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AN EXPERIMENTAL PINEAPPLE JUICE CONCENTRATE PLANT INCORPORATING FLAVOUR RECOVERY. 2. INSTRUMENTATION AND CONTROL OF THE TURBULENT THIN FILM EVAPORATOR

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SUMMARY

The control of a Luwa turbulent thin film evaporator has been greatly improved by the installation of pneumatic steam temperature control devices, a vacuum steam supply, a cartesian manostat for vapour pressure control and positive delivery pumps for feeding juice and discharging concentrate. Appropriate recording instruments have also been installed.

I. INTRODUCTION

The suitability of Queensland-produced pineapple juice for concentration has been under investigation at this Laboratory for several years, using the plant described by Leverington and Morgan (1964). During this work the main difficulty experienced was loss of control in one or more of the variables affecting evaporation rate, namely steam temperature, product vapour pressure and product feed rate.

The effect of varying evaporation rates on the final product Brix is shown in Figure 1. It can be seen that variations in evaporation rates become increasingly important as the final concentration is increased. For example, in order to obtain 65% soluble solids in one pass, it is necessary to distill off 80% of the juice. An increase of only 4% in evaporation rate is sufficient to increase soluble solids to 80% while an increase of 7% in evaporation rate would completely dehydrate the juice.

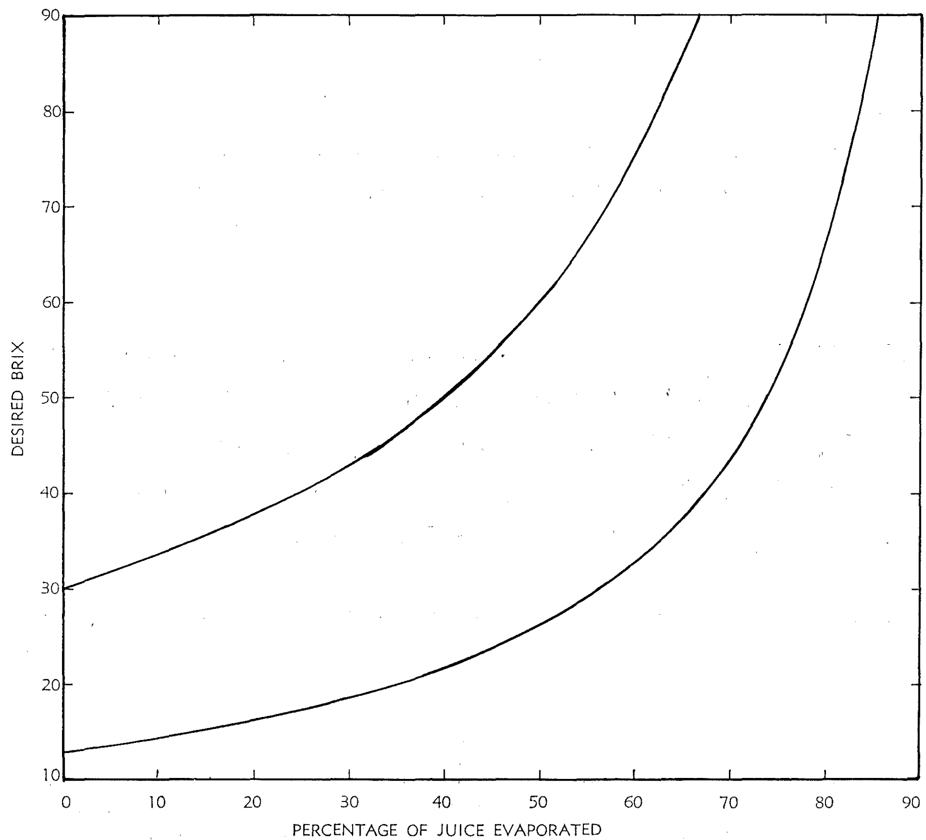

Fig. 1.—Effect of evaporation rate on final product Brix.

Methods used in industry to regulate the process of evaporation are:—

(1) keep flow rate constant and vary heat applied;

(2) keep heat constant and vary flow rate; and

(3) a combination of 1 and 2.

As most commercial evaporators monitor the concentration of liquid as it leaves the plant, the feasibility of installing this type of control was investigated. In-line refractometers as used in the sugar industry were considered, but due to the high viscosity and optical density of the concentrate, as well as the physical size of the units, they were unsuitable. The possibility of using continuous-weighing density meters such as the U-tube type was studied, but the size of the weighing tube was such that the juice would take some minutes to pass through it and therefore the time lag would be too great to obtain accurate control.

