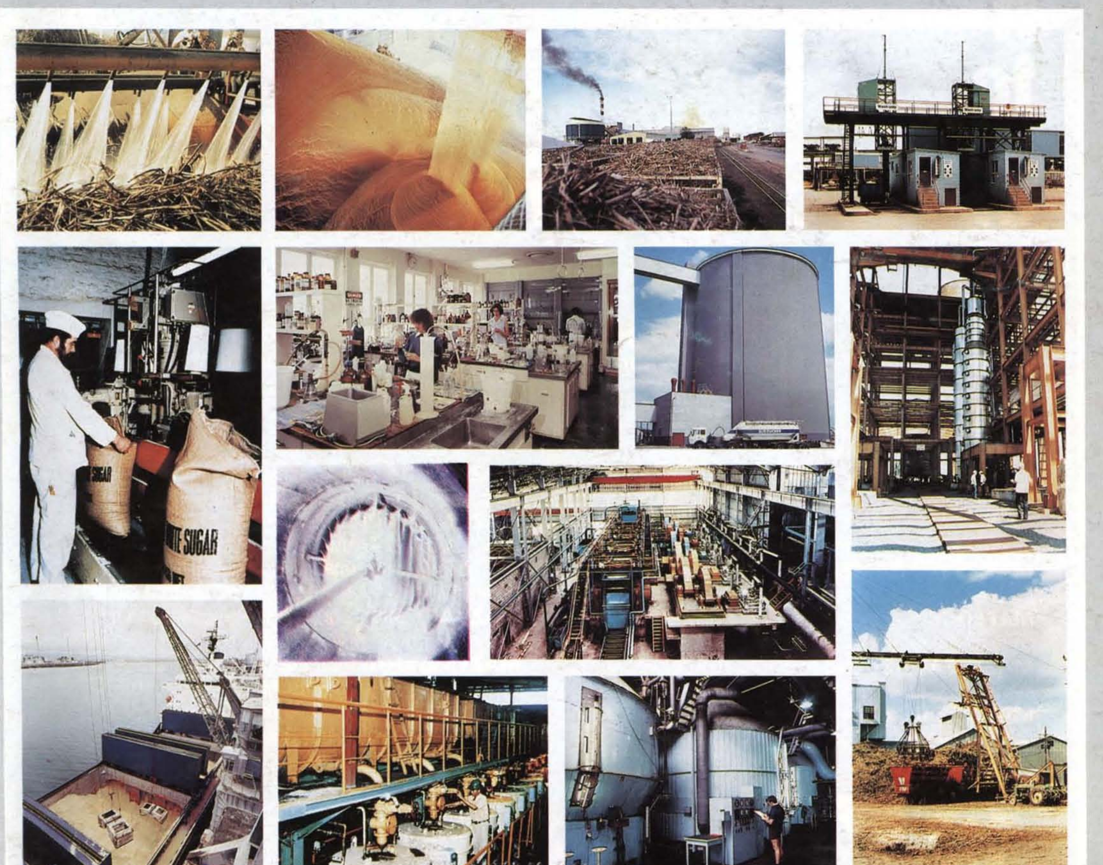
INTERNATIONAL SUGAR JOURNAL



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MAY 1988

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News and views

ISJ articles in other languages than English

We are occasionally asked for copies of articles we have abstracted but with the specification that they should be in a designated language, usually English. On other occasions, however, the language specified is another and we believe that there may be a demand for copies of our own articles, translated into, for example, Spanish or German. We are therefore making arrangements whereby translations into these two languages will be available and we invite inquiries for their supply. The tentative price will be US $30.00 per page for translation ito Spanish and DM 40.00 per page for translation into German, but a definitive quotation will be provided to interested inquirers.

We also hope to be able to provide translations of abstracted articles into required languages; the cost will depend on the charges made by the translator and this will depend on the languages concerned. Again, we will have to provide definitive quotations for individual inquiries.

International Society of Sugar Cane Technologists

A second newsletter has been published by the organizing committee of the 20th Congress of the ISSCT, to be held in Brazil in 1989. It provides detailed rules for the preparation of manuscripts of papers to be presented at the Congress as well as a guide for successful visual aids. Papers will only be accepted from Society members who will either be present at the Congress or have nominated a representative to deliver the paper. Copies of the newsletter may be obtained from the General Secretary-Treasurer, Dr. José Paulo Stupiello, Caixa Postal 532, Piracicaba, SP, Brazil 13400.

World sugar prices

Purchases of 300,000 tonnes of sugar by China and by traders thought to be supplying the USSR with replacements for Cuban shortfalls, as well as rumours of buying interest by other countries, had a firming effect on the world sugar market and the London Daily Price rose during the first half of March from $213.40 per tonne to $221, with an intermediate peak of $224.80 on March 8. A small setback followed, but the rise resumed and the LDP reached $237 by March 29 before falling back to $235.40 on March 31.

The white sugar premium started at $16.60 on March 1 but there was less strength in demand than for raw sugar and the premium fell quickly to about $11 for the first half of the month and then to around $6 during the second half. As a consequence the climb in the LDP(W) was smaller, from $230 to $241.50 after peaking at $242.50 on March 29.

EEC sugar balance

The total of A-quotas for sugar in the European Community for the period July 1987 - June 1988 amounts to 10,540,000 tonnes, white value, not too far from its consumption estimated at 10,750,000 tonnes by C. Czarnikow Ltd.[1], while the B- or "insurance" quota totals 2,288,000 tonnes. Not all the member countries produce their full A- plus B-quotas, however – in fact, the French Overseas Departments, Greece and Portugal do not produce their full A-quota entitlement – so that, even with sugar carried over from 1986/87, only 12,393,000 tonnes is calculated by Czarnikow to be within quota out of an available total (production plus carry-over) of 14,476,000 tonnes. Of the balance of 2,083,000 tonnes, the various member countries have announced that 1,260,000 tonnes in all will be carried over to 1988/89 so that 823,000 tonnes of C- or non-quota sugar will be available for export.

In addition, export availability will include the within-quota surplus of 1,643,000 tonnes, the ACP imports of 1,305,000 tonnes and other imports of 265,000 tonnes from East Germany, etc., but less the 190,000 tonnes of net exports in goods. Of the net total of 3,846,000 tonnes, the C-sugar amount of 823,000 tonnes will be sold without any subsidy, while the remainder will be subject to the various taxes and levies intended to recover the subsidy provided under EEC rules.

Czarnikow refers to the possibility that improved world market prices might have led to lower carry-forward tonnages so that producers could benefit from these prices; however, when translated into national currencies, the higher dollar prices have not proved sufficient inducement and the carry-over is virtually the same as from 1986/87. Further, Czarnikow considers that the levies are such that, with the exception of Greece, most countries are unlikely to expand beet areas.

World sugar balance

F. O. Licht GmbH recently published their second estimate of the world sugar balance for the period September 1987/August 1988 and the details appear below[2]. Production is now set at 104.2 million tonnes, raw value, which compares with the earlier estimate of 103.5 million tonnes and the previous season's total of 104.4 million tonnes. The higher figure is mainly the result of the excellent Soviet crop, which surpassed all expectations. Major uncertainty still surrounds production in China, India and Cuba; in the first, the area devoted to sugar crops has fallen and, while higher prices are to be paid, they will have little or no effect on the current crop. Assessment of the damage caused by drought and floods in different parts of India have varied widely, while the true effects this season of the 1987 drought in Cuba are difficult to gauge, especially with conflicting reports from the island.

Following an abnormally high growth rate in 1986/87, consumption is only expected to rise by not more than 0.6%, to 105.5 million tonnes; this is partly because demand has weakened in Brazil following the collapse of the Cruzado Plan. China's consumption increased markedly in the last season but rationing has been introduced and could effectively halt or even reverse growth.

1 *Czarnikow Sugar Review*, 1988, (1770), 23 - 24.
2 *Int. Sugar Rpt.*, 1988, **120**, 89 - 98.

Similarly, consumption growth in India, which has been high, is likely to fall, especially with a drop in disposable incomes where there has been drought. In the USSR retail sales are indicated to have returned to normal after recent fast growth, which also signals little or no growth in 1987/88.

Next year's outlook will depend on weather conditions; if these return to normal there should be sufficient additional supplies to prevent any further draw-down of stocks. But there is always the danger of a series of weather-induced production shortfalls. Licht notes that, for this eventuality, surplus stocks are no longer high enough to prevent a full boom. "This would be a disaster for the sucrose industry as a sharp price rally would only benefit alternative sweeteners such as HFS. The price depression over the past few years was a direct result of the high prices in 1973 and 1980, which made substitute sweeteners such as HFS more competitive. It is only natural that sugar producers and exporters welcome higher prices after such a long period of lean profits. But it would be dangerous to lose sight of the problems that high prices involve".

The balance is as follows:

	1987/88	1986/87
	---- *tonnes, raw value* ----	
Initial stocks	35,482,000	37,065,000
Production	104,212,000	104,392,000
Imports	27,402,000	28,174,000
	167,096,000	169,631,000
Consumption	105,530,000	104,907,000
Exports	27,993,000	29,242,000
Final stocks	33,573,000	35,482,000
% consumption	31.81	33.82

US sugar re-export plan dropped[3]

The United States has officially dropped plans for a program to benefit sugar exporters in the Caribbean and Philippines on the basis that the Department of Agriculture did not have the legal authority to implement it. The USDA officially notified Senator Inouye, author of the program, that it could not go ahead without additional action by Congress.

The program, passed as part of the Congressional 1988 budget package, was intended to increase sugar imports from the Caribbean and the Philippines, in compensation for the decline in their sugar quotas, and then to subsidize their re-export to the world market after conversion to white sugar by US refineries. The necessary funds for this subsidy were not provided by Congress, however, and the USDA has no authority to provide CCC money for aiding foreign farmers.

Caribbean exporters said they plan to seek the help of the Reagan Administration to help reverse the decision. This is unlikely, however, as the Administration does not wish to ease pressure on Congress to reform the price-support sugar program which is the cause of the reduction in access to the US market for the Caribbean and other sugar suppliers.

European beet area, 1988

Their first estimate of areas to be planted to beet in Europe was recently published by F. O. Licht GmbH[4]. It suggests that, unless exceptionally good weather raises yields above the average for the past six years, there is unlikely to be an increase in sugar production on the continent.

In spite of higher sugar prices on the world market, producers in the EEC are extremely cautious about expanding, particularly as they are required to pay a very high levy to wipe out, before the end of the 1990/91 season, losses on sugar exports. Although the last EEC crop was poor in some countries owing to bad weather, overall results were not bad and, if average yields are obtained in 1988, overall sugar production can be expected to decline.

No significant changes are expected in the rest of Western Europe, except Yugoslavia where the official forecast is for a reduction of area from 165,000 to a maximum 125,000 ha. A high beet crop is also forecast, however, and one or other of these figures may have to be adjusted.

The area sown to sugar beet in East Europe is unresponsive to changes in world market prices and the 1988 area is likely to remain at around last year's level, but with some small increases and decreases.

The key to 1988/89 production lies in the USSR, which had an extremely good crop last campaign. Whether the rise was due to structural change or if the good results were an aberration remains to be seen. In any case, the high level of output in 1987/88 is no argument for expanding the area sown this year as the industry would have difficulties processing more than last campaign. Thus, no increase is forecast and, since good harvesting conditions contributed to the good results so that a repeat of this performance may be difficult in 1988/89, a crop of 9 million tonnes would be no surprise.

