THE PERFORMANCE OF A FLUIDIZED BED REFINED SUGAR DRYER

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Abstract

The performance of a fluidized bed drying system installed at Smith Sugar's Gledhow Mill to handle the production from the factory's back-end refinery is discussed. Effect on sugar quality, optimum operating conditions and residence time distributions are considered.

Introduction

Since the beginning of the 1977/78 Season, Gledhow Mill has operated a fluidized bed drying plant to dry the refined sugar produced in the factory's back-end refinery. The decision to consider this type of plant was motivated largely on the grounds of cost although the limitations imposed by the available sites for any proposed installation made its compactness and floor loading equally attractive.

A fluidized bed has many attractive properties for handling granular solids. The intimate contact between fluidizing media and the solids offers good mixing and heat transfer characteristics while the near absence of moving parts reduces maintenance costs substantially.

When considering the application of this type of equipment to the drying of sugar, more particularly refined sugar, two reservations are often expressed:

(i) the danger of crystal damage by mutual attrition of the particles.
(ii) the difficulty in fluidizing damp or wet sugar.

Theoretically, during fluidization each individual crystal is separately supported in the fluidizing air although a certain amount of contact can be expected to occur in the more violent aggregative phase, a condition resulting from the distribution of particle sizes normally encountered in real systems.

However, tests conducted on sugar fluidized for up to 10 times the expected residence time recorded no apparent damage. As crystal size characterisation proved to be impractical, a more direct measure of surface damage was attempted using the Pulp and Paper Industries' reflectance test. No discernible trend was detected, either visually or statistically, and all results were reported to be within the reproducibility of the method.

To ensure the fluidization of wet material, two methods may be used:

(i) Mechanical assistance
(ii) Back blending of dry material to lower the moisture of the input.

It was decided to opt for the second approach and to utilize the back blending characteristics inherent in a back mixed or perfectly stirred reactor design.

Description and performance of plant

The basic principles of fluidization are well known and will not be covered here while the main components of the system are shown in Fig. 1.

The main body of the drying unit consists of a vertical cylinder in which the upper section is separated from the lower by a perforated distributor plate through which the heated air is passed. Wet sugar from the centrifugals is fed to the upper compartment on a belt conveyor and distributed by means of a rotating spreader. The sugar forms a bed above the perforated distributor plate and freshly added feed is uniformly blended into this bed. In the fluidized condition, the sugar behaves like a liquid and flows out through the side of the dryer via a vibrating tray, the control mechanism of which is linked to the pressure drop monitored across the bed. Heated air is introduced, under pressure, into the lower compartment and passes upwards through the distributor plate and sugar bed. An exhaust air system withdraws the air and water vapour