The use of the boiling point method of specific gravity measurement was also checked, but due to the confined space at the bottom of the evaporator and the difficulty of maintaining a constant level of liquid above the sensing element, this method was postponed until suitable miniature equipment was available.

In the absence of a suitable density controller and because the evaporation rate is mainly dependent on steam-jacket temperature, vapour pressure in the evaporator and juice feed rate, it was concluded that these variables would have to be controlled accurately. Modifications were therefore made to the plant to incorporate several automatic controlling and recording devices, bearing in mind that considerable versatility was required so that it would operate satisfactorily under a wide range of processing conditions. These new installations are the subject of this paper.

II. CONTROL SYSTEMS

A general view of the plant is shown in Figure 2, while the schematic diagram in Figure 3 illustrates the control systems used.

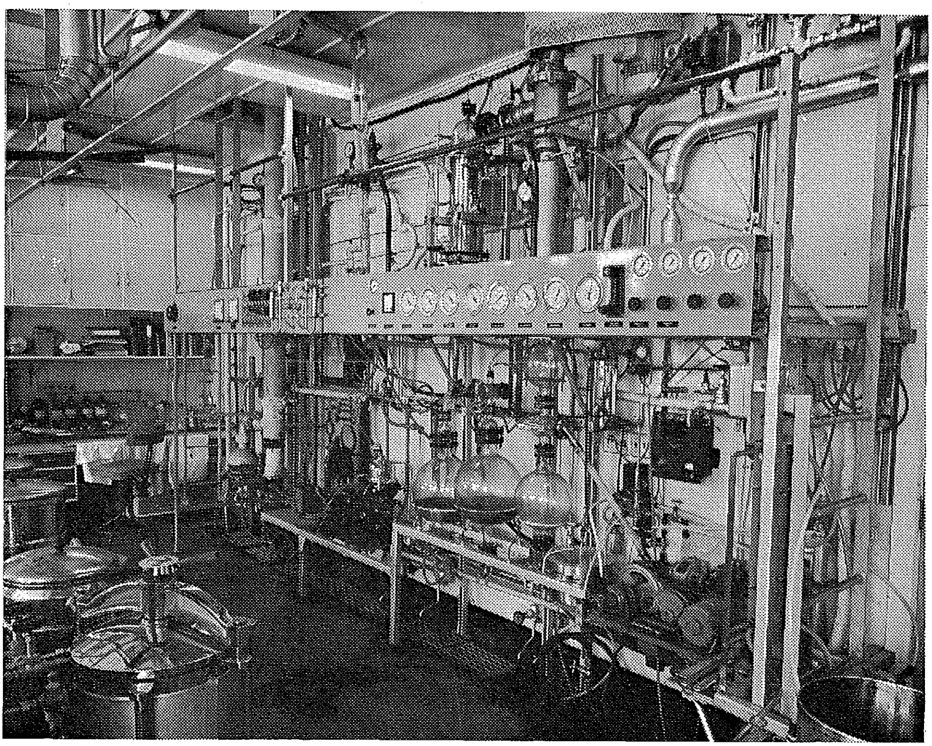

Fig. 2.—General view of the concentrate plant.

Automatic steam temperature control.—Although reasonable control was maintained when the jacket temperature was within the range 105–120°C and the product was evaporating at 60°C, the operation of the plant became more difficult the lower the evaporation temperature. This was assumed to be due (*a*) to the inability of the $\frac{1}{2}$-in. pilot-operated steam-reducing valve to control accurately the low steam pressure and (*b*) to the temperature differential between steam and product being excessive.