The area forecasts are as follows:

	1988	1987
	------- *hectares* -------	
Belgium	117,000	111,000
Denmark	67,000	67,000
France	420,000	420,000
Germany, West	385,000	384,000
Greece	40,000	29,000
Holland	123,000	128,000
Ireland	33,000	36,0000
Italy	271,000	270,000
Portugal	1,000	1,000
Spain	195,000	182,000
UK	200,000	200,000
Total EEC	*1,852,000*	*1,828,000*
Austria	39,000	39,000
Finland	30,000	30,000
Sweden	51,000	52,000
Switzerland	15,000	15,000
Turkey	400,000	392,000
Yugoslavia	125,000	165,000
Total W. Europe	*2,512,000*	*2,521,000*
Albania	10,000	10,000
Bulgaria	50,000	48,000
Czechoslovakia	192,000	195,000
Germany, East	230,000	215,000
Hungary	101,000	108,000
Poland	440,000	428,000
Rumania	265,000	275,000
USSR	3,400,000	3,415,000
Total E. Europe	*4,688,000*	*4,694,000*
Total Europe	7,200,000	7,215,000

3 *Public Ledger's Commodity Week*, March 12, 1988.
4 *Int. Sugar Rpt.*, 1988, **120**, 123 - 127.

Sugar Industry Technologists Inc.

47th Annual Meeting, 1988

The 1988 Meeting of Sugar Industry Technologists Inc. is to take place during May 8 - 11 at the Hyatt Regency Hotel in Savannah, Georgia, USA. Members will gather during the week-end, will register on May 8, and will be entertained to a mixer cocktail party in the evening.

The first technical session will be held on the morning of Monday, May 9, when they will be welcomed by the President Leon A. Anheiser (Imperial Sugar Company). Under the chairmanship of James F. Fenske (Western States Machine Company), the technical papers will begin with a presentation by Brian T. Harrison (Imperial Sugar Company) on "Multiple hearth furnace – the years in review". This will be followed by a paper on "Safety instructions for sugar silos" by G. Dupire (Béghin-Say) and, after coffee, a presentation on "Optimizing filter aid filtration cycles" by Robert H. Rees (Manville Sales Corporation). "Demonstration of an advanced liquid treatment system – Amstar Sugar Corporation's reverse osmosis project" will be described by Dr. Chung Chi Chou and Henry C. Weber of the company, after which the election of the Board of Directors for 1988/89 will take place at the Annual General Meeting of SIT, and subsequently a reception and luncheon.

The second session is to be under the chairmanship of Léon Sué (Raffinerie Tirlemontoise) and will be in the form of a symposium on "Process control of sugar boiling" with four panelists: Marcel Braeckman (Raffinerie Tirlemontoise), François Defacoz (Abay S.A.), D. J. Radford (Tongaat-Hulett Sugar Ltd.) and Michael Donovan (Tate & Lyle Group Research and Development).

On Tuesday morning, May 10, Thomas E. Wilson (Colonial Sugars Company) will be in the chair for the third technical session which will be addressed by Charles D. Shamel who will describe how "Sugar turns the corner". Dr. Margaret A. Clarke (Sugar Processing Research Inc.) will then present a paper on "Chemical and microbial products from sucrose – recent developments", and "Experience of polystyrene and acrylic resins in dual-column decolorization systems" will be described by Leif Ramm-Schmidt, Erkki Talvitie, Matti Tylli and Erkki Luoma (Suomen Sokeri Oy.)

After coffee, Mary An Godshall and Margaret A. Clarke (Sugar Processing Research Inc.) will discuss "High molecular weight (HMW) colour in raw and refined sugars" and this will be followed by the description of a "New magnesia clarifying process for sugar refining" by Tadasi Nakamura, Hiromi Iwabe and Masayuki Kawakami (Mitsui Sugar Company). The final paper of the session will be an account of "Mist elimination in evaporators and pans" by Ronald D. Belden (Koch Engineering Company).

The fourth session, after luncheon, will be opened by a paper on "Production and refining of Hawaiian high pol sugars" by David M. Humm (California & Hawaiian Sugar Company) and this will be followed by a description of "Leucrose production and use" by D. Schwengers, H. Benecke and H. Giehring (Pfeifer & Langen). Richard Priester (Savannah Sugar Refinery) will discuss "Statistical process control – the new wave in quality assurance" while the last paper of the meeting will be presented by Bryan Tungland on the subject of "Analysis by ion chromatography in the sugar industry".

At the banquet to be held in the evening, the winners of the Meade Award for the best paper of 1987 will be presented with their plaques while the SIT Crystal Award for service to the sugar industry will be made. The final act of the evening will be the passing of the Presidential gavel from Mr. Anheiser

The new Executive Director of SIT

Bruce Foster was born and raised in Montreal, Quebec, Canada. In 1950, he graduated from McGill University in Mechanical Engineering.

He joined the St. Lawrence Sugar Company as a project engineer and, as the years passed, moved into management of the refinery. He remained with the Company, recently bought out by Lantic Sugar, until his retirement in 1987.

Bruce, and his wife Barbara, produced four fine children, two boys and two girls. As a family, they enjoyed many outdoor activities. Now the children are off making their own way in the world.

Bruce joined Sugar Industry Technologists, Inc. in the early sixties, became a member of the Executive Committee in 1979, and served as President in 1984. In November, 1987, he replaced R. Stuart Patterson, the retiring Executive Director of SIT.

to the new President, Frank Stowe (American Sugar Division of Amstar Corporation).

On Wednesday morning, a tour has been arranged of the Savannah sugar refinery, at the invitation of Joseph W. Alberino, whose wife has organized a ladies program which includes a tour of the city of Savannah, which was founded in 1733, and some of the Georgia hinterland.

Efficiency: the result of quality and productivity

An operational overview of Savannah Sugar Refinery

By Gil Clarke
(Savannah Foods & Industries Inc., Savannah, GA, USA)

Without some understanding of the background of our employees and company philosophy, it is hard to explain the progress that thas been made in the seventy years of Savannah Sugar. We have come from an average daily melt of less than 350 short tons in 1918, when employment was 400, to 2915 tons in 1987, with employment at 493. It would also be difficult to explain how production records are established almost routinely only to be broken within several months.

First, it starts with our employees. They come from proud families where hard work and dedication were and are a way of life. They are conservative, innovative, and determined. They have the desire to accept the challenge of improvement and to be a contributing part of a successful company. Without these qualities expenditure on technical improvements would be like throwing good money after bad.

The second factor in this equation is Savannah Sugar's philosophy about effciency. It is to make a product that meets customers' requirements, to operate at a capacity that uses energy, labour, and facilities to their best potential, and to maintain a technologically advancing process.

The Savannah Sugar Refinery is a raw cane sugar refinery that utilizes carbonatation, char/carbon decolorization, and a three-strike boiling system to produce a full range of consumer and industrial products. Melt capacity, with the plant in its present configuration, appears to be near 7.2 million pounds (3600 short tons) of raw sugar per 24 hours, with the current record melt being 6760 million pounds (3380 tons). During 1987, the plant melted in excess of 1,300,000,000 pounds (650,000 tons) of raws. Because of our coastal location and proximity to Florida, raws can be shipped by barge and rail from Florida, and in the last year, we have had vessel arrivals from Mexico, Peru, the Dominican Republic, Colombia, Brazil, Honduras, Belize, Costa Rica and Guatemala to supplement Florida's production.

Raw sugar arrivals are handled in one of two areas. Barge- and ship-borne raws are unloaded at a 300-foot wharf on the Savannah River by means of three contract cranes equipped with clamshell buckets handling 200 tons per hour each. The sugar is conveyed to an elevator which discharges to a 22,000 pound capacity Parsons scale. The scale weight is checked and recorded by two representatives of the seller and one from our plant. Twice daily, the scale is checked for accuracy: repeatability is within ± 2 pounds. Automatic composite sampling of each 22,000 pounds is performed for laboratory evaluation and price determination. From the scale, sugar is distributed to one of six warehouses or two bins. Sugar going into process can be ploughed off into "C" bin, or into "B" bin, the day bin. Sugar destined for warehouse storage can be ploughed off into No. 2, conveyed into No. 3, or trucked to Nos. 4, 8, 9 or 10. The second area for raw unloading, the Florida "cut-in", consists of four unloading spots for the simultaneous unloading of 110,000 to 157,000 pound railcars by front-end bucket payloaders every forty minutes. Although somewhat seasonal, rail shipment can average 25 to 30 cars per day.

Sugar is elevated and conveyed to a 22,000 pound capacity Parsons scale, where weights are checked and recorded by one representative of the seller and one from Savannah Sugar. Regardless of the means of shipment, the automatic composite is split into five identical samples, and sealed in air tight cans for the determination of the variables that enter into the pricing formula for raws. One can is sent to the seller, two to the New York Sugar Trade Laboratory, one to our laboratory for analysis and one is kept as a spare. Each composite represents 700,000 pounds of raw sugar purchased according to a contract based on 97° pol with either a positive or a negative factor applied to the price, proportional to the pol deviation from 97° pol. The customary closest two pol average is employed. Other factors entering the pricing formula are moisture, ash, colour, affined colour, dextrans, and grain size. Weighed and sampled Florida sugar can be conveyed to "C" bin, or other storage areas, and is normally used to refill "B" bin drained by the previous night's run.

Sugar from "B" bin enters the refining process via a belt controlled by five level indicators in electrical series. The conveyor discharges to an elevator (1st interlock) which carries the raws to a surge hopper (2nd interlock) above a 6400-pound capacity Parsons scale (3rd interlock). This scale is calibrated daily and the test weight is reproducible to within ± 2 pounds. An automatic sampler composites raws from each shift and with the scale weight, process accountability and performance are evaluated. As the raw sugar is discharged from the scale into the mingler, 72°Brix affination syrup at 60°C is dosed by timer into the mingler directly below the scale. The magma leaving the mingler is scrolled to a north and south mixer bank (4th interlock) that supplies twelve 48 × 36 inch Western States batch centrifugals.

Through the use of an ISSC-300 programmable controller, the sequence of centrifugal starts and intervals between them can be controlled, so as to assimilate the press floor liquor flow. This is accomplished by monitoring the high pressure "blow-up" level (Sweetland press supply tank) and assigning a 12-second machine start cycle to a 50% tank level. By maintaining a tank level of 50% or less, the affination centrifugals are at a minimum (144 seconds) cycle for all twelve machines. If the tank level increases by 5%, the interval between centrifugal starts increases by one second, slowing down the station. This may continue uintil the high pressure "blow-up" is full and the centrifugals are stopped, or it may balance the flow requirements of the press floor and continue to start centrifugals only at a slower rate. Within the 144-seconds machine cycle, each centrifugal's Westinghouse Numa Logic controller sequences the various functions of wash, load, brake, etc. The wash water timer control remains an operational variable, from 10 to 20 pounds per cycle depending on occluded ash and washed raw colour. Our target raw sugar colour is 5000 RBU (ICUMSA Mod.4) and the washed raw colour target is 1500 RBU. Target ash in raw sugar is 0.30% and that in washed raws 0.10%. These numbers represent experienced regression targets (derived by determining respective colour and ash removal efficiencies and applying them to the finished product analysis desired for that station). The washed raw sugar is melted to 70°Brix with high purity sweet water and heated to 70°C with exhaust steam and heat reclaimed in an *A*-liquor vapour heater in the evaporator system.

The low-grade system handles about 10% of the solids coming in as raw sugar. Besides the recovery of sucrose from affination syrup and sweetwater, Everglades Sugar Refinery's Florida syrup is also one of the syrups boiled in the four calandria pans using 10 psig exhaust steam. Two pans, the "C" (1300 cu.ft.) and the "D" (1440 cu.ft.), are

dedicated to boiling number one remelt massecuites. Each pan is charged with 55% of its strike volume in affination syrup or *C*-remelt syrup. A vacuum of 25 inches is applied in the pan until a 66°C temperature is reached, at which time the pans are seeded with one pound of extra fine grain (EFG) sugar in $^{1}/_{2}$ gallon of isopropanol, reduced for $4^{1}/_{2}$ hours in a ball mill. Vacuum is held for ten minutes without steam, and then boiling is resumed with sweetwater, tightening up with affination and 1st remelt syrup to a target Brix of 92 - 93°. Boiling time is about 3 hours.