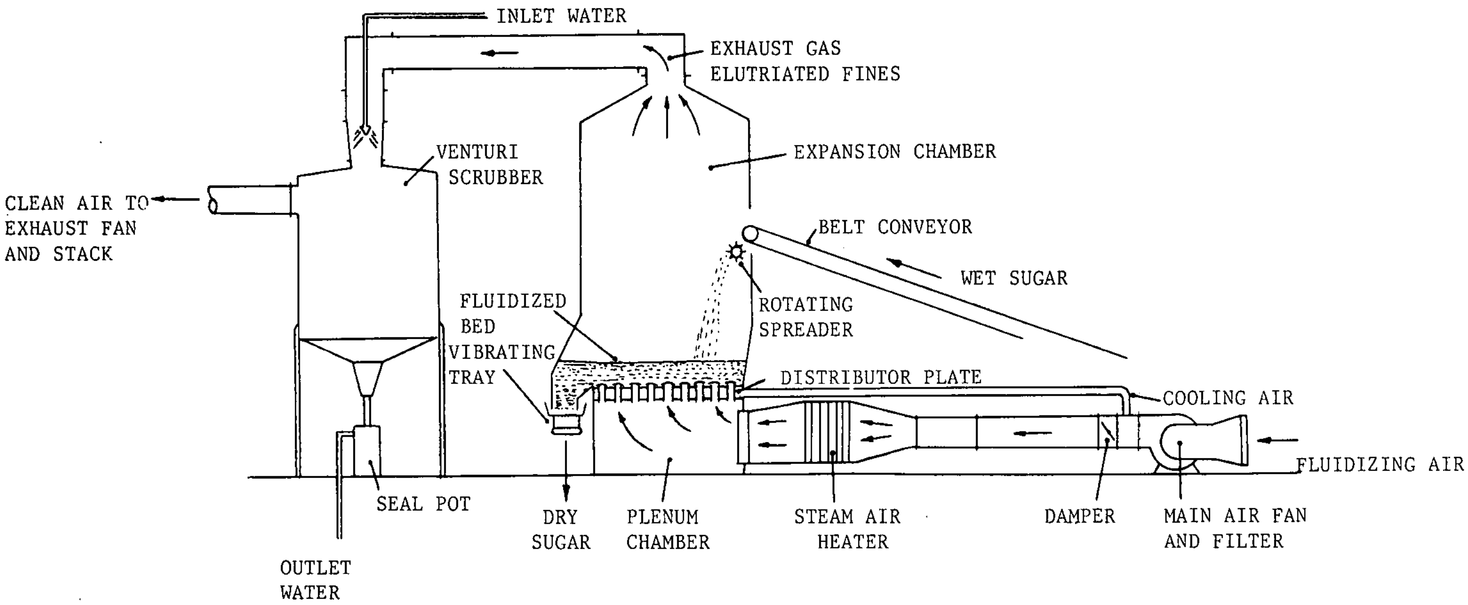

FIGURE 1 Schematic arrangement of fluidized bed sugar dryer.

from the top of the upper compartment and through a venturi scrubber, removing any fine crystals which may have been entrained.

Drying is controlled by maintaining the correct depth and temperature in the fluidized bed. As mentioned earlier, the former is controlled by a vibrating tray by means of a pressure drop based level controller, whilst the latter is achieved by regulating the steam flow to the air heater.

From the outset the installation confirmed the expected mechanical reliability but difficulties were experienced with the presence of lumps as well as with a variable outlet moisture content of the dried product. The problems associated with the lumps were largely eliminated when the new refined sugar centrifugal station was commissioned the following season, allowing a more suitable curing cycle to be practised on the machines.

The second problem however proved to be more tenuous and despite extensive investigations by all parties concerned with the installation, defied solution until the apparent cause was traced from the appearance of the damp sugar at the outlet from the dryer. At the required low level of outlet moisture content, any degree of by-passing could prove to be detrimental to performance and to overcome this problem it was considered necessary to introduce a baffle system to eliminate any short-circuiting. The success of this modification can be judged by the following results:

Average moisture content (oven dried)	
Before modification	0,085 % (w/w)
After modification	0,033 % (w/w)

The solution of this problem allowed a more meaningful assessment to be made of the unit. This took the form of 3 investigations which are considered below.

Residence time distribution tests

The main purpose behind the residence time distribution tests was to confirm the hypothesis that the variable moisture levels initially achieved were the result of by-passing of the damp sugar across the fluidized bed. By introducing a simple baffle arrangement — *cf* Fig. 2 — it was postulated that a finite minimum contact time would be imposed on each individual sugar crystal. This would have the effect of changing the shape of the distribution itself as opposed to altering the mean residence time, as no additional bed area had been introduced.

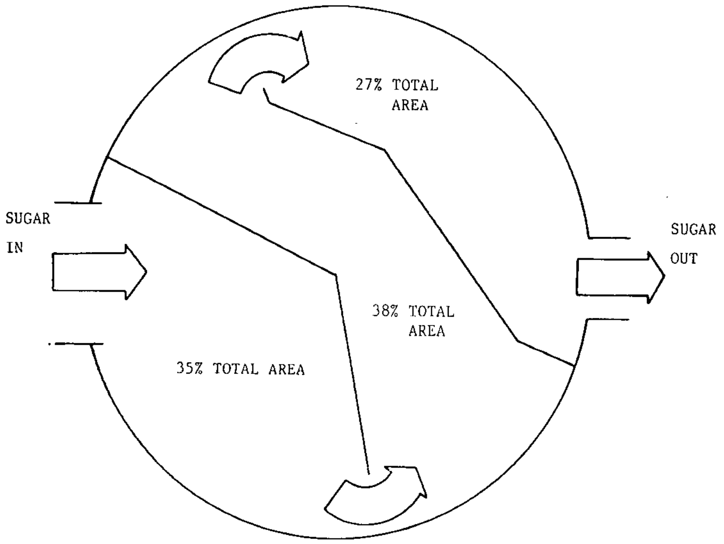

FIGURE 2 Plan view on baffle arrangement.

A series of pulse tracer response tests was conducted using 0,5 kg of fluorescein coated refined sugar as the input signal while the response was determined by measuring the optical absorbance at 485 nm on the outlet samples. It was found that at the concentration levels encountered, the attenuation index was directly proportional to the dye concentration which facilitated subsequent computations.