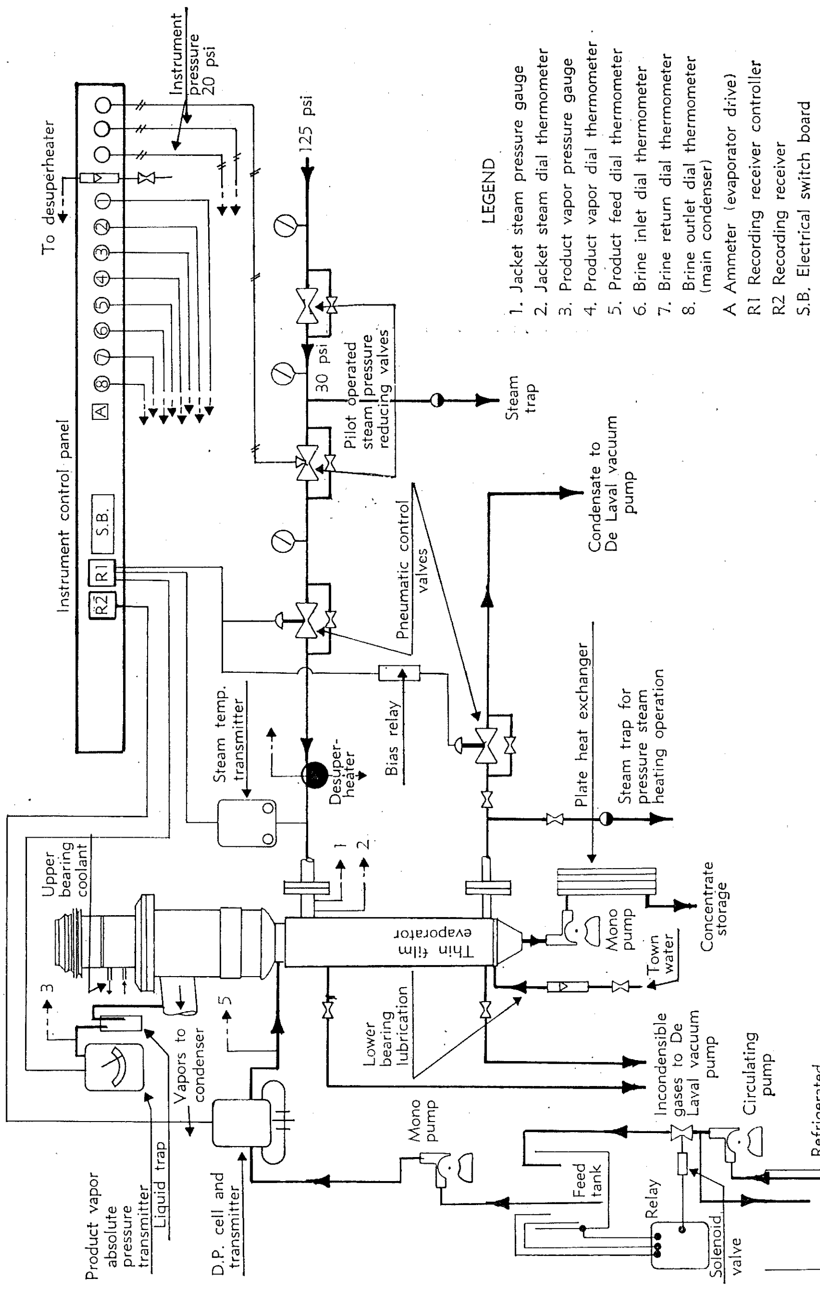

Fig. 3.—Diagram of control systems.

A new steam temperature control system was therefore designed and installed to overcome these difficulties. The temperature in the steam jacket is measured with a liquid-filled, fully compensated temperature sensing element which operates a Fisher and Porter temperature transmitter Model No. 1450, whose signals go to a Fisher and Porter 3-term pneumatic recorder controller Model No. 1212. The output from the controller passes to a 1-in. Fisher and Porter pneumatic control valve type 3560 V/P set in the main steam line. This signal also passes to a proportioning device which in turn controls the setting of another pneumatic control valve installed in the condensate line.

In order to reduce the temperature differential when required, a vacuum steam system was installed, utilizing an AD-3 de Laval vacuum pump connected with 2-in. lines to act as a vacuum reservoir and to avoid pressure drops along the line.

Because the evaporator was not fitted with non-condensable removal devices, two ¼-in. brass needle valves were fitted to the top and bottom of the steam jacket and coupled to one section of the de Laval vacuum pump. During operation the valves were cracked just sufficiently to allow connecting lines to warm up, thereby ensuring that all non-condensables were removed.

The steam supply was desuperheated by spraying metered water into a chamber in the steam line just ahead of the evaporator. The correct flow of water was obtained by regulating it until the steam temperature of the jacket was equivalent to the saturated steam temperature at the indicated pressure.