The massecuites, of target purity 82 - 83, are cured in three 37-inch 30° Western States continuous centrifugals at a rate of 300 cu.ft./hr/machine. The centrifugals are equipped with syrup separators sending 1st remelt wash to the affination syrup supply tank and 1st remelt syrup to low-grade recovery. Centrifugal basket screens are of type "W" (0.0035 × 0.067 in). Water use on the feed rod is maintained at 0.5 gpm, and 1.75 gpm on the basket wash. Saturated steam also is used on basket wash to assist purging. The No. 1 remelt sugar is melted with high purity sweetwater to 65°Brix and mixed with melted washed raw sugar to form melt liquor going forward to the press floor.

The "A" pan (1440 cu.ft.) is dedicated to boiling No. 2 remelt massecuites which can be boiled in one of two ways. A strike to be cured has a massecuite target purity of 69 - 71; a seed massecuite, used as footings for three strikes, has a target purity of 72 - 73. A footing of affination and 1st remelt syrup, amounting to 44% of the pan capacity, is pulled into the pan by vacuum and boiled. Upon reaching a temperature of 74°C under $23^{1}/_{2}$ inches of vacuum, the pan is grained by the addition of one half-gallon of fondant slurry as in the "C" and "D" pans, after which the steam is similarly cut off for 10 minutes. The vacuum is elevated to 25 inches and 1st remelt syrup is used to feed the pan to capacity. The resulting Brix for a seed strike is 92° to 94°, and for a massecuite to centrifugals is 93° to 95°.

Pan boiling cycles are typically 3 to 4 hours, but may vary considerably according to demand and tank levels. Massecuites are cured in two 37-inch 30° Western States continuous centrifugals equipped with "W" screens, as are all of Savannah's continuous machines. Capacities are 160 cu.ft./hour/ machine and they typically require no feed rod or basket wash water, although halo and basket steam is used. The 2nd remelt syrup is returned to low-grade recovery for boiling in a 3-massecuite system.

The fourth pan in the recovery system – the "B" pan (1260 cu.ft. capacity) – boils solely No. 3 massecuites. Its footing of 44% of pan capacity can be pulled by vacuum from the seed mixer or from the "A" pan. It is boiled at 26 inches of vacuum, and fed 2nd remelt syrup to a target Brix of 94° to 95°. Boiling time is typically six hours. Massecuites are dropped to a crystallizer that is connected in series with seven more. As the massecuite moves through by gravity, heating/cooling coils maintain the temperature at 55°C, during a residence time of 40 hours.

Two 37-inch 30° Western States continuous centrifugals cure the massecuite at a rate of 90 cu.ft./hour/machine. Feed rod water is typically found to be 0.4 gpm, but is dependent on temperatures and Brix of massecuite leaving the crystallizer. Basket and halo steam are always used. Molasses apparent purity falls to 42 when we are able to boil target purities; by minimizing feed rod water, the differential between batch centrifugal molasses purities and continuous machine purities can be kept to about 3 purity points.

Sugars from the No.2 remelt and No.3 remelt machines are melted to 72°Brix and pumped to low-grade recovery as *C*-remelt syrup which is the footing of No.1 remelt massecuites. Melting these sugars, rather than handling them as a magma, has given a tremendous improvement in low-grade recovery. It has improved grain quality in all of the massecuites, which has resulted in better separation and demanded less feed rod and wash water, resulting in less remelt syrups and overall less recycling of impurities.

From the affination station, melt liquor is pumped to the clarification house where the removal of colloidal material, high molecular weight colour components and insoluble matter is accomplished by the precipitation and filtration of calcium carbonate. Clarification starts with a milk of lime slurry that is made by slaking $^{3}/_{8}$ - $^{3}/_{4}$-in diameter pebble lime with water to form calcium hydroxide. The reaction is exothermic and, by controlling the milk of lime temperature at 65°C by the addition of 45°C water, a milk of lime of constant 20°Brix specific gravity can be achieved. From the slaker, milk of lime overflows to the dosing tank which maintains a constant head pressure. The press floor programmable controller doses milk of lime at a preset interval into the distribution tank where melt liquor is mixed with the lime in a cyclical vortex and is distributed to one of three 10,900 gallon carbonatation cells in parallel. Normal sugars requires 0.6% CaO on melt solids. Carbon dioxide from either "B" boiler (gas fired stack CO_2 is about 8 - 9%) or from "D" boiler (coal fired stack CO_2 is about 12 - 13%) is cleaned of particulate and sulphur compounds in a Peabody scrubber. The CO_2 is then compressed in a water seal type rotary compressor which is paired with a carbonatation cell and is released into the bottom of the cell. The temperature of the three carbonatation cells is controlled at 75°C. An excess of lime has been introduced into the cells relative to the volume of gas reacting with it, and no pH control is sought in the three cells. However, the flow from all three cells converges in No. 4 cell, with which two compressors are paired, allowing further gassing to a target pH of 8.6 and temperature of 85°C.

From the 4th carbonatation cell, liquor flows to the high pressure "blowups" in preparation for filtration. There are thirteen Sweetland presses that are operated by two ISSC-90 programmable controllers in a 77-minute cycle, and sequenced by an ISSC-300 controller which also communicates with the affination controller. The press floor ISSC-300 monitors the level in tanks 84 and 85, which are the char house supply tanks, and controls the intervals between presses coming on line. As the level in tanks 84 and 85 drops, the interval is decreased between press starts which initiates the sequence in the ISSC-90 press controllers. A typical press cycle starts with precoating from the low pressure "blow-up" with 50 pounds of kieselghur (3.3 minutes), clear to the excess tank (3 minutes), high pressure "blow up"-liquor forward (60 minutes) sluice to the mud tank (0.75 minutes), and drain (5 minutes). Generally presses are on line with a combined flow rate of 700 gpm being sent to the char house. In the formation and filtration of calcium carbonate, approximately 50% of the colour, 20% of the ash, and almost all of the insolubles are removed. Target colour and ash are now 900 RBU and 0.10%, respectively. The major by-product of the carbonate reaction, and sometimes a difficult material to handle, is the 95,000 pounds of press cake produced daily. This material is sluiced to the mud tank from the Sweetland presses at 15% solids and 10 - 12°Brix sucrose. Lime is dosed to the mud tank in the amount of 0.2% CaO on melt; the slurry is then pumped to the saturator where 10 psig CO_2 gas further precipitates $CaCO_3$ at 10.5 pH and distributes the mud to one of three Eimco rotary drum filters for dewatering. Sweetwater is extracted leaving a dry filter-cake on the polypropylene blanket which is slurried with char wash water and pumped to our oxidation pond for future disposal. The sweetwater is clarified and sent to the granulator wet dust collectors, and thence to the affination station as dilution water for melting washed raw sugar.

From tanks 84 and 85, SP liquor is pumped to a constant head tank so that 12 - 15 psig can be maintained on the char house filters. There are 42 bone char filters, each holding 32 tons of char and 5000 gallons of 65°Brix liquor. Of the forty-two, thirty filters are on liquor at a 25 gpm flow rate to maintain the 700 gpm process flow. The remainder are being sweetened off, washed, being filled or being emptied. The total cycle time is about 140 hours of which the "on liquor" duration is 100 hours. After sweetening off to 0.37 pol and washing, char is conveyed to a 7 hearth Nichols-Herreshoff kiln. Char feed problems caused by high moisture content in the louvre section of the kiln have been eliminated by using three feed screws to maintain constant flow from the storage hopper to the first hearth. Char temperatures reach 1000°F in the seventh hearth where conditions are maintained at 1750°F and less than 3% excess oxygen to minimize fires within the char. Two coolers, below the kiln, cool the char to 50°C before it returns to service. Constant screening and a density table remove the exhausted portions of char. The system requires the cut-in of about 6000 pounds of char per day. Char liquor colour and ash have now reached 120 and 0.08%, respectively. From the liquor gallery, the best ten filters are selected to go forward as true char liquor (60 colour and 0.08% ash) whereas the remaining 20 filters are diverted to carbon decolorization.

The carbon system consists of seven 1500 cu.ft. stainless steel filters. Each filter can handle 80 gpm but, because of the revivification cycle, normally only six are on line. A typical liquor cycle may last 450 hours, with 400 hours "on liquor". As the decolorization effciency drops off, a time frame must be selected where additional sweetwater is not a problem and 72 hours can be arranged for the carbon kiln to be at operating temperature. Normally at the beginning of a week, the filter is sweetened-off and the carbon is transported hydraulically to the spent tank. An inclined scroll dewaters and transports the carbon to a five-hearth Nichols-Herreshoff kiln where the carbon is brought up to 1750°F with no excess O_2 control. The carbon is quenched by the water that will transfer it to the regenerated carbon tank and further to the carbon filter. The filter is drained of water and liquor is started. Liquor from the six carbon filters is then mixed with liquor from the ten best char filters. The carbon decolorization efficiency of 50 to 60% with no further ash removal, has now put us in a position to boil acceptable sugar. To ensure that no particulates accompany the *A*-liquor, the combined stream is filtered in either of two Sweetland polishing liquor filters; a third stands ready but idle. From the filters, liquor is pumped through a Dedert quadruple-effect falling-film evaporator. Exhaust steam at 10 psig, entering No. 1 evaporator, encounters a counter-current liquor flow from No. 3 evaporator to 4 to 2 to 1. Steam vapour from the liquor side of No. 1 heats No. 2 evaporator; its vapour heats No. 3 and No. 3's vapour heats No. 4. No. 4 evaporator's vapour is condensed in a barometric condenser to create 26 inches of vacuum for the system. The efficiency of the evaporator system is 2.8, i.e. one pound of steam evaporates 2.8 pounds of water. A Brix increase from 65° to 77° at 180°F is typical with no measurable colour increase.

From the evaporators, *A*-liquor is pumped to the white sugar pan floor (and used very quickly) where there are four pans. The "E" pan is a 2400 cubic foot ribbon type pan boiling on 35 psig steam. It is charged with 1133 cu.ft. of *A*-liquor, grained at 25 inches vacuum and is fed *A*-liquor at 22 to 23 inches vacuum. Its boiling cycle of 75 - 80 minutes produces a strike at 92°Brix with a 55% yield. Six 48 × 36 inch Western States batch centrifugals, using a 4-seconds water wash, are dedicated to the "E" pan massecuites. *A*-Sugar colour is typically 8 - 10 RBU.

The "F" pan is an 1800 cu.ft. ribbon type pan also boiling on 35 psig steam. It is charged with 695 cu.ft. of "A" liquor, grained at 25 inches of vacuum and fed a combination of 1st and 2nd syrup. Its boiling cycle of 80 minutes produces a strike yielding 53% sucrose. Nine 48 × 36 inch Western States batch centrifugals, using a 4 seconds water wash, are dedicated to the mixer that is common to the "F" pan and "H" pan massecuites. This sugar is typically of 14 - 15 RBU colour.

