit the measured output

sugar temperatures. The main indication we get from this “negative” result is that the sugar dryer is likely to be characterised by operating modes different from the “standard” one.

An alternative operating mode, which is quite intuitive and simple to interpret is the mode that will be called the “overdried-mode”. The “overdried-mode” is characterised by sugar moisture content effectively equal to zero from a certain point of the drum on. This mode refers to the case when the sugar at the output is overdried. It is worth noticing that a zero-moisture condition is just an ideal condition that, in practice, never happens, since a little amount of moisture always remains in the sugar.

This operating mode of the model, however, is useful since it clearly stresses and represents what actually happens in the drum when the evaporation process of the sugar moisture becomes negligible. As it is apparent from Fig. 5 (where an example of sugar temperature and moisture content profile is depicted), in “overdried-mode” the temperature profile has a discontinuity in its first derivative when the zero-moisture condition is reached. This causes a “knee” in the sugar temperature profile, the slope of the latter part of the temperature profile being much lower, since — “beyond the knee” — the temperature decreases only due to convection.

The discontinuity between the standard and the overdried mode, clearly visible in Fig. 5, is just an approximation of the plant behaviour: the *true* moisture and temperature profiles are obviously characterised by a smooth transition between these two working conditions. As already said, this is due to the fact that, for low-moisture conditions, the model presented here cannot accurately predict the moisture content; however, for the purposes of this work, a finer model of the transition region is somewhat unnecessary and redundant.

It is worth observing here that, in the literature, this “two-stage behaviour” has already been observed in materials other than sugar, and labelled with a different nomenclature. In van Boxtel and Knol (1996) (where a fluid-bed dryer model is developed and proposed), for instance, the “standard” and “overdried” modes are named *constant flux period* (CFP) and *constant surface period*, respectively. *Constant rate period* (CRP) and *falling rate period* (FRP) is another way of naming the same phenomenon. Another piece of literature where multiple-stage behaviour was observed and described is the work of Toyoda and co-authors (Toyoda, 1989, 1992; Toyoda, Farkas, & Kojima, 1995, 1996), where the so-called “two-tanks model” of rough rice is proposed. In these works three different drying conditions (named *drying period* I, II, and III) are outlined: the intermediate working condition refers to the transition phase between standard and overdried modes. Finally, note that in these papers different working conditions are named “periods”, since it is assumed that different working modes appear at different times; this is not the case of a rotary sugar dryer, where different drying stages can exist simultaneously in different regions of the drum.

The new search for the “optimal” model parameters has been done under the following assumptions:

- The output sugar moisture content is not taken into account in the fitting procedure, since in all trials the output moisture content is very low. It was expected that the sugar output moisture content predicted by the model is somewhere in the range $[0, 0.1]\%$.
- The model was allowed to work both in “standard-mode” and in “overdried-mode”.

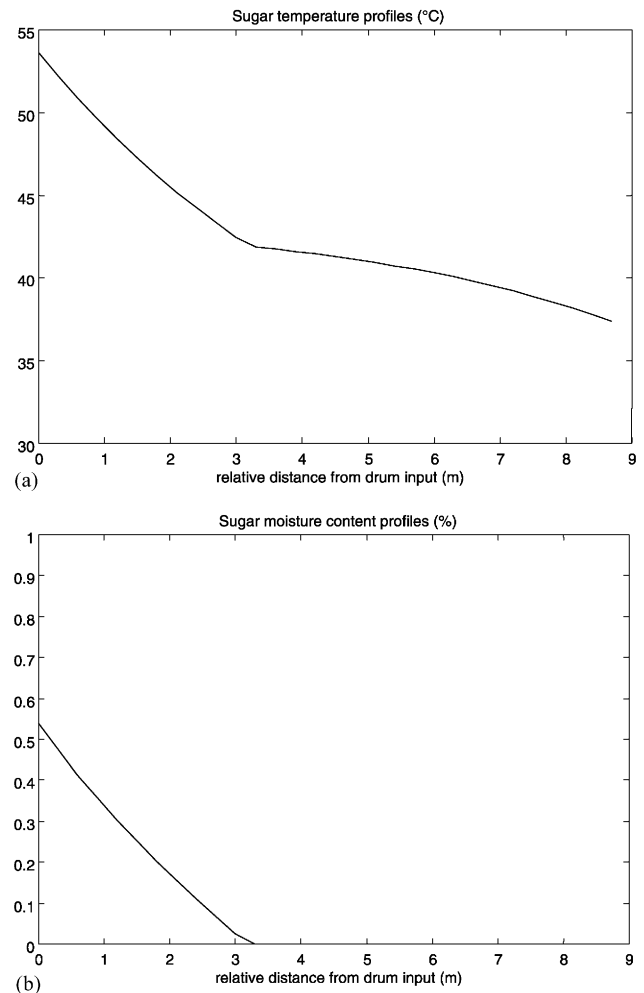


Fig. 5. (a) Example of sugar temperature profile, in “overdried-mode”. (b) Example of sugar moisture content profile, in “overdried-mode”.