Vapour pressure control.—As pointed out by Leverington and Morgan (1964), the main vacuum system was controlled by a modified cartesian manostat and this has proved to be very reliable and capable of maintaining a pressure within $\pm 0 \cdot 5$ mm Hg under normal conditions. In order to have complete data on the pressure during operation, a line from the vapour tube of the evaporator was connected to a Fisher and Porter absolute pressure transmitter Model No. 1101, the output of which was fed to a Fisher and Porter strip chart recorder Model No. 1212. Subsequent recordings showed that sudden changes in pressure sometimes occurred during the connection of receivers due to slight differences in pressure between the two vacuum systems. A second cartesian manostat was therefore installed in the auxiliary vacuum line to ensure that the receiver pressures were exactly balanced with the evaporator before connecting the flasks back into operation.

Juice feed rate control.—As the throttling devices described by Leverington and Morgan (1964) were not entirely reliable due to occasional clogging of the fine orifices by small particles of pulp, attention was again turned towards non-pulsating positive delivery pumps. Of the small-capacity pumps tested, the Mono IBD 15 fitted with a variable-speed drive proved satisfactory but only after modifications had been made. These pumps normally operate with a positive pressure on the discharge side, but to maintain these conditions when pumping into a vacuum it was necessary to reverse the direction of rotation and connections of the pump. To assist further in steady flow, the pump was fed from a tank fitted with an electronic constant-level device. In order to have a record of feed rates throughout the processing periods, particularly as a check on pump performance, the juice feed was passed through a Fisher and Porter D.P. cell Model No. 10B1465, the output signal from which was fed into a pneumatic strip chart recorder Model No. 1202.

It should be noted that concentrate discharge and cooling were improved by the installation of a variable-speed IBD 15 Mono pump and a de Laval plate heat exchanger chilled by glycol brine. It was essential to use an oversize pump and to run it at very low speeds when handling viscous syrups, as recommended by Balanowski (1964).

Leverington and Morgan (1964) pointed out that foaming of the juice after centrifuging could cause fluctuations in feed rate. It was deduced that this was due to the use of centrifugal discharge centrifuges which actually aerated the product. This was not a mechanical problem when juice was canned at single strength. However, the frothing problems were completely overcome by using pressure discharge centrifuges in which no aeration of the product occurred by virtue of their principle of operation.

According to Tressler and Joslyn (1954), pineapple juice in Hawaii is concentrated in two stages, the esters being recovered from the first distillate only. Although the Luwa evaporator is a single-pass unit, the system was changed to a double-pass concentration because (*a*) by evaporating about 50% of the water in the fresh juice (giving a 26° Brix concentrate) and then evaporating the product to 65°–70° Brix the changes in the abovementioned variables were not so critical, and (*b*) the amount of distillate to be handled by the fractionating system was reduced to half, thereby almost doubling its capacity based on fresh juice feed rate and also assisting the production of higher fold essences.

III. OPERATING CONDITIONS

In a series of experiments in which the effect of evaporating temperature on juice quality was examined (the subject of a separate paper), the plant was adjusted to give evaporating temperatures of 32°C, 46°C and 61°C. The operating conditions are set out in Table 1.

TABLE 1

TYPICAL OPERATING CONDITIONS

—	Product Vapour Temperature (°C)		
	32	46	61
Average product vapour pressure ($\pm$ 0·5 mm) ..	30	95	190
Average steam temperature for 4–hr run ($\pm$ 0·5°C)	85	105	122
Average Brix of juice feed for 4–hr run	12·5	12·5	12·5
Average Brix of concentrate, 1st pass	30·4	30·9	28·8
Average Brix of concentrate, 2nd pass	67·6	68·2	69·1
Rotor speed (r.p.m.)	2,000	2,000	2,000
Average overall heat transfer coefficient (kcal/hr/m^2/°C)..	1,740–1,290	2,060–1,350	2,130–1,550

It was found most convenient to start up the plant using water rather than juice and, in general, operating conditions varied little for a number of runs at the same evaporating temperature. If the feed rate is kept constant and the Brix of the feed is known, it is a simple matter to calculate the distillate rate required to give any desired final Brix. In starting up, water is pumped into the evaporator at the same rate as the juice will be fed and the vacuum and steam pressure adjusted until the operating conditions are attained. Juice is then fed in and small adjustments made to the steam pressure until the product Brix is reached. Further minor adjustments are sometimes necessary at the lowest evaporation temperature to compensate for lower heat transfer rates being brought about by some scale build-up in the evaporator and for changes in ambient temperature.

IV. ACKNOWLEDGEMENTS

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