The "H" pan installation is the

latest technological improvement made in the plant. It is an all stainless steel 2400 cu.ft. calandria-type pan, boiling on 10 psig exhaust steam. The calandria is constructed with 1700 tubes of $3^1/_2$ inches diameter by 42 inches long, resulting in a heating surface to pan volume ratio of 2.16 to 1. It is equipped with a 72-inch diameter nautical-type propellor and a 150 h.p. motor. It is charged with 1000 cu.ft. of *A*-liquor, grained at 25 inches vacuum, and fed on a combination of 1st and 2nd syrup at 22 to 23 inches of vacuum. Basically, except for pan size difference, the "F" and "H" pans are boiled identically. Wash water and sugar colours are identical, but boiling time in the "H" pan may average 90 minutes, which includes steaming-out and lubrication times. By installing the "H" pan our electrical-steam load has become more balanced; there are two more pans boiling *A*-liquor, and less tankage is required, owing to the constant back-boiling of syrups. Because of its timely installation, and the degree of efficiency it has brought to the process, the "H" pan has been appropriately nicknamed "the Saturday pan", as it has all but eliminated Saturday operations and overtime along with it.

The fourth "G" pan is of 2160 cu.ft capacity and has a concentric ribbon heating surface. it is boiled on 35 psig steam in about 90 -100 minutes. A "G" pan charge is a combination of 1st and 2nd syrup. It is grained at 25 inches and fed 3rd syrup at 23 inches of vacuum. This pan is constantly back-boiling 3rd syrup which results in a gradual increase in sugar colour from 20 RBU, increasing 10 RBU per strike, to an 80 RBU level which is the "kickout" sugar colour. Generally speaking, we expect to get nine cycles before colour becomes unacceptable. Six 48 × 36 inch Western States batch centrifugals are dedicated to the "G" pan, and wash water is incremental from 8 seconds to 20 seconds.

Even though the pan floor is not integrated into the control scheme with the char house, press floor, and affination station, that is not to say it is not automated. On the contrary, it is fully automated, involving three ISSC-90's and one ISSC-300 controller communicating with pneumatic and electronic instrumentation. It is sophisticated to the point that a controller can completely boil and drop a pan with the only input being the graining and feed vacuum desired. Ours are typical sugar boilers, however, in that they almost always want to be able to operate at least one manual control during the cycle.

From the centrifugals, sugar is conveyed to one of three granulator systems. The "F" and "H" sugars are dried in one pair of heating and cooling granulator systems of the Standard design. Both are 8.5 feet in diameter, with sugar passage counter-current to air flow. The upper or heating granulator has a drying temperature of 90°C, whereas the lower or cooling unit uses ambient air. The "G" granulators are an identical system to the "F" and "H" granulators except that the upper temperature is controlled at 80°C. The "E" sugars are dried in three 6-foot Hersey-type granulators at 100°C and cooled in three of the same design using ambient air. All extra fine grain (EFG) sugar, with targets of 28 - 30 RBU colour and 0.04% moisture, is conveyed to No.2 or No.3 silo. Each of the three silos has a 6 million pound capacity and sugar can be withdrawn as packaging demands. Only one silo is used for bulk sugar deliveries, which requires that it be reserved for sugar that has been screened and conveyed around the packaging loop. This loop is designed to handle 200 tons per hour and supply screened sugar to the packing house.

The packing house handled almost 70% of the products shipped in 1987, bulk and liquid sugar accounting for the remainder. The installation of three new industrial bag packing machines and automated palletizers has streamlined the 100-pound EFG, 100-pound specialty sugar and the 50- and 25-pound EFG bag operations so as to allow any combination of desired products to be run simultaneously. The 100-lb Consolidated Bag Pak line packages specialty sugar at 17 bags per minute (BPM), as does the Bemis PBOM bagger using EFG sugar. Both the 50-lb and 25-lb machines are Bemis PBOM baggers with capacities of 17 BPM and 22 BPM. To continue in the granulated area, 5-lb bags are produced on nine Parsons scales with 48 BPM capacity each.

Parsons scales 1 - 3, 4 - 6 and 7 - 9 are paired with an International Paper/Griswold bundler producing twelve unprinted bales per minute. Brand and code dates are applied by a non-contact dot matrix coder. Production of 2-lb bags is handled by three Parsons scales operating at 48 BPM feeding a standard Knapp packer, while filling of 10-lb bags is by two Parsons scales operating at 32 BPM and paired to an I.P./Griswold bundler also equipped with a non-contact coder. The economy of using unprinted stock becomes apparent when it is realised that fifty-seven different 5-lb labels and nine 10-lb labels must be produced.

Powdered sugar production has been on the increase over the past several years, demanding that the two Mikro atomizers and six Mikro pulverizers operate at full capacity. Production of 100-lb bags requires four pulverizers to pack 3 BPM; that of 50-lb bags requires one pulverizer to produce 1.5 BPM and the 2-lb polyethylene bag line uses one pulverizer to supply the form-fill seal Rovema/Mateer Burt filler running at 24 bags per minute. All of our powdered sugar is made from granulated sugar combined with 3% corn starch by a K-Tron volumetric feeder.

Production of 100-lb, 50-lb and 25-lb soft sugar packs is handled through a manually-operated Thayer scale. The rate for 100-lb bags is one bag per minute. One pound carton production is packaged on four automated Rovema form-fill seal machines with Schindler scales, pneumatic side seamer and a Rovema cartoner, at a rate of 120 cartons per minute. Production of 2-lb polyethylene bags of soft sugar is also handled by a Rovema form-fill sealer/Schindler scale running at 18 bags per minute.

The white sugar warehouse is home for six automatic and two semi-automatic palletizers and 100,000 square feet of product storage. Two FMC

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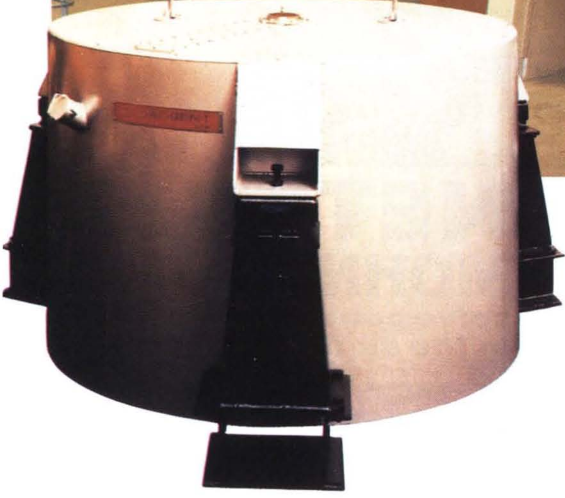

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palletizers are constantly running 12/5 bales. One FMC can alternate between 12/5 bales, 6/10 bales and 30/2 bales. All of the cases are palletized on a Litton/Von Gal, 5 accumulating lane automatic palletizer. The 100-pound powder and specialty 100-pound sugars are each stacked on a Shields semi-automatic Air-Float palletizer. Two Litton/Von Gal palletizers automatically stack 50-lb and 100-lb bags of EFG sugar. Even with all of this capacity we are in need of more floor space and we are at present expanding storage capacity to 132,000 square feet with 17 loading spots.

Sugar that has been transferred between silos is conveyed at 80 tons/hr to the bulk system which consists of eight epoxy-coated 300,000-pounds capacity bins. Each bin is purged with Kathabar-dried air, filled and held for a prescribed period of time. A Parsons 500-pound scale handles the rail car loading requirements whereas a Wallace & Tiernan weigh belt feeder is used for loading of trucks. The rail system loading rate is 60 tons/hour, handling a maximum of 8 cars per day. The truck system loads at 20 tons per hour, handling 8 - 10 trucks. Both can operate simultaneously.

Liquid sugar is currently manufactured as liquid sucrose and invert. Sucrose is manufactured from melted granulated sugar, carbon-treated, filtered and flash-cooled. Invert is manufactured from melted granulated sugar, 50% inverted with hydrochloric acid, neutralized, treated with carbon, filtered and cooled.

One of the largest projects in recent years was the installation of the Babcock and Wilcox coal-fired boiler and peripherals at a cost of nearly $13 million. It has the capability of producing 250,000 lb of steam per hour at 300 psig, firing coal, gas or oil, with coal being the most cost-effective at present. It has three compartments that can be independently controlled by a Bailey Network 90 computer equipped with an Operator Interface Unit. The boiler burns 200 tons per day of compliance coal from either Kentucky or Virginia (0.75% sulphur and 13,000 BTU/lb) which is delivered in 100-ton rail cars. Upon arrival, these can be unloaded onto the pile in about one hour. Coal can then be trucked to the coal bunkers within the boiler room in 24-ton trailers. Ten deliveries per day are necessary when the boiler is on line. The three bunkers hold 127 tons of coal each and supply the three pulverizers and compartment burners. The disposal of 30 tons per day of fly and bottom ash is performed on alternate days by a contract hauler who is also handling the disposal.

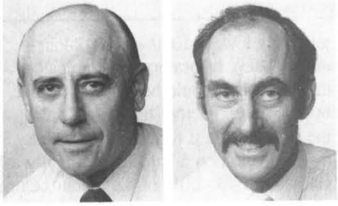
At the same time that the new boiler was installed, a 5000 kW turbo-generator was installed, creating the balance needed between the electrical demand of the new boiler and the desired supply of low-pressure steam to operate the "H" white sugar pan. It brought the electrical generating capacity to 11,700 kW. This balance has worked out very well as we seldom are required to "cut over" 300 psig to 15 psig steam without going through a turbine. As a back-up to the new boiler and for additional steam and flexibility, three older boilers are maintained and kept on standby status. Even though this is just one isolated example of our operational flexibility it serves as a good point with which to end. For the word efficiency seems synonymous with flexibility. Man's very nature is to improve, be it for himself, the world around him or a world for his children. Our dedication to quality and productivity gives us, we hope, the opportunity to continue to improve and be here for the next seventy years and beyond.

High pol sugar refining using continuous affination

By Geoffrey E. Mitchell and Cameron G. Smith
(Millaquin Sugar Company Pty. Ltd., Bundaberg, Australia)

Introduction

The Millaquin complex consists of a raw sugar factory, refinery and distillery located on the Burnett River near the regional city of Bundaberg. The refinery was built first (in 1881), and produced white sugar from limed cane juice pumped through a cast iron pipe network to the factory from small crushing mills.



G. E. Mitchell C. G. Smith

Millaquin Refinery is a small facility processing some 30,000 tonnes of raw sugar annually. Located next to a sugar factory, it has traditionally been supplied with a high pol raw sugar. As a result the refinery has a lower than usual impurity loading.

A common boiler station provides steam to the three units of the complex during the crushing season, using bagasse as fuel. In the slack season, the refinery and distillery continue to be supplied with steam produced from a range of alternative fuels. Power is generated on site during the crushing

Paper presented to 46th Meeting, Sugar Industry Technologists, 1987.

season and is purchased from the regional electricity board in the slack season.

All granulated sugar produced is delivered either in 30 kg industrial packs or 1 kg and 2 kg domestic packs. Soft sugar and golden syrup is also manufactured. The Millaquin refinery does not supply significant tonnages to the bulk granulated or liquid sugar markets as these have been largely met by a capital city refinery.

Refining process (1978)

The refinery steps in use at Millaquin in 1978 are shown in Figure 1. Phosphatation was carried out in a circular multicell clarifier, of 2.1 m diameter and 2.7 m high, using three chambers mounted vertically. Scums from the clarifier were handled on two Sweetland leaf filters, each 93 m² (1000 ft²) in filtering area.