- According to the results obtained in the previous subsection (Table 4), within their “range of indetermination”, the inputs of Trial #1 which maximise the output temperature, central input values of Trial #2, inputs of Trials #3 and #4 which minimise the output temperature (see Table 5) have been selected. Using these input values, the goal is to fit the central values of output sugar temperatures.

Under the above assumptions, it is easy to see that an exact matching of all the central values of output sugar temperatures can be achieved. This confirms the fact that the “overdried-mode” closely reflects what really happens in the sugar dryer, and that the model (1) is a simple but effective way of explaining the main phenomena occurring in the sugar dryer. Specifically, the best fit with the measured data has been obtained using the following values for the tuning parameters:

$$h_x = 0.0038 \text{ (kW/m}^2\text{K)},$$

$$m_x = 4.05 \times 10^{-6} \text{ (kg/m}^2\text{s kPa)}.$$

Table 5
Input values used for parameter-tuning

	Trial # 1	Trial # 2	Trial # 3	Trial # 4
Sugar temperature (°C)	55.1	53.8	55.0	55.0
Sugar flow-rate (ton/h)	40.6	39.1	50.0	31.4
Sugar moisture content (%)	0.606	0.775	0.888	0.888
Air temperature (°C)	27.9	20.4	22.6	22.6
Air flow-rate (ton/h)	19.2	19.3	19.5	19.5
Air absolute humidity (%)	0.844	0.590	0.675	0.675

Table 6
Fitting results

	Trial # 1	Trial # 2	Trial # 3	Trial # 4
Estimated output sugar temperature	37.5	30.7	31.8	29.1
Measured output sugar temperature (central-val.)	37.5	30.7	31.8	29.1
Estimated output sugar moisture content	0	0	0.005	0
Measured output sugar moist. content (central val.)	0.048	0.063	0.042	0.029

The estimation results are summarised in Table 6, whereas in Fig. 6 the corresponding temperatures and moisture content profiles along the drum are depicted.

By inspecting the results above outlined, the following remarks can be made:

- The dryer works “beyond the knee” in 3 cases out of 4. The only case of output moisture content higher than 0 (“standard-mode”) is Trial # 3.
- The sugar in Trial # 1 is, by far, the most “overdried”, since the moisture content gets to zero just 3 m from the drum inlet. This fact provides a simple explanation for the very high output sugar temperature, which cannot be otherwise explained.
- The low temperature of the output sugar in Trial # 3, despite the comparatively high input sugar temperature, air temperature, and sugar flow rate, is explained by the fact that it is the only “non-zero-moisture” condition, thus allowing a better sugar cooling.

The “two-stage” temperature profile predicted by the model is very difficult to validate by real data. As a matter of fact, it is a “spatial” profile (not a “time” profile). This implies that, in order to be validated, input/output measurements are not enough: sugar samples must be *simultaneously* taken at different positions along the drum. Apparently, the collection of such a sample set is inherently a very difficult task. For the Plane Creek plant only input/output data are available, and this validation is not possible. However, a few simultaneous data snapshots along the drum were taken (with an ad-hoc experiment and instruments) at the Macknade plant (experiments made in 1997 — see Bitmead et al., 1997). In Fig. 7 the temperature profile measured in a working condition where the output sugar moisture is very low (overdrying condition) is displayed. This profile confirms