The results of these tests are presented graphically in Fig. 3 — without the baffle installed, and Fig. 4 — with the baffle in place.

A comparison of these figures confirmed that the introduction of the baffles had imposed a finite minimum time for each crystal in the dryer as is evidence by the reduction in the spread of the residence times which in turn lead to a more symmetrical distribution. Quantitatively, this improvement can be measured by means of one of the following methods:

(*a*) *Dispersion Model*[1] — in this model the degree of mixing is characterised by the magnitude of the dimensionless group D/uL.

(*b*) *Moments Analysis* — A statistical analysis of the spread and shape of the distribution. Any change may be characterised by either:

  (i) Mixing Index[2] *or*

  (ii) Measure of Skewness[3]

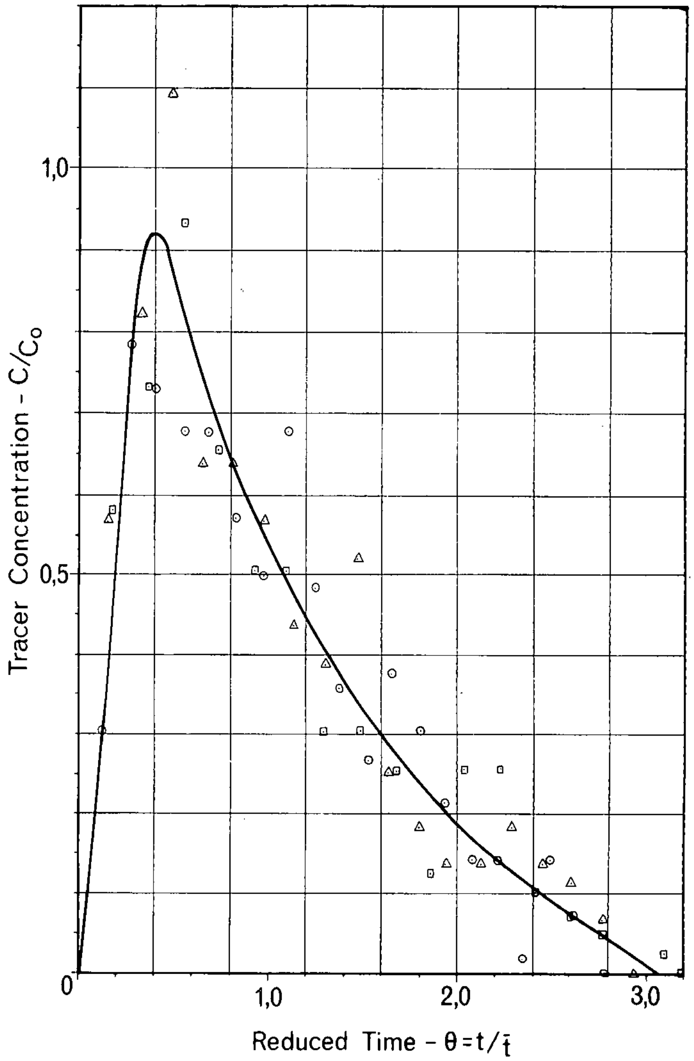

FIGURE 3 Response to a delta-function tracer input signal — no baffle installed.