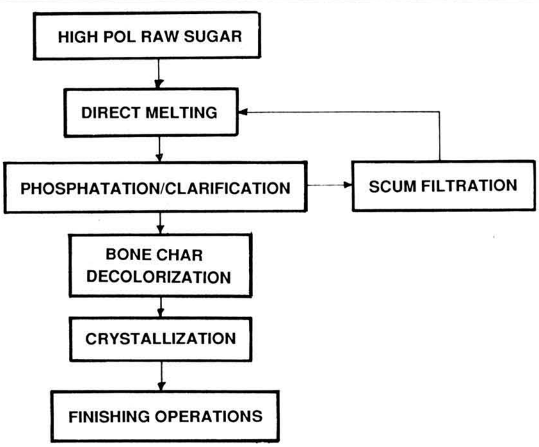


Figure 1. Refining process, 1978

A bone mill on site operated during the mill slack season to provide char for decolorization. The char house comprised 28 cast iron cisterns of 16 m³ capacity and four Scotch kilns. Two kilns met normal regeneration requirements and these were fired with fuel oil.

Polish filtration

As is often the case with bone char following clarification, carryover from phosphatation restricted liquor flow through the cisterns. Polish filtration of clarified liquor commenced in 1979 using one of the Sweetland filters previously dedicated to scum handling. This filter became available following work which demonstrated that clarifier scum could be filtered at reasonable rates after dilution to 20°Brix.

Liquor cycles were initially 8 - 10 hours. These cycles improved later, to average 20 hours, with the addition of polyacrylamide at the clarifier. Diatomite filter aid was required for both precoat and body feed. Flow rates through char increased dramatically as a result of polish filtration and over time this improvement led to a reduction in char usage from 18 to 12% on melt.

Affination

Even though the colour of the incoming high pol raw sugar is relatively low, at 1000 - 1200 colour units (RBU), laboratory tests indicated that the colour could be reduced further by affination. Factory trials confirmed the laboratory test results and showed that affination on a batch centrifugal was viable with 40 - 50% of the colour being readily removed. Following the trials, an affination system was designed and built utilizing an existing batch centrifugal. This plant was commissioned in 1983.

Affination of the high pol raw sugar provides a surprisingly high level of benefit in respect of lower melter liquor colour and lower impurity levels in the main course, typical of affination of a normal raw sugar.

Phosphatation

Following the success of affination, the multicell phosphatation clarifier was bypassed in 1984. The effect of this change on filtered raw sugar liquor (F.R.L.) colours and a comparison with colour prior to affination is given in Table I.

Table I. Comparison of F.R.L. colours for various operating systems

Process characteristics	F.R.L. colours (RBU units @ pH 7)
(Raw sugar)	(1100)
Direct melting/ Clarification (prior to affination)	700
Affination/ Clarification	410
Affination only	590

It can be seen that, while the absence of clarification resulted in an increase of some 180 units in F.R.L. colour, this level is lower by a similar amount than the colours experienced initially when direct melting and clarification were in use.

The main feature of the operation without clarification is that the chemical requirements for the refining process are now negligible and only a small quantity of lime continues to be used for pH control.

Canesorb

The future of the char house had been under consideration for some time owing to the age of the cisterns and the state of the wooden building. In 1984, after several alternative decolorizing media had been assessed, a commitment was made to build new char cisterns and at the same time convert to the Canesorb blend for liquor decolorization. This change came into effect in mid-1985 with the installation of four new adsorbers, each 28 m³ in volume.

Introduction of Canesorb effectively

ISJ Abstracts

Cane sugar manufacture

Evaluation of evaporator scaling inhibitors

D. Hsu. *Ann. Rpt. Hawaiian Sugar Planters' Assoc.*, 1986, 46 - 48.

Tests in 1986, in which Mazu Evap 711 (claimed by the manufacturers to control the formation of organic scale) was injected into the juice entering a pre-evaporator cell and Mazvap 900 (intended to prevent inorganic scale formation) was introduced into the juice entering the 4th effect of a quadruple-effect evaporator, showed that neither agent significantly inhibited scale formation nor raised the overall heat transfer coefficients, confirming unfavourable results obtained in the previous year.

Deterioration of standing mixed juice

K. M. Onna and R. R. Tamaye. *Ann. Rpt. Hawaiian Sugar Planters' Assoc.*, 1986, 48 - 49.

Investigations of the behaviour of mixed juice samples stood for up to approx. 7 hours in sealed containers at 30 - 35°C showed little or no change in purity and pol content up to 2 hours, after which the values of both parameters fell; on the other hand, refractometric Brix fell at a slight rate from the moment the juice was collected and was associated with the loss of volatile gases, primarily CO_2, as the products of microbial activity. An initial increase in refractometric Brix in some samples was attributed possibly to evaporation of water despite precautions to minimize it, while another possibility was solubilization of fibre and soil components in the juice; over a number of years, fibre solubilization has been held responsible for a higher RDS weight after extraction than before it.

Dextran content of juice and clarifier mud streams

K. M. Onna and R. R. Tamaye. *Ann. Rpt. Hawaiian Sugar Planters' Assoc.*, 1986, 49 - 50.

The apparent dextran % refractometer solids content in juice and clarification mud as measured by the haze method at a number of factories was found at an average of 0.33% to be higher in the mud than in mixed juice (0.12%), while a content of 0.17% in filtrate indicated removal in filter cake; the dextran in filtrate could have contributed to the content in the mud or in clarified juice (0.16%). It was concluded that the differences in dextran content between the various streams could have been caused by changes in the molecular structure of the dextran or by microbial activity in the mud-handling streams, particularly where the temperature was low. A single test showed that the dextran content in syrup was similar to that in clarified juice, so that results obtained for clear juice should be applicable to syrup. It is suggested that the dextran content in mixed juice alone may not be a reliable measure of the degree of deterioration of harvested cane; assuming removal of 90 - 95% of the soil in cane by washing as well as juice exposed to wash water, there is need to establish how cane cleaning changes the dextran content in prepared cane.

Pan control study

G. E. Sloane. *Ann. Rpt. Hawaiian Sugar Planters' Assoc.*, 1986, 51 - 52.

Two control algorithms were developed and tested. One was a two-loop feedback system in which mother liquor Brix was controlled by regulating the opening of the steam valve supplying the calandria while massecuite consistency was controlled by regulating the opening of the pan feed valve; the solids control loop was a cascade loop with the output of the primary (solids) controller providing the set point to the secondary controller regulating the evaporation rate. The second algorithm was a combination feedback/feed forward loop used to estimate the evaporation rate needed to maintain the mother liquor Brix at a desired level for known values of feed rate, massecuite level, crystal content and the programmed rate of increase in mother liquor Brix; the mother liquor Brix control loop was used to adjust the calculated evaporation rate to correct for any bias in the feed forward loop and the corrected value used as the set point for the evaporation rate loop. The two-loop control system worked reasonably well when the characteristics of the footing at the start of a strike were at or near their required values, but difficulties sometimes arose in bringing the footing parameters back to normal levels within a given time when they were abnormal at the start of boiling. Much of the problem was caused by the dead time in the responses of the variables to changes or disturbances in the set points; the necessary use of low controller gain settings (to avoid instability) results in a slow recovery after a disturbance. The combined feedback/feed forward system did not perform as well as the former system, mainly because of its inability to provide a sufficiently accurate estimate of massecuite crystal content.

Pol losses during storage of sugar

T. Moritsugu, B. J. Somera and S. Goya. *Ann. Rpt. Hawaiian Sugar Planters' Assoc.*, 1986, 53 - 54.

Subsamples of three composites of raw sugar having a dry pol of 99.18, 99.31 and 99.29 were stored under three different conditions: (1) in sealed plastic bags left in the HSPA control room at 20°C and 65% relative humidity, (2) in open pans under identical conditions and (3) in open pans in a refinery warehouse. Within the expected variations in pol and moisture determinations, there were no apparent changes in dry pol nor were there any apparent increases in crystal colour for up to 20½ weeks, whereas whole raw colour increased substantially, particularly in the case of sugar stored under conditions (3). A study of sugar samples collected from four locations found to provide sugar of low pol and high moisture showed that pol fell considerably and moisture increased during storage, while high glucose and fructose levels suggested that sucrose inversion was largely responsible for the drop in pol, at least during the initial period. The action of micro-organisms in the molasses film on sugar crystals also affects deterioration, and investigations showed that the yeast and bacterial populations in one of the stored samples were too high to allow them to be counted, while fresh unstored sugar contained <40 yeast colonies and <20 bacterial colonies per g.

Prospects for energy saving by the use of optimum thicknesses of insulations

C. Vázquez B., G. González M. and R. Manduí G. *CubaAzúcar*, 1985, (April-June), 3 - 8 *(Spanish)*.

A nomogram has been developed which can be used to determine the appropriate thickness of insulating material for horizontal or vertical tubing of given diameter and holding steam of a given temperature where the permissible percentage of heat loss has been specified.

Growth of sucrose crystals as a function of purity

M. Wong, S. Ameneiro and G. Rivera. *CubaAzúcar*, 1985, (April-June), 13 - 16 *(Spanish)*.

Sucrose crystals were suspended in solutions of 60, 70, 80, 90 and 100 purity and held at different values of supersaturation (1.02 to 1.10) each at temperatures of 50°, 60° and 70°C. At intervals the crystals were washed with saturated solution, dried and weighed; the surface area was calculated from Kukharenko's formula ($S^3= 69.96\ P^2$), where S is the area in cm^2 and P is the weight in grams. Calculated values of the crystal growth rate in $mg/m^2/min$ are tabulated and illustrated in graph form. They show that the growth rate is greatly affected by the impurities, the greatest effect being by reduction of purity from 100 to 90; subsequent reductions of purity have less effect. For each purity there are no substantial qualitative changes in the dependence of growth on the temperature and supersaturation.

Information system for the management of operations in the area of crystallization

J. E. Cueto R. and R. Consuegra R. *CubaAzúcar*, 1985, (April-June), 17 - 20 *(Spanish)*.

The development of computers has permitted their use for calculation of flows, volumes and times in the boiling process, and a system for such a means of control is proposed.

Method for organizing the operations and stabilizing steam consumption at the pan station considering the milling rate and the thermoenergetic balance

L. Toledo C., H. P. de Alejo C., G. Riera G., E. Domínguez A. and J. R. Taboada. *CubaAzúcar*, 1985, (April-June), 38 - 42 *(Spanish)*.

Theoretical work developed in Cuba, plus experimental data obtained in a Cuban sugar factory, has been used to develop a model which integrates in a systematic way the convenient effects of the methods proposed by the authors quoted, and also includes a mechanism for the stabilization of steam consumption at the pan station. The mechanism is based on the application of an extension of the critical path technique known as "allocation of limited resources". The model allows consideration of the operations programming at the pan station and their interaction with the milling rate - heat balance interrelationship.

Evaluation of the influence of different surfactants on wetting, surface tension, adherence and extension coefficient of sugar cane juices

M. Derivet Z., R. González Q. and M. E. Castellanos G. *CubaAzúcar*, 1985, (April-June), 43 - 47 *(Spanish)*.

The effects of different concentrations of three Cuban surfactants (Espumols E, H and K) on the title characteristics have been measured and are illustrated by tables and graphs. Espumol E produced the greatest reduction of the surface tension and contact angle. The extension coefficient becomes less negative in diluted solutions than in concentrated solutions with increase in surfactant dose.

Usina Açucareira Ester S.A.

Anon. *STAB*, 1986, **4**, (3), 4 - 6, 8 - 10 *(Portuguese)*.

A detailed account is given of agricultural conditions affecting the supply of cane to the title sugar factory/distillery as well as details of the equipment of the plant and processing to sugar and alcohol. Production data for the 1985/86 season are tabulated.