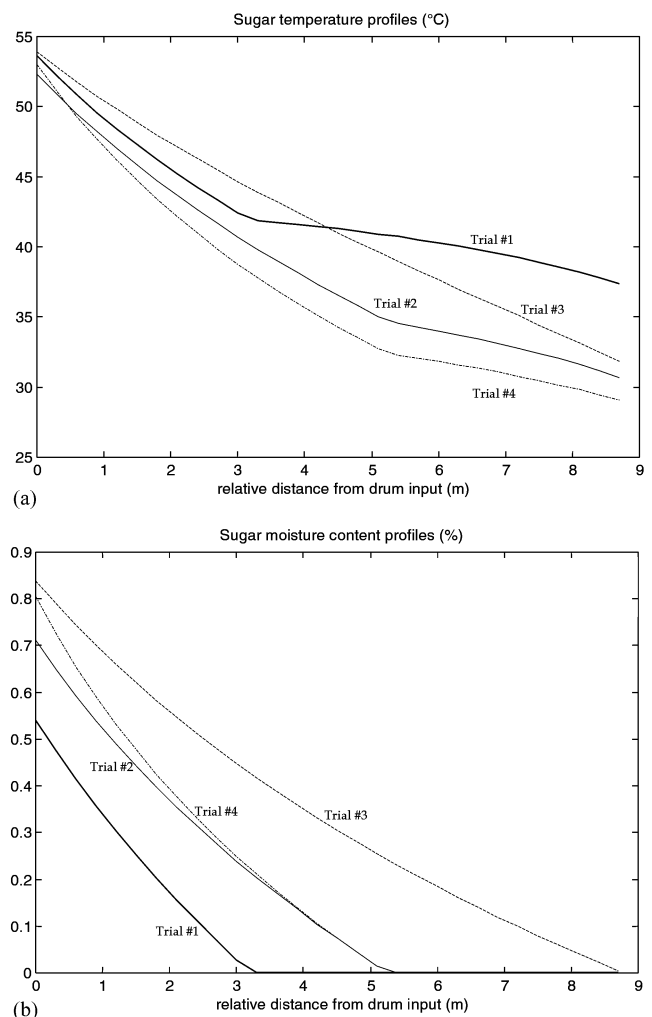


Fig. 6. (a) Sugar temperature profiles predicted by the model. (b) Sugar moisture content profiles predicted by the model.

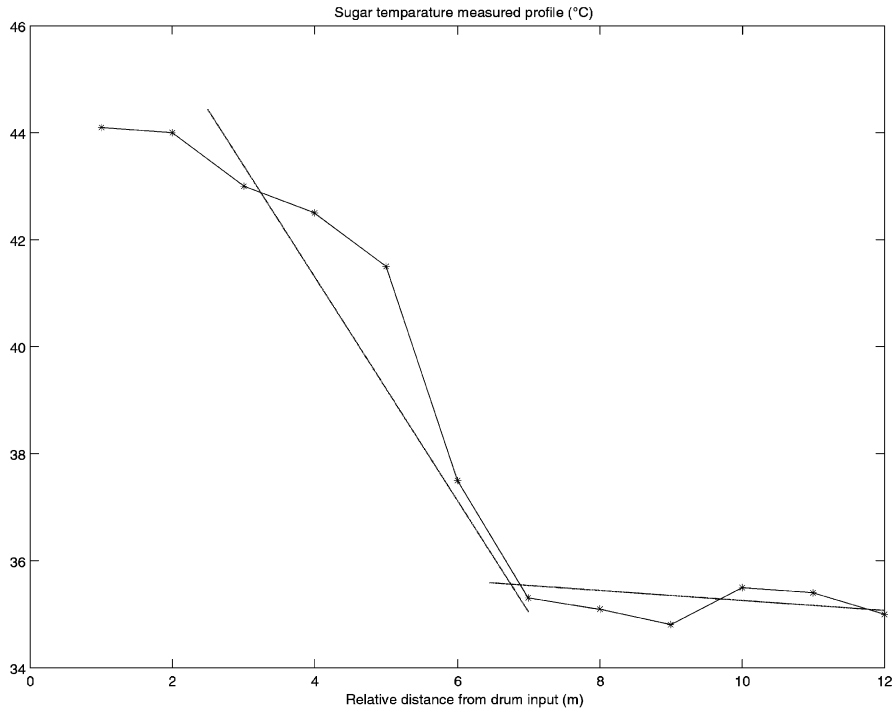


Fig. 7. A sugar temperature content profile measured at the *Macknade* plant, in an overdrying condition. The two main working conditions are highlighted with two straight dashed lines.

the two-stage behaviour predicted by the model. The two working stages (highlighted with two straight dashed lines in Fig. 7) are clearly separated and visible. Moreover, note that the transition phase between standard and overdried mode (occurring at about 6–7 m from the drum inlet) is very short. This also confirms that, in a simplified model for control purposes, the accurate modelling of such an intermediate phase can be neglected.

5. Controlling the output sugar of a dryer: hints and observations

The aim of this section is to propose control strategies to regulate the output temperature and moisture content of a rotary drum sugar dryer. First, in Section 5.1 a preliminary qualitative discussion on the best control actions for different working conditions is proposed. All the possible working states of the dryer are clustered into nine different operating conditions, and a control strategy is outlined for each of them.

In Section 5.2 an innovative regulation algorithm is proposed. Its main features are:

- it uses output sugar temperature measurements only;
- it self-detects the dryer’s working conditions;
- it keeps the dryer working “on the edge” between standard and overdried mode;
- it is very simple.

In order to develop the control strategies presented in the rest of this section, it is useful first to compute the linear approximation of the model working in “overdried-mode”. The linear approximation of the model (1) when working in “standard-mode” has been already derived in Section 4.1.