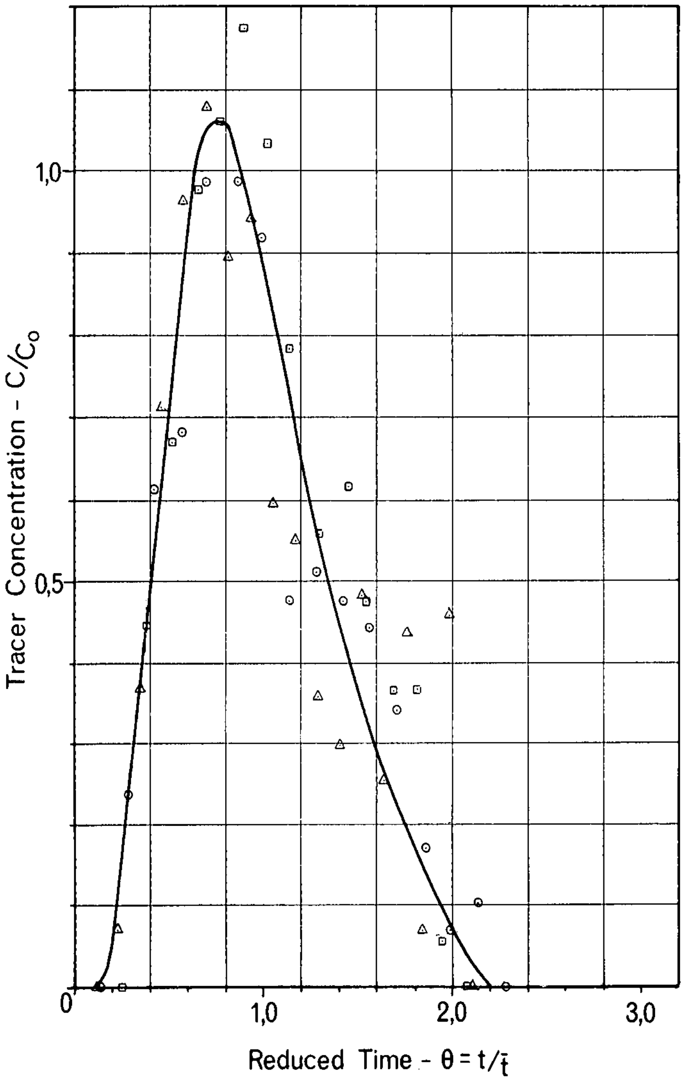

FIGURE 4 Response to a delta-function tracer input signal — baffle installed.

The results obtained using these methods are tabulated in Table 1.

TABLE 1

Residence time distribution analysis

	Perfectly mixed	Plug flow	Without baffle	With baffle
Dispersion No. D/uL	∞	0	0,22	0,09
Moment analysis—				
Mixing index	1	∞	36,73	45,81
Skewness	∞	0	36,56	6,72

It is evident from Table 1 that the introduction of the baffle appears to have increased the plug flow characteristic of the system and an attempt was made to obtain a more absolute measure of this increase.

For a plug flow vessel the response to a delta-function or pulse input is a similar pulse at the output T units of time later, while for a perfectly mixed or backmix flow vessel, the residence time distribution is represented by exponential decay. This latter response on integration gives the well known 'F' curve of the form:

$$F(t) = 1 - \exp(-t/T).$$

Wolf & Resnick[4] proposed a more generalised form of this equation for real systems:

$$F(t) = 1 - \exp - \eta \frac{(t - \varepsilon)}{T} \quad \text{for } t \geqq \varepsilon$$

$$F(t) = 0 \quad \text{for } 0 < t < \varepsilon$$

where η = measure of mixing efficiency

ε/T = system phase shift

T = mean residence time

Using the results obtained in the tests, the following values were obtained for the various coefficients:

No baffle installed $\eta = 1{,}2818$

$\varepsilon/T = 0{,}2525$

Baffle installed $\eta = 1{,}9280$

$\varepsilon/T = 0{,}3602$

With perfect mixing occurring when $\eta = 1$ and $\varepsilon/T = 0$ and plug flow when $\eta \rightarrow \infty$ and $\varepsilon/T \rightarrow 1$, the results confirm the substantial increase in the plug flow content by the installation of the baffles. Applying the specific model proposed by Wolf & Resnick for perfect mixing + plug flow in parallel, the percentage of the material in the vessel moving in plug flow through the bed increased from 21,1 % to 44,5 % following the introduction of the baffle.

Optimum operating conditions

In a fluidized bed drying system, the only system parameters that may be varied are:

Air velocity	(A)
Bed temperature	(B)
Bed depth (or residence time of the sugar in the dryer)	(C)

A full factorially designed experiment[5, 6] was carried out at 2 selected levels comprising 11 runs — 8 design and 3 replicates. The operating levels selected, being as wide as possible with due regard to both plant capability and dryer stability, are shown in Table 2.

TABLE 2

Fluidized bed dryer operating levels

Variable	Variable Designation	Operating level	
		High	Low
Fluidizing air flowrate (m^3/hr)	A	25 000	23 000
Bed temperature (°C)	B	80	50
Bed depth (mm)	C	350	250

The results of the above factorial experiment are summarised in Table 3 for a confidence level of 95 %.

TABLE 3

Factorial design experiment results

Variable	Response Sugar Moistures (%)	Confidence Interval at 95 %	Significance
A	0,060	– 0,00825 0,00525	Not significant
B	0,049	– 0,0120 0,00150	Not significant
AB	0,028	– 0,0075 0,00575	Not significant
C	0,030	– 0,00140 – 0,000500	Significant
AC	0,039	– 0,000250 0,0132	Not significant
BC	0,035	0,00150 0,0150	Significant
ABC	0,046	– 0,00525 0,00825	Not significant

Average inlet sugar conditions: Moisture 0,7 % w/w
Temperature 61° C.