Mill imbibition: influence on extraction, evaporation and the heat balance

G. M. M. Costa. *STAB*, 1986, **4**, (3), 40 - 42, 44, 46 *(Portuguese)*.

Imbibition is generally considered essential to obtain high extraction but the means of applying water or dilute juice and the various types of simple and multiple systems are numerous and the latter selected on a basis of the equipment available rather than a conviction of the efficiency of the system. Calculations are presented of a tandem of four mills with and without triple simple imbibition and the improvement in extraction demonstrated. Bagasse moisture is higher and, for heat economy, it should be dried, perhaps using boiler flue gases. Other factors in achieving good results from imbibition are good cane preparation and use of a suitable means of application, and milling conditions to achieve high primary juice separation. The use of imbibition increases the load on the evaporator but with multiple-effect systems this is not usually a problem, particularly if compound imbibition is employed; calculations are made on a similar basis to the earlier ones with no and simple imbibition showing the reduction in evaporator load. The problem of mill feeding with wetter bagasse is mentioned and means of overcoming this – pressure feeders, etc. – are briefly discussed.

How to preserve the efficiency of polyacrylamide solutions from corrosion during their preparation and storage in sugar factories

E. Angulo A., R. Monduí G. and V. Chopik. *CubaAzúcar*, 1985, (July-Sept.), 3 - 7 *(Spanish).*

The efficiency of polyacrylamides used in clarification of juices, syrup and refinery liquor is diminished when their solutions become contaminated with metallic ions produced by corrosion of the containers. Use of resin and epoxy paint coatings seem to be the best way of preventing corrosion, while aluminium and copper can also be used for making containers. Neither mild steel nor $X_{18}H_{10}T$ stainless steel are suitable.

Study on the application of thermochemical treatment of diffused chroming on iron castings subjected to corrosion and erosion in the sugar industry

Y. S. Novygrod and T. G. Porva. *CubaAzúcar*, 1985, (July-Sept.), 12 - 16 *(Spanish).*

Diffused chromium coatings were given to grey cast iron by treatment with an active mixture of 50% X65 ferro-chrome, 48% aluminium oxide and 2% ammonium chloride. This produced a hard and corrosion-resistant layer of chromium carbides on the surface of the castings which improved their suitability for use as pump impellers and covers, etc., while being more economical than bronze castings.

Influence of the temperature on the saturation coefficient of final molasses

R. Morera C. and M. de la Caridad G. *CubaAzúcar*, 1985, (July-Sept.), 16 - 21 *(Spanish).*

To examine the limit of molasses exhaustion requires exact knowledge of the sucrose solubility in that molasses. The saturation coefficient is supposed to be independent of temperature but a study using temperatures at 5° intervals between 40 and 60°C showed that the coefficient varied significantly with a minimum at 50°C. This should therefore be the mean temperature during the crystallization of final massecuites, especially in crystallizers with short residence times.

Effective method of treatment of water for steam generators. Quality and quantity required

A. Rosales and I. Díaz. *CubaAzúcar*, 1985, (July-Sept.), 22 - 29 *(Spanish).*

The impurities present in raw water and the methods of removing them are discussed and five of these reviewed in turn: filtration through a bed which removes insolubles, clarification with lime, aluminium sulphate, etc., softening by ion exchange of Na^+ for Ca^{++} and Mg^{++}, complete demineralization by various ion exchange sequences, and evaporation. Degasification is also used to remove e.g. CO_2 after ion exchange treatment. The extent of impurities removal by the various methods is illustrated in a graph and the suitability of methods for various kinds of pressure operation tabulated; only complete demineralization is suitable where the boilers operate at high pressure (5000 - 10,000 kPa). The amounts of water available as condensate from various evaporator/pan schemes are calculated for the same throughput of cane (and thus water demand).

Industrial application of chemical treatment to retard deterioration

E. L. Ramos, J. A. Urrutia and R. Linares. *CubaAzúcar*, 1985, (July-Sept.), 36 - 41 *(Spanish).*

Deterioration of raw sugar in bulk storage can sometimes require its re-processing, with consequent losses up to 15%, and an account is given of the application of two methods designed to increase stability of the sugar. The first is to neutralize the syrup with NaOH at the temperature at which it is received, measuring the pH of the samples at hourly intervals. The second involves treatment with NaOH to an unspecified alkaline pH at a constant unspecified temperature and for a suitable time whereby the amino-acids present are removed. Sugars were obtained from the neutralized, alkali-treated and standard syrups and their pH and colour observed at the start and after 3 and 6 months storage. These showed that the alkaline treatment was more effective in avoiding deterioration.

Determination of the residence time and flow patterns in the "RETO" cassette crystallizer

S. R. Mesa and J. Griffith. *CubaAzúcar*, 1985, (July-Sept.), 46 - 50 *(Spanish).*

A new crystallizer design, described by Díaz *et al.*[1] was examined using radio-active tracer and stimulus-response techniques, in order to study the flow patterns and residence time distribution. The times were found to be greater than calculated from the overall dimensions and flow rate, and indicated large dead zones. The authors propose design changes to improve the crystallizer's efficiency.

Direct contact heater for C-massecuite; its advantages

P. Morales, A. Valdés and Z. Hernández. *CubaAzúcar*, 1985, (Oct.-Dec.), 3 - 6 *(Spanish).*

The heater described is a jacket fitted around the massecuite feed pipe of a continuous centrifugal; it is provided with an upper steam inlet and a perforated ring forming its base. Steam entering the jacket heats the massecuite passing down the feed pipe by transmission and then, passing through the ring, heats it directly. Condensate also mixes with the massecuite so that it further reduces the viscosity which is already reduced by the temperature rise. Trials carried out on massecuite of different temperatures and with varying steam flow are reported and the results

1 *Poster presentation at the 18th Congr. ISSCT*, 1983.

tabulated and discussed in respect of heated massecuite Brix and purity, viscosity and the final molasses Brix and purity. It was possible to raise the massecuite temperature by 10°C and reduce viscosity by one-third; this permitted a rise in centrifugal capacity without any increase in molasses purity.

Açucareira Zillo Lorenzetti S.A. – Usina São José ZL

Anon. *STAB*, 1986, **4**, (5), 4 - 6, 8 - 11 *(Portuguese)*.

The title sugar factory/distillery in São Paulo state was established in 1946 and has produced alcohol since 1952/53. Details are provided of its management; of the cane area supplying the factory (45,798 ha, of which 19,112 ha is company-owned); on varieties, agronomic practices, harvesting, loading and transport of cane; of the milling plant and other equipment and processes employed for the production of sugar and alcohol. Data from the 1985/86 season are provided.

Balance of materials by permanent and variable components from clarified juice to refined sugar

O. Janigova and D. Fernández A. *CubaAzúcar*, 1985, (Oct.-Dec.), 21 - 26 *(Spanish)*.

The variation of colour in sugar factory and refinery products has been studied by relating colorant content to soluble inorganic matter in the material; this involves a small error in assumption of a constant amount of the latter but is useful to detect the stages in the process where colour is removed. The method enables monitoring of the movement of colorants and detection of technological errors which may cause high or unnecessary expenditure on decolorizing agents.

Industrial application of chemical treatment retarding the deterioration of raw sugar. Part II

E. L. Ramos S., J. A. Urrutia F. and R. A. Linares C. *CubaAzúcar*, 1985, (Oct.-Dec.), 31 - 36 *(Spanish)*.

The method described earlier[2], whereby syrup is treated with alkali to improve its stability, was applied on an industrial scale. As in the laboratory, there was a considerable reduction in the amino-acid content of the raw sugar producted, but no other effect on the quality indices of the sugar or molasses produced. The low cost of installation and application of the method provides a technique available to producers if their sugar is to be stored for a long time.

Clarification of various kinds of technical sugar solutions by ultrafiltration through a self-rejecting membrane

M. Komoto *et al. Proc. Research Soc. Japan Sugar Refineries' Technol.*, 1987, **35**, 1 - 11 *(Japanese)*.

Ultrafiltration, in which colloids and high-molecular substances built up on a thin layer of gel to form a dynamic (self-rejecting) membrane on a porous ceramic tube, was applied to clarification of cane juices and molasses as well as to run-off and affination syrup from a refinery. In the case of diluted final molasses, flux decreased and rejection of colouring matter increased with time, although flux could be increased by raising the temperature and reached a maximum at 30°Bx; performance of the membrane became almost constant, i.e. its formation was almost complete, after approx. 2 hr. Thickness of the membrane was a function of the circulation rate and its permeability was governed by the operating pressure. At 90% rejection and operating pressures of 0.5 and 2 MPa, the cut-off molecular weights of the membranes formed were 40,000 and 20,000, respectively. The flux was approx. 4 times greater than through an asymmetric organic membrane but the rejection of colouring matter was lower; at 50°Bx, 60°C, 0.2 MPa and a circulation rate of 4 m/sec, the flux was 20 litres/m^2/hr and colour rejection was 10%. Turbidity and suspended matter that could not be satisfactorily removed by Celite filtration or centrifuging were completely removed by the ultrafiltration process, and the taste of the permeate was much better than that of the original molasses. The flux of raw and limed juice was nearly 5 times that through an asymmetric organic membrane; purity of raw juice rose from 83.2 to 84.7 at a flux of 75 litres (9.4 kg solids)/m^2/hr, 60°C, 0.2 MPa and 3.3 m/sec, while that of limed juice rose from 86.7 to 87.9 at 114 litres (12.6 kg solids)/m^2/hr under the same conditions. The flux of rotary vacuum filtrate was 91 litres (6.7 kg solids)/m^2/hr under these conditions, raising the purity from 77.6 to 79.5 and yielding a completely clear permeate, while middle juice and syrup solids flux was maximum at about 12 kg/m^2/hr and 20 - 30°Bx, giving a 1.8 unit rise in purity. The fluxes of 2nd run-off from a factory and of refinery affination syrup were slightly lower at 5.3 - 8.4 and 3.1 kg/m^2/hr, respectively, and purity rose from 54.8 to 56.5 and from 70.1 to 71.2, respectively. Ultrafiltration gave 100% turbidity removal and a colour removal in the range 16 - 44%.

Some ideas on the design of batch and continuous pans

E. E. A. Rouillard. *Proc. 61st Ann. Congr. S. African Sugar Tech. Assoc.*, 1987, 76 - 82.

A method is described for calculating the circulation and evaporation rates in vacuum pans and thus optimize their design. It is shown that for batch pans there is an optimum combination of tube length, tube diameter and circulation ratio that will provide maximum circulation for a given graining volume and heating surface:volume ratio; this combination does not correspond to the shortest tube nor to the lowest circulation ratio. It is advisable to use tubes having a diameter greater than 0.098m. In the case of continuous pans, there is also an optimum tube length governed by the heating surface:volume ratio, and again the use of tubes having a diameter greater than that generally used is advocated for improved evaporation and circulation.

2 Ramos *et al.*: *I.S.J.*, 1988, **90**, 51A.

Beet sugar manufacture

New approach to the sugar factory

D. Herve. *Ind. Alim. Agric.*, 1987, **104**, 665 - 670 *(French)*.

It is suggested that, in the face of keen competition from manufacturers of alternative sweeteners, the sugar industry should consider replacing the conventional sugar factory with two plants, one operating during the campaign and producing only thick juice and the other processing the thick juice in the post-campaign period to provide a variety of products, including liquid and invert sugar (possibly also enriched with fructose), refined sugar, organic hydrolysates, fertilizers and dextrose for use as raw material in the manufacture of chemicals. The use of ion exchange for treatment of the thick juice would provide a high-purity sugar solution some of which could be crystallized in a single strike coupled with the use of vacuum crystallizers, while the remainder of the syrup plus run-off could be further processed to liquid sugar and invert syrup. The non-sugars separated by ion exchange could be further treated to provide fertilizers from the various salts, leaving proteinaceous material for use in animal fodder. The characteristics of the products, consumption of reagents and energy per tonne of sugar and of non-sugar, production schemes and material balances are presented for both beet and cane processing.