In Table 7 the nominal inputs, outputs, and parameters about which the linear model is computed are listed. Notice that the inputs are the central values of

Table 7
Nominal values of the model used for local linear analysis for a model in “overdried-mode”

	Nominal value
<i>Inputs</i>	
Sugar temperature (°C)	53.8
Sugar flow-rate (ton/h)	39.1
Sugar moisture content (%)	0.825
Air temperature (°C)	20.4
Air flow-rate (ton/h)	19.3
Air absolute humidity (%)	0.59
<i>Outputs</i>	
Sugar temperature (°C)	29.814
Sugar moisture content (%)	0
<i>Parameters</i>	
Heat transfer coefficient h_x (kW/m ² K)	0.0038
Mass transfer coefficient m_x (kg/m ² s kPa)	405×10^{-6}

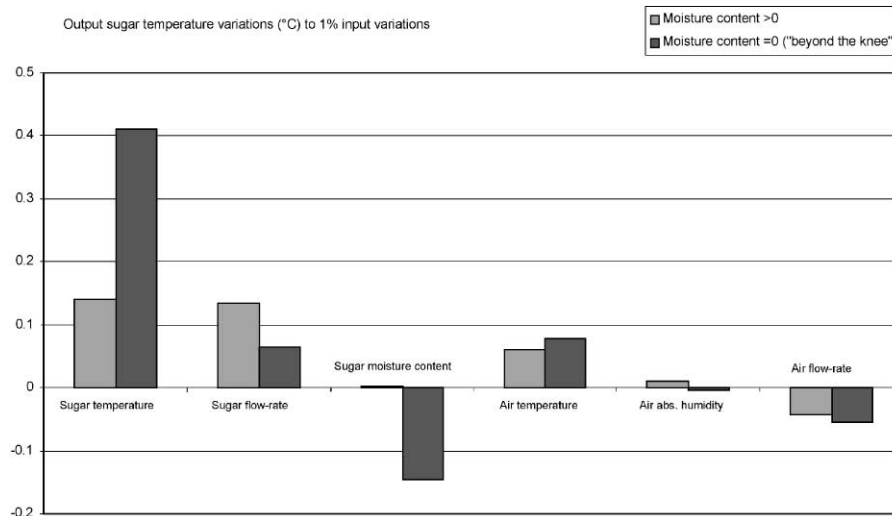


Fig. 8. Effect of +1% input variations on the output sugar temperature (“standard-mode” and “overdried-mode”).

Trial #2, whereas the parameters are those estimated in Section 4.2.

The equations of the local linear model in “overdried-mode” condition are given by:

$$\begin{aligned}
 (\text{SugarTempOut}-29.81) = & 0.7647(\text{SugarTempIn}-53.8) \\
 & + 0.1627(\text{SugarFlow}-39.1) \\
 & - 17.648(\text{SugarMoistIn}-0.825) \\
 & + 0.3868(\text{AirTemp}-20.4) \\
 & - 0.2793(\text{AirFlow}-19.3) \\
 & - 0.4576(\text{AirHumid}-0.59), \quad (3a)
 \end{aligned}$$

$$\text{SugarMoistOut} = 0. \quad (3b)$$

Obviously, for small input variations, the output sugar moisture content remains zero.

In Fig. 8 the absolute variations of output sugar temperature to +1% input variations are displayed, both in the case of “standard-mode” and “overdried-mode”. The analysis of Fig. 8 suggests the following remarks:

- The most impressive difference between the behaviour of the dryer in “standard-mode” and in “overdried-mode” is the effect of the input sugar moisture content on the output sugar temperature: in “standard-mode”, increasing the input sugar moisture content has little or no effect on the output sugar temperature; whereas, when the dryer works in “overdried-mode”, even a small increase of the input moisture content results in a remarkable reduction of the output sugar temperature. This is easily explained by the fact that in “overdried-mode” in the last part of the drum there is no evaporation, which is the main phenomenon helping the sugar cooling. The huge benefits of moving a drier

from “overdried-mode” to “standard-mode” can be fully appreciated in Fig. 9. The two curves refer to the same situation (in particular, central input values of Trial #1 are used), the only difference being that the input sugar moisture content is 0.65 in the “overdried-mode”, and 1.0 in the “standard-mode”. Apparently, by simply adding some water at input, the output sugar experiences a 6.5°C cooling.

- Input air humidity variation has opposite effects on output sugar temperatures in “standard-mode” and in “overdried-mode”: in the first case increasing the air humidity results in an increase of the sugar output temperature, whereas in “overdried-mode” this reduces the temperature.
- The sensitivity of output sugar temperature to input sugar temperature variations is much higher in “overdried-mode” than in “standard-mode”. This is due to the fact that, in “standard-mode”, the moisture in the sugar has a sort of filtering effect (the gain is about 0.4), whereas in “overdried-mode” a change in sugar input temperature affects more directly the output temperature (the gain is about 0.8). This is true also for air temperature variations.