From Table 3 it can be seen that only bed depth and the interaction between bed depth and temperature are shown to be of any significance. In addition, this table also indicates that in order to minimize the sugar moisture, the bed depth should be maintained at the high level and the bed temperature at the low level.

The non-significance of air velocity is assumed to be entirely due to the excess quantity of air required for fluidization.

It is however generally acknowledged that high temperatures promote rapid drying.[7] This aspect was investigated further and the results obtained, which are listed in Table 4, confirm the original findings. No reasonable explanation has yet been found for this phenomenon.

TABLE 4

Effect of bed temperature on outlet moisture

Temperature °C	Sugar Moisture (%)
50	0,032
55	0,045
60	0,030
70	0,088
80	0,086

Operating conditions: Bed depth 350 mm
Air flow 25 000 $m^3 hr^{-1}$
Average inlet sugar conditions: Moisture 0,87 % w/w
Temperature 62° C

A multi-linear regression on the significant parameters yielded the following relationship:

$$\text{Moisture } \% = 0{,}088 - 9{,}05 \times 10^{-5}(C) - 8{,}34 \times 10^{-7}(B \times C)$$

Operating the dryer at the optimum design condition of:

Bed depth	350 mm
Bed temperature	50° C

yielded a dried sugar moisture of 0,03 % as against a predicted moisture content of 0,04 %. The difference in performance can most probably be attributed to a marginal reduction in throughput during the test.

Dryer performance

The original specification for the dryer was to handle a throughput of 40 tph of refined sugar with a maximum discharge moisture of 0,05 %. To confirm the capability of the installation, a series of capacity trials were undertaken, the results of which are shown in Table 5.

TABLE 5

Fluidized bed dryer — capacity tests

Throughput (tons hr^{-1})	Sugar Moisture (%)
35	0,046
39	0,050
37	0,044
35	0,044
39	0,038

Operating conditions: Bed depth 350 mm
Air flow 25 000 $m^3 hr^{-1}$
Bed temperature 50° C
Average inlet sugar conditions: Moisture 0,85 % w/w
Temperature 60° C

As mentioned previously reservations have often been expressed regarding possible deterioration of refined sugar when handled in a fluidized bed system. Further tests were carried out on the actual plant to establish whether or not any measurable damage had been done to the final product, the results of which are presented in Table 6.

TABLE 6

Comparison of inlet and outlet sugar properties

	Mean Value		t — Value	
	Entering Dryer	Exiting Dryer	Calculated	At 95 % Confidence
S.G.S. (mm) ..	0,54	0,58	1,63	2,034
M.A. (mm) ..	0,64	0,66	1,13	2,034
C.V. (%)	43	41	1,37	2,034
Colour (420 nm)	98	106	0,77	2,064

Statistically there is no difference between the mean values recorded. Further confirmation is obtained by the low entrainment levels currently being recorded, which are running at approximately 0,1 % on the final product.

Conclusion

The resulting success of this installation proves that a fluidized bed drying system offers a highly attractive alternative to the dryers which are currently installed in the Industry. The system is compact, has few moving parts, requires only minor maintenance, can be fully automated and produces an undamaged refined sugar with a moisture content well below that currently specified.

It is proposed in the future to conduct tests on VHP sugar but there is little, if any, doubt that the required standard will be achieved. In addition, with the proper adjustment of the air velocities, dust and fines can be removed without mechanical screening which could have significant benefits in subsequent bulk handling.

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REFERENCES

1. Levenspiel, O. (1962). Chemical Reaction Engineering, Wiley and Sons, New York.
2. Otto, R. E. and Stout Jr., L. E. (1961). The moments method of analyses and its application to mixing dynamics, Chem Eng. Progr. *Symposium Ser.* **36** (57): 69-73.
3. Wetherill, G. B. (1972). Elementary Statistical Methods, Chapman and Hall, London.
4. Wolf, D. and Resnick, W. (1963). Residence time distribution in real systems, I. & E.C. Fundamentals 4 (2): 287-293.
5. Lionnet, G. R. E. and Baker, S. M. (1977). Experiment planning and the use of factorial designs. *SASTA Proc* **51**: 125-128.
6. Matthesius, G. A., Graham, W. S. and Pillay, J. V. (1978). An application of the fractional factorial experiment to continuous pan boiling. *SASTA Proc* **52**: 89-92.
7. Schwer, F. W. (1964). Some fundamentals in drying granulated sugar. *Int. Sug. J.* **23**: 115-129.