Juice purification - practice and development

A. Vígh. *Cukoripar*, 1987, **40**, 95 - 98 *(Hungarian)*.

Developments in carbonatation theory and practice are discussed (with 18 references to the literature), starting with the ideas of Spengler, Dedek & Vasatko and Brieghel-Müller on preliming and continuing with examination of optimum milk-of-lime preparation and feeding conditions. The positive effects on settling of carbonatation mud recycling are discussed and the mechanics of Ca carbonate formation and non-sugars adsorption explained. The contribution made by carbonatation and the positive effect of 2nd carbonatation at elevated temperatures on residual non-sugars separation are considered, and choice of suitable carbonatation vessel is discussed, with particular mention of the tank described by Møller and used at Gørlev in Denmark[1] and the Soviet vessel described by Anikeev *et al.*[2].

Sugar house model calculations applied to examination of the relationship between low-grade raw sugar purity and the (boiling) scheme

L. Megyeri, J. Gerse and K. Hangyál. *Cukoripar*, 1987, **40**, 99 - 105 *(Hungarian)*.

A computer program obtained by mathematical modelling of the boiling process was applied to a simple 3-boiling scheme. Results showed how compensation could easily be made for changes in thick juice purity by adjusting the ratio of 1st wash syrup to 3rd sugar in the remelt liquor. Examination of the effect of low-grade sugar purity on the amount of massecuite in circulation demonstrated how a 2 unit change in purity had the same effect as if thick juice purity had changed by 1 unit. In the centrifugal, a mother liquor viscosity lower than 6.5 Pa/sec was needed if the raw sugar purity was to be >90; approximate linearity was found between the two parameters ($r = -0.89$). Laboratory studies indicated that the raw sugar purity fell with smaller crystal size, while a hyperbolic curve defined the relationship between viscosity and grain size.

Thin juice alkalinity and change in alkalinity during evaporation and crystallization

K. Hangyál and V. Gryllus. *Cukoripar*, 1987, **39**, 110 - 113 *(Hungarian)*.

The significance of juice pH and alkalinity in regard to corrosion, sucrose hydrolysis, non-sugars concentration and carbonatation efficiency is discussed and optimum values are given for thin juice, juice in the 1st, 2nd and 3rd effects of an evaporator, thick juice and 1st and 2nd green syrup at which inversion is minimal. The quality of beet in Hungary is sufficiently poor that alkalization is necessary, either by (i) partial anion exchange, (ii) partial removal of ammonia, (iii) supplementary liming, (iv) adding NaOH in the evaporator or (v) adding NaOH or soda ash in carbonatation. Comparison of the effects of the various methods or their combinations showed that supplementary liming + anion exchange + ammonia removal was the best in that it resulted in a reduction in molasses sugar and in sugar production costs, while merely adding soda ash in carbonatation caused greatest increases in both parameters. There was little difference between the effects on the two parameters of supplementary liming combined with ammonia removal or with NaOH treatment in evaporation.

Energy management

P. Wertán. *Cukoripar*, 1987, **39**, 113 - 116 *(Hungarian)*.

The energy consumption in Hungarian sugar factories in 1986/87 is surveyed and comparison made with the values for the previous campaign, showing a slight improvement. Aspects discussed include fuel consumption by the boilers and lime kilns, thick juice Brix and the use of pulp pressing aids.

Application of the vertical design of crystallizers. IV. Experimental operation of a prototype vertical crystallizer

S. Kucera, J. Klepal and M. Zalabák. *Listy Cukr.*, 1987, **103**, 207 - 212 *(Czech)*.

Trials conducted in 1983 and 1984 on a prototype 63 m^3 vertical crystallizer are reported in which the unit failed to increase the volume of massecuite cooled in the existing battery of trough-type crystallizers; the low-grade massecuite initially of 73 - 85°C was cooled at a rate of just over 1°C/hr. Problems with water flow through the perforated horizontal disc sections led to the

1 *I. S. J.*, 1986, **88**, 84 - 88.
2 *ibid.*, 127A.

conclusion that it would be better to abandon the design in favour of a tubular cooling element where larger crystallizers were concerned. Improvements were achieved in massecuite flow which was approx. 6 tonnes/hr, but problems were caused by excessive crystal contents, and exhaustion varied from 1.3% to approx. 10% purity reduction.

The effect of certain parameters on the process of sugar extraction from beet cossettes

I. A. Oleinik, A. V. Sadych, V. V. Mank and V. T. Kober. *Sakhar. Prom.*, 1987, (9), 21 - 23 *(Russian).*

The effects on sugar losses in pulp of the diffusion coefficient D, the coefficient of mass transfer from the cossette surface to extraction liquid β and of temperature change in the cossettes-juice mixture in the initial zone of a diffuser were investigated. Results showed that prescalding the cossettes with saturated steam to 70°C reduced the losses by comparison with heating only the juice, while increase in both D and β also caused a decrease in sugar losses. Values obtained by calculation using non-linear differential equations agreed to within ± 8 - 10% with experimental results.

The design of main liming vessels

L. I. Pankin *et al. Sakhar. Prom.*, 1987, (9), 23 - 25 *(Russian).*

After a brief survey of some Soviet liming tanks, details are given of a patented vessel design which takes the form of a vertical tank with a conical bottom which is divided into sections by horizontal baffles extending from alternating walls but not stretching to the opposite wall, so that juice flows from bottom to top of the vessel following a snaked path. Just above each baffle are paddles, one on each side of a vertical shaft. A mixing chamber located below the lowest baffle contains horizontal discs mounted on the vertical shaft; each disc carries concentric rows of perforations, and inclined blades attached to the edges of the perforations, both below and above the disc surface but with the angle of inclination below the surface the complete opposite of that above the surface. Juice enters the chamber via a downward sloping pipe and is accompanied by lime which enters the pipe via a vertical port just before the wall of the liming vessel. The inclined blades act as miniature pumps, forcing the juice/lime mixture down between the first pair of discs as they rotate, while the mixture is forced up between the 2nd and 3rd discs, and so on, thus causing intensive mixing. The juice finally leaves the chamber and flows up the main vessel, again being subjected to mixing by the paddles as they rotate at low speed (20 rpm).

A counter for registering the quantity of beets being processed

V. V. Smagin. *Sakhar. Prom.*, 1987, (9), 32 - 33 *(Russian).*

A circuit diagram is presented of a scheme for automatically registering the quantity of beet in process; mechanical counters at the beet weigher are interfaced with a transmitter that sends a signal to a display counter at the diffuser operator's desk. The system can also be modified to register the amount of sugar produced; in this case, the signals are received from the weigher in the packeting section.

Restoration of worn steam chests in inclined diffusers

A. E. Mil'man and Yu. R. Semchuk. *Sakhar. Prom.*, 1987, (9), 33 - 34 *(Russian).*

Advice is given on restoring steam jacket sections on a DDS-type diffuser so as to prolong the useful life of the diffuser by 5 - 8 years without considerable expenditure on new components.

Determination of optimum juice heating by groups of sugar factory heaters

K. O. Shtangeev and V. N. Gorokh. *Sakhar. Prom.*, 1987, (9), 34 - 37 *(Russian).*

Heaters are used for juice at different steps in the overall process and are grouped according to the process stage, e.g. for raw or prelimed juice, 1st carbonatation juice before 2nd carbonatation, thin juice and thick juice. It is shown that there is an optimum level of heating for individual groups of heaters which is defined by the difference in temperature (preferably about 10°C) between the reheat steam and the juice leaving the heater and is a function of the type and flow rate of the juice, the system of heating, cost factors, etc.; while increase in the difference between the two temperatures reduces the heat exchange surface requirement, decrease in the difference (except for the last group of heaters, in which the juice temperature is determined by the technology used) creates conditions for increasing heat economy. The steam consumed in the evaporator is considered as the sum of vapour bled to the pan station, that bled for heating of process materials and the remainder going to the condensers, and the effect of redistribution between the three components is examined; two extremes can arise: where thick juice Brix is unchanged but the amount of vapour passing to the condenser alters and vice versa. The effect of optimum redistribution between juice heaters (that at which the costs of heating for all units in any one group are lowest) is also examined. Calculations are made to show the levels and patterns of the optimum juice differences for various forms of juice, and the desirability of reducing both heating surface area and the difference in juice-vapour temperature is emphasized.

Utilization of the heat from continuous boiler blowdown at Uvarovo sugar factory

I. F. Popov. *Sakhar. Prom.*, 1987, (9), 37 - 39 *(Russian).*

Details are given of the economics of a scheme to heat untreated boiler feed using blowdown water, subsequent disposal of which is also described.

Sugar refining

Modern procedures in the automation of raw sugar refineries

G. R. Alonso G., S. M. Orue V. and V. G. Menéndez Z. *CubaAzúcar*, 1985, (Oct.-Dec.), 7 - 13 *(Spanish)*.

Sugar refineries in different parts of the world have adopted completely centralized control systems such as direct digital control systems or supervisory control with man-machine interaction. From analysis of work with this system of control and management it is concluded that a refinery treating 800 tonnes of raw material per day can operate with 7 men per shift, with consequent gain by diminution of wages and also with an improvement in the final product. An application of these superior means of automation for the control and management of a modern sugar refinery is described.

Installation for obtaining amorphous sugar from pure refinery liquor

C. Fabre, A. Aguilar and M. Delgado. *CubaAzúcar*, 1985, (Oct.-Dec.), 48 - 52 *(Spanish)*.

"Amorphous" sugar is a fine-grain product obtained from whole refinery liquor[1] by a method which gives no molasses. An account is given of trial production at two installations in Cuba, the first being a pilot plant.

Optimization of energy costs in refined sugar drying

A. S. Ginzburg, A. F. Zaborsin, A. R. Kazimirov and A. N. Kashurin. *Izv. Vuzov, Pishch. Tekh.*, 1987, (2), 100 - 103 *(Russian)*.

While infra-red radiation has a number of advantages over convective drying of refined sugar, it consumes more energy; it has therefore been suggested to use it as a 2nd stage when the rate of drying is very much reduced, conventional convective drying being used for the 1st stage. From calculations for an initial sugar moisture content of 1.05% and a final one of 0.2%, it was found that optimum for energy expenditure was a 1st stage drying period of 10.5 minutes (to a sugar moisture content of 0.8 - 0.5%), a 2nd stage period of 5.5 min, a drying air temperature of 80°C and an infra-red radiation of 5 W/m^2.

Fine tuning a refinery: "sucrose"

D. E. Webster. *Sugar J.*, 1987, **50**, (1), 10 - 14.

See *I.S.J.*, 1987, **89**, 40A.

Fluid bed drying of bone char

J. P. Merle, L. H. Bates and L. A. Zemanek. *Sugar J.*, 1987, **50**, (1), 15 - 20.

See *I.S.J.*, 1987, **89**, 87 - 92.

Continuous moistening of sugar before the cube press

S. Lavroff. *Ind. Alim. Agric.*, 1987, **104**, 689 - 690 *(French)*.

The performance and output of a cube press depend on the quality of the sugar/water or sugar/syrup mixture and on the blending accuracy, and an automatic continuous sugar moistening system is described which operates at 95°C and mixes sugar with 1 - 1.8% hot water at an hourly rate of 300 - 1400 kg depending on the cube press demands. Requirements of the moistening unit included a sugar dosing accuracy of ± 0.5% over the entire range of feed rate, a water feed accuracy of ± 1% and a blending accuracy of ± 1.5% with wetting of all the crystals. Full details and diagrams are given of the equipment and its components.

An improved unit for carbonatation of cane raw sugar remelt

M. T. Ibragimov, A. R. Sapronov, A. A. Slavyanskii, S. T. Cherikov and U. Sh. Shabdanbekov. *Sakhar. Prom.*, 1987, (9), 30 - 31 *(Russian)*.