5.1. Control strategies

Using the linearised models (2) and (3), whose behaviour is graphically summarised in Fig. 8, some simple qualitative control strategies can now be easily figured out. It is worth observing that these basic control actions could be, in principle, derived by elementary mass and heat balance equations (or even by direct data inspection). The linearised models (2) and (3) (graphically condensed in Fig. 8), however, provide a very simple, intuitive and effective decision tool for a human operator.

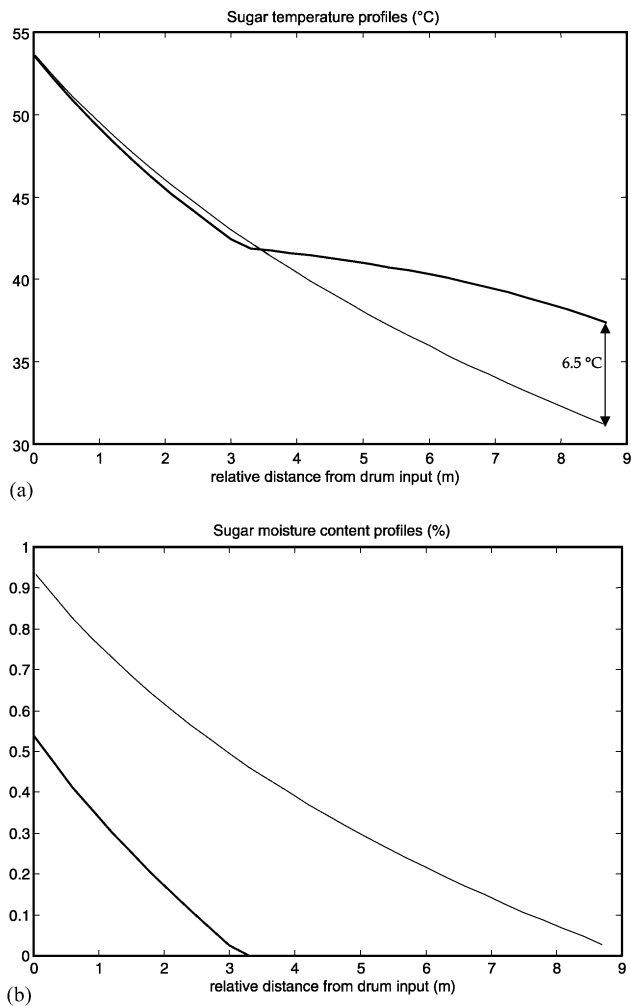


Fig. 9. (a) Example of sugar temperature profiles, in the two operating modes: “standard-mode” (thin line) and “overdried-mode” (bold line). (b) Example of sugar moisture content profiles, in the two operating modes: “standard-mode” (thin line) and “overdried-mode” (bold line).

For the sake of simplicity, the working domain of the dryer was divided into nine main clusters, according to the combination of three possible output sugar moisture contents (“low”, “O.K.”, or “high”) and three possible output sugar temperatures (“low”, “O.K.”, or “high”).

The control strategies suggested by the model in these nine conditions are the following (see Table 8):

Moisture content low–Temperature low: The dryer is working “beyond the knee”. The easiest thing to do is to spray water at the drum inlet. If water is sprayed, an additional cooling of the sugar is expected. Therefore, it is possible to take advantage of sugar cooling by decreasing the air flow-rate, turning the cooler off, or even increasing the sugar flow-rate.

Moisture content low–Temperature O.K.: The dryer is working “beyond the knee”. The easiest thing to do is to spray water at the drum inlet. If water is sprayed, the sugar is expected to cool down. Therefore, it is possible to

take advantage of sugar cooling by, e.g. decreasing the air flow-rate.

Moisture content low–Temperature high: The dryer is working “beyond the knee”. The easiest thing to do is to spray water at the drum inlet. If water is sprayed, the sugar is expected to cool down. This might be enough to get the right sugar output temperature. If not, input air must be cooled.

Moisture content O.K.–Temperature low: The input variables affecting the sugar moisture content should be kept untouched. If the sugar is well cooled, the possibility may be considered of turning the cooler off or reducing the air flow-rate.

Moisture content O.K.–Temperature O.K.: Keep untouched.

Moisture content O.K.–Temperature high: The input variables affecting the sugar moisture content should be kept untouched. Cool input air.

Moisture content high–Temperature low: The best fix is to increase input sugar temperature. If this is not enough, de-humidification of input air can be considered.

Moisture content high–Temperature O.K.: Increase input sugar temperature and cool input air. If not enough, de-humidification of input air can be considered.

Moisture content high–Temperature high: The first thing to do is to decrease input air humidity (it reduces both temperature and moisture content of the output sugar) and cool input air. If the moisture content is still high, but the sugar is over-cooled, raise the input sugar temperature. If not enough, the only thing to do is reduce sugar input flow-rate.

The above scheme provides useful and non-trivial guidelines for the control of the dryer. Unfortunately, the practical implementation of such control actions is difficult since it requires an accurate measurement of both output sugar temperature and moisture content. This problem will be addressed in the following subsection.

5.2. Keeping the dryer “on the edge”: the control scheme

As already pointed out, considerable benefits can be obtained by keeping the rotary sugar dryer working in “standard conditions”. Apparently, this might be easily obtained by spraying water at the input, according to the actual output moisture content.

The implementation of this control scheme is — however — very challenging. As a matter of fact the on-line un-manned measurement of the output sugar moisture content is quite a difficult task, since an accurate moisture content measurement can be done only by laboratory analysis of sugar samples, or by means sophisticated and expensive sensor equipment (see e.g. Rodriguez, Vasseur, & Courtois, 1996a; Toyoda et al., 1995; Toyoda, Kojima, Miyamoto, & Takeuchi, 1997). As a consequence, a control loop using the output sugar moisture content as feedback variable is, at the present stage, hardly practicable.

Table 8
Control strategies

Temperature → ↓Moisture content	Low	O.K.	High
Low (“beyond the knee”)	Spray water at input Decrease air flow-rate Turn air-cooler off Increase sugar flow-rate	Spray water at input Decrease air flow-rate	Spray water at input Cool input air
O.K.	(Turn air cooler off) (Decrease air flow-rate)	—	Cool input air
High	Increase sugar temperature Decrease air humidity	Increase sugar temperature & cool input air Decrease air humidity	Decrease air humidity Cool input air (& increase sugar temperature) (Decrease sugar flow-rate)

To overcome this problem, a simple and implementable control scheme is proposed, having the aim of keeping the sugar dryer working “on the edge” between “standard-mode” and “overdried-mode”. Notice that this is a wise control goal since:

- the “edge-condition” is as good as the “standard condition”;
- the value of output moisture content is only slightly lower than the target moisture content; this sub-optimality can be easily overcome by spraying a tiny amount of water at the output of the dryer.

The most appealing feature of the control scheme proposed is that it just requires the measurement of the output sugar temperature, which is a much easier task than measuring the moisture content. To this purpose, the traditional probe thermometers used at CSR plants, which suffer a performance decay due to sugar sticking around the probe, are being replaced by infra-red thermometers, providing a much more reliable measurement, and requiring little or no maintenance.

Before presenting the algorithm, it is worth pointing out that it provides sub-optimal tracking performance: the regulation of the output sugar moisture is comparatively rough, energy consumption is not minimised, and start-up trajectories are not optimised. As such, it can hardly be compared with more sophisticated techniques used (or proposed) for different types of dryers (see e.g. Courtois, 1996 or Quirijns, van Willigenburg, van Boxtel, & van Straten, 1999 for an overview). However, here optimality was not the goal in designing the controller: at the present stage robustness and simplicity are the crucial features for a rotary sugar dryer regulator.

The basic idea of the control scheme proposed is to exploit fully the fact that switching from standard to overdried mode results in the change of the sign of the gain from input moisture content to output sugar

temperature. In other words, when the dryer is working in over-drying conditions adding water at the dryer inlet results in a temperature decrease at the output, whereas when the dryer is working in “standard mode”, adding water results in a slight increase of the temperature.

The bulk of the control scheme proposed is depicted in Fig. 10. The entire control scheme is based upon the injection of a “impulse-like” signal (whose duration τ is quite short, e.g. 3 min), which is periodically superimposed (e.g. every $T = 30$ min) to the average amount of water sprayed at the sugar dryer inlet (signal ① in Fig. 10). The role of this signal is to excite the system, in order to detect the current operating condition of the dryer, and to change the amount of sprayed water accordingly.

Following clock-wise the block diagram of Fig. 10, from signal ① to signal ⑥ (see also Fig. 11 where all the signals involved are displayed in the time-domain), the role of each element of the control scheme can be summarised and explained as follows:

- Signal ② is the short-term response of the output sugar temperature to the water spray impulse; as a matter of fact it is obtained as the difference between the actual output sugar temperature and its average (low-pass filtered) value. According to the results depicted in Fig. 8, it was expected that signal ② is characterised by large negative impulses if the dryer is working in overdried-mode, and by small positive impulses if it is working in standard mode.
- The multiplication between signal ② and the delayed signal ① (producing signal ③) has just the role of ensuring that signal ③ is exactly zero outside the “impulse-window”, in order to avoid feeding the integrator with small but quite obnoxious non-zero low-frequency signals. Notice that the delay r_t , which must