A two-stage carbonatation system is described which comprises two identical vertical tanks with conical bottom and a top section that is wider than the main section. Remelt liquor is fed into an annular gutter at the base of the top section and overflows into the central space to splash down onto a central plate from which it spills over into the juice space; CO_2 enters the vessel at the bottom of the juice space. Above the bubbler are three grids composed of naturally vibrating thin metal sheets of uniform height and sloping at an angle of 70° from the horizontal, but with the slope in one direction in the top and bottom grids and in the opposite direction in the intermediate grid, so that the juice and gas streams follow a zigzag path. Oversaturated juice from the 2nd stage is recycled to the 1st stage. Operation of the modified system over 225 days demonstrated its advantages over the original scheme (without 2nd stage juice recycling, the annular feed gutter and the grids) in the form of reduced overall lime consumption (although reducing matter degradation in liming was almost complete and the 2nd stage liquor contained only 0.01% reducing matter), greater CO_2 utilization, and lower remelt colour content and better filtration and settling properties; white sugar colour was also lower as a result.

A line for mechanized packaging of sugar packets

V. I. Prishchepa, N. G. Rudenko and V. T. Kireichenkov. *Sakhar. Prom.*, 1987, (10), 37 - 39 *(Russian)*.

Details are given of an automatic line for placing 0.5-kg or 1-kg packets of sugar in 20-kg packages at the rate of 2 packages per minute; the line is located at Odessa refinery after a Chambon packeting line of 2100 packets/hr capacity and is manned by 3 operators. The annual monetary saving from the line, three of which were installed in 1986, is mentioned. A packeting line is also under test for granulated sugar at the refinery; the 1-kg packets are filled at a maximum rate of 81 per min, and are then packaged in corrugated paper parcels, 16 packets per parcel.

1 Leão: *Proc. 15th Congr. ISSCT*, 1974, 1246 - 1254.

Laboratory studies

Saturation temperature relationships

G. E. Sloane and B. J. Somera. *Ann. Rpt. Hawaiian Sugar Planters' Assoc.*, 1986, 50 - 51.

Saturation temperatures of three final molasses samples having significantly different reducing matter:ash ratios were determined at various refractometric dry solids (RDS) levels at their original purities and at a number of higher purities obtained by adding known quantities of sucrose. From the results were calculated the solubility coefficients (SC) which were then correlated with the reducing matter:ash ratio, RDS, soluble solids by drying, Clerget sucrose and pol purities on the basis of both RDS and soluble solids by drying. The SC values on the basis of RDS were quite different, the sample having the lowest reducing matter:ash ratio having the highest SC at low purities (as was expected) while the sample with an intermediate ratio had the lowest SC (which was unexpected). This may have been due to the fact that the sample having the highest ratio contained about 50% more chloride ions than the sample of intermediate ratio – chloride ions in the presence of potassium ions adversely affect sucrose crystallization properties. Temperature had a significant effect at the lower purities but only a small effect at the higher purities. Calculation of SC on the basis of Clerget sucrose and soluble solids by drying gave results that were significantly different from those calculated on a RDS basis, all the samples having SC values <1; the effect of temperature was negligible at both high and low purities. Corresponding relationships were found when pol was substituted for Clerget sucrose for the two methods of solids determination, whereas only fair correlations were obtained on a RDS basis; when soluble solids by drying was used as basis, temperature had little or no effect and the correlations were very good. The ratio between soluble solids by drying and RDS as a function of purity was essentially linear, with no significant difference between the curves for the samples having high and low reducing matter:ash ratios; a significant but small difference was found between these and the curve for the sample having an intermediate ratio. The possibility of using an equation to convert RDS to soluble solids by drying is suggested where the reducing matter:ash ratio does not differ greatly for different boiling materials.

HPLC analysis of sugar and factory products

R. R. Tamaye. *Ann. Rpt. Hawaiian Sugar Planters' Assoc.*, 1986, 52.

Analysis of clear juice and syrup samples that had been stored frozen for six years showed that the samples kept remarkably well and demonstrated good agreement between HPLC and classical methods of sucrose measurement, despite an earlier fear that pol and sucrose values as determined by conventional methods would be too high.

Enzymatic treatment of juice and molasses to improve recovery of sugar

N. Nomura and H. W. Hilton. *Ann. Rpt. Hawaiian Sugar Planters' Assoc.*, 1986, 70 - 71.

Enzymes for treatment of cane juice and molasses were evaluated by measurement of viscosity and by HPLC in which the dextran of higher molecular weight is eluted first and produces a relatively sharp peak at retention times of about 5 minutes; as the M.W. falls, the polysaccharide peaks broaden. Oligosaccharides, simple sugars and salts are eluted together at about 11 - 12 minutes' retention. With dextranase treatment, the breakdown of the dextrans of higher M.W. is demonstrated by the shifting of the peaks towards a "catch-all" peak at >11 minutes. Treatment of low-grade molasses with ß-glucanase caused a reduction in viscosity. Pairing of enzymes to enhance activity failed to give encouraging results; several enzymes were combined with very low levels of dextranase and tested on polysaccharides isolated from sour cane juice. It was shown to be better to increase the amount of dextranase and omit the other enzyme.

Influence of polysaccharides on the elongation of the sugar crystal

J. Hormaza and E. L. Ramos. *ATAC*, 1985, **44**, (3), 31 - 34 *(Spanish)*.

Tests were made at a sugar factory which produced raw sugar having a high proportion of elongated crystals, whereby the dextran content of the syrup was eliminated by the use of enzyme. The elongation persisted, however, indicating that dextran is not the cause and that other polysaccharides induce the elongation.

Sampling of disintegrated cane and its influence on the system of cane payment on the sucrose content

C. G. de Oliveira and R. M. Prata. *STAB*, 1985, **4**, (1), 38, 40 - 41 *(Portuguese)*.

For cane payment purposes a 500-gram sample of disintegrated cane is analysed, and this requires sub-sampling a larger amount. A compartmented tray is illustrated which is used for sub-sampling by a factor of 20; tests showed no significant differences between the material in all 20 compartments, the greatest variation – 5.38% – occuring in the fibre % cane figure.

Influence of sugar crystal size on its quality

C. H. Lopes. *Brasil Açuc.*, 1985, **103**, (2/3), 5 - 8 *(Portuguese)*.

Samples of sugar were screened using a set of Tyler sieves and the fractions, of different sizes, analysed for ICUMSA colour and ash content. The results when plotted give parabolic curves with minima at 0.7 mm, so that this is the ideal value; while a CV of 25% would be best, one of 25 - 30% is acceptable, but crystals varying in size by more than 30% from 0.7 mm have too high colour and ash content.

Study of metallic ions in sugar cane. Influence of the soil

R. Hernández, L. García, J. González and E. Rodríguez. *Revista ICIDCA*, 1984, (2/3), 37 - 41 *(Spanish)*.

The ash content and constituents of bagasse influence the characteristics of the paper pulp obtained from it and so a study was made of the relationship between the content of different metals in the bagasse and the soils on which the cane was grown.

Detection of intermediate compounds from the initial stages of the Maillard reaction in model systems of sugars and amino-acids

M. A. Otero and D. Vondrak. *Revista ICIDCA*, 1985, **19**, (1), 1 - 6 *(Spanish)*.

Aqueous 40% sucrose solutions were treated with 0.6% w/v conc. HCl and heated in sealed tubes at 90°C for 0 - 6 hours. They showed an increase in formation of 5-hydroxymethyl furfural with time and also a dark precipitate, presumed to be an aldol condensation product. During the first hour the colour became a pale yellow but the UV absorption increased 2500 times. Maximum inversion occurred in 30 minutes. The solution after 6 hours was treated with equal volume of 2% lysine solution and heated at 70°C, using as control an invert solution (20.94% glucose + 21.72% fructose) similarly treated. Amino-sugars were detected in the solutions, the amount in the inverted sugar solutions being greater than in the control.

Crystal distribution by size, its determination and description in crystallization and fine grain strikes

N. Vara C., J. Lorenzo G. and R. Valdés A. *ATAC*, 1985, **44**, (5), 24 - 28 *(Spanish)*.

The method described for measuring grain size and distribution is to take the sample of massecuite, add it to three times its volume of glycerine and agitate to separate the molasses from the crystals. These are then filtered under vacuum, washed well with sucrose-saturated alcohol, dried in an oven for about 15 minutes at a temperature no higher than 80°C, and subjected to sieve analysis. Using mesh sizes from 30 to 170, corresponding to openings of 0.711 to 0.097 mm, it has been possible to make analyses of massecuites from normal and fine grain sugar with the aid of a special sampler which is described and illustrated; this provides a sample of adequate size but avoids crystal damage, and can be employed irrespective of the vacuum in the pan. Values of the weight fractions of the two types of massecuite are tabulated and shown in graphs which demonstrate the change in size distribution during the strike.

Isolation and determination of oligosaccharides in sugar products

J. V. Hormaza and E. L. Ramos. *CubaAzúcar*, 1985, (July-Sept.), 51 - 55 *(Spanish)*.

The oligosaccharides concerned are those forming in sugar products as a result of microbial action between cutting the cane and milling it. Methods are described for their quantitative determination and preparative separation; the first involves adsorption from a 20°Bx solution on a column of an equal-weight mixture of active carbon and Celite filter-aid. Sucrose is eluted with 3% alcohol, the eluate being monitored by colour formation with anthrone and sulphuric acid. The oligosaccharides are then eluted with 30% alcohol. Recovery of raffinose was 98% from that added to a pure solution of sucrose and 94.4% from a B-molasses. Standard deviation of the measurements was 0.257. For separation of the oligosaccharides the 20°Bx molasses solution was treated with carbon in small portions, separating each before adding the next. The carbon was combined, washed with 3% alcohol to remove sucrose and with 30% alcohol to recover the oligosaccharides which were then separated by fractionation on a column of Sephadex G-10 gel.

Carbohydrates and soluble proteins in sugar cane at different stages of its growth

C. W. Rodríguez, P. Valdés and M. Martínez. *Cienc. Agric.*, 1985, (24), 55 - 61 *(Spanish)*.

Proteins and high M.W. soluble polysaccharides (HSP) are the most abundant constituents of the colloidal fraction of cane juice. Both are considered injurious to the industrial process, especially the HSP. However, it is demonstrated that increasing concentrations after ripening of carbohydrates having a molecular weight of 700 to 10,000 could be a major factor in industrial juice quality.

Determination of nucleic acids in process materials from the manufacture of sugar

M. Quincoses, M. Darias and M. Villalonga. *ATAC*, 1985, **44**, (6), 35 - 37 *(Spanish)*.

Hourly samples of mixed juice, clear juice and syrup were taken during 8-hour daily shifts duiring normal factory processing, while sugar and molasses samples were taken at 10-day intervals. Analyses were made of the nucleic acid contents of the samples and the variations studied. The fall from mixed juice to syrup, i.e. during clarification and evaporation, is possibly through partial adsorption on the calcium phosphate precipitate and/or elimination in the form of scale. The content fell between syrup and sugar, and rose between syrup and molasses resulting from the normal characteristics of the boiling and centrifugalling processes.

Study of the determination of supersaturation by means of the temperature of saturation

M. Wong Q. and S. Ameneiro P. *ATAC*, 1985, **44**, (6), 46 - 49 *(Spanish)*.

The theoretical basis for the use of saturoscopes for determining supersaturation is explained.

By-products

Experimental study of bagasse compaction

O. Suárez and J. Lois. *Revista ICIDCA*, 1984, **19**, (1), 75 - 80 *(Spanish)*.

A study was made of compaction of bagasse to improve its storage and transport. Raw and depithed bagasse, both wet (50% moisture) and pre-dried (13.2% moisture