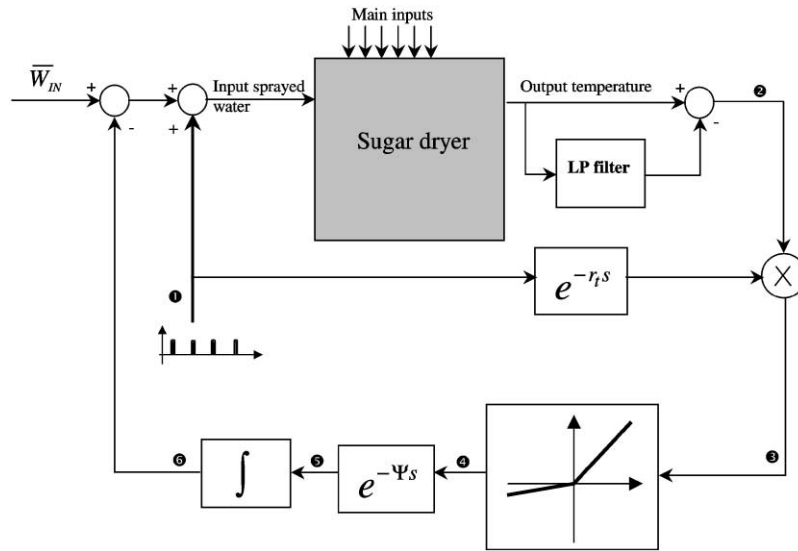


Fig. 10. The control scheme.

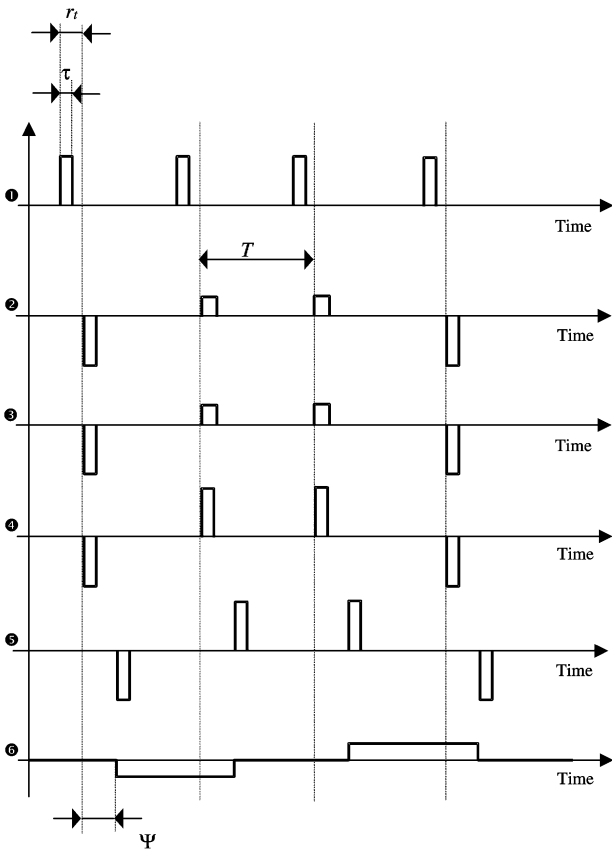


Fig. 11. The signals involved in the control scheme (example).

equal the residence time of the sugar in the drum (about 7 min in the case of the plane creek plant) is necessary since the effect of an impulse on ① appears on ② only after r_t minutes.

- The non-linear static characteristic has the role of re-balancing the size of positive and negative impulses in ③ (basically, it keeps unchanged the positive impulses, and multiplies the negative impulses by

$$\frac{(\Delta Temp_{out}/\Delta Moist_{in})_{standard-mode}}{(\Delta Temp_{out}/\Delta Moist_{in})_{overdried-mode}},$$

namely the absolute value of the ratio between the moisture-temperature gains of the dryer in standard and in overdried mode.

- The time delay Ψ (which must be strictly larger than r_t) is used to avoid interference between the “testing-phase” (during the impulse injection) and the “adjustment-phase” (after impulse injection).
- The role of the integrator is to produce a permanent adjustment to the amount of sprayed water, till the next impulse injection. The adjustment signal ⑥ is superimposed on the constant value of sprayed water \bar{W}_{IN} determined by the plant operator or, alternatively, by some automatic supervision algorithm according to the average external weather conditions.

The most interesting and peculiar features of the control scheme outlined above are the following:

- The “excitation” signal made up by a row of impulse-like signals allows a reliable estimation of the working condition of the dryer, since it is locally very strong in amplitude. However, being quite short in time, its effect on the average characteristics of the output sugar is negligible, especially after mixing in the hopper. To this purpose, it is important to stress that a “strong” excitation signal is called for, since the high variance of the six main input variables (temperature, flow and humidity of input sugar and

air) of the dryer would hide the effects of a weak excitation signal.

- The time-decoupling between the excitation impulse-like signal and the actual variation of the average amount of water sprayed simplifies the dynamic behaviour of the regulation loop.

Needless to say, the control scheme proposed here and outlined in Fig. 10 is only the framework of the overall control algorithm used in a “real-world” implementation. Problems like the pre-treatment of air, supervision of this control loop and its interaction with the slowly varying characteristics of the sugar and air characteristics at the dryer inlet must obviously be considered.

6. Related work

The *modelling* and *control* of dryers is an issue that has been extensively studied in recent years both from a theoretical and from a practical point of view. The existing literature on this topic can be given a first classification according to the type of dryers, and to the desiccated materials. Many types of dryers are used: drum dryers (Courtois & Trystram, 1994; Rodriguez et al., 1996a, Rodriguez, Vasseur, & Courtois, 1996b), cross-flow dryers (Douglas, Jones, & Mallick, 1994a–c), solar dryers (Farkas, Mészáros, & Seres, 1998a), fluidised-bed (van Boxtel, & Knol, 1996; Temple & van Boxtel, 2000) and fixed-bed (Farkas, Remenyi, & Biro, 1998b) dryers, mixed-flow (Courtois, Nouafo, & Trystram, 1995) and cross-flow (Platt, Rumsay, & Palazoglu, 1992) dryers, and *rotary dryers* (Douglas et al., 1993) are just the main example of existing dryers. The variety of desiccated material is huge: maize (Courtois, Lebert, Duquenoy, Lasseran, & Bimbinet, 1991), rice (Bonazzi et al., 1994; Toyoda, 1989, 1992), corn (Courtois, 1995; Courtois et al., 1995; Trelea, Trystram, & Courtois, 1997), agricultural products (Toyoda et al., 1997), tea leaves (Temple & van Boxtel, 1999) are just a few examples.

An important feature that can be used to categorise the literature is the modelling approach: *black-box* or *first-principles*. Traditionally, black-box approaches are rarely used in this field, since physical insight in the model is important, and a large set of “good” (stationary and characterised by high signal-to-noise ratio) input/output measurements is rarely available. The black-box models typically used are standard *ARMA*, *BJ*, or *OE* models, if linearity is assumed (Toyoda et al., 1995), or neural-nets (or similar non-linear parametric functions) if non-linear phenomena must be captured (Farkas et al., 1998b, 1998c; Trelea et al. 1997). On the other hand, traditional first-principles models range from partial differential equations (sometimes replaced by lumped-parameter models obtained by spatial dis-

cretisation — see e.g. Douglas et al., 1993) to sophisticated stochastic models (Mészáros, Farkas, & Balint, 1999).

The control strategy used to regulate the dryer is another main distinctive feature. A vast selection of control approaches has been tested: from classical (PID-based or LQG) and recently developed (H_∞) techniques for linear systems, up to “trendy” non-linear control schemes (neural-networks or fuzzy-logic based approaches — see e.g. Zhang & Litchfield, 1993). A comprehensive overview of control strategies for dryers regulation is in Courtois (1996) or in Quirijns et al. (1999).

It is worth mentioning that, in dryer control problems, the issue of measuring the moisture content of the desiccated material has always attracted a special attention. The fast, cheap and accurate measurement of moisture content is a formidable problem, which is still open. This problem is particularly important since it typically imposes strong limitations on the control strategies that can be used in practice. Papers explicitly dealing with this issue are, e.g. Rodriguez et al. (1996b), Toyoda et al. (1995, 1997).

Finally, as already pointed out, the authors wish to recall that (to the best of their knowledge), in the recent literature there are very few works specifically dealing with the problem of modelling and controlling rotary sugar dryers. The work most closely related to the present paper is Douglas et al. (1993), where a first-principles model is proposed and validated, starting from the early works by Friedman and Marshall (1949) and Thorne and Kelly (1980).

7. Conclusions

In this paper the problem of fitting a set of data collected in June 1998 in the rotary sugar dryer at *CSR Ltd Plane Creek* by means of a recently developed grey-box model is considered.

The main results obtained here are:

- The model can accurately predict the output temperature if the moisture content predicted by the model is allowed to be exactly zero from one point of the drum on. This working-mode of the model has been called the “overdried-mode”, and is characterised by peculiar behaviour, considerably different from the “standard-mode”. The model’s ability to predicting an “overdried-mode” is extremely useful for developing a suitable control strategy.
- The model (both in “standard-mode” and in “overdried-mode”) provides useful guidelines for developing a sound control strategy, in the main working conditions of the sugar dryer.
- By exploiting a special feature of the gain from input moisture content to output sugar temperature, a very

simple control scheme was proposed, capable of keeping the dryer working on the edge between standard and overdried mode, using output sugar temperature measurements only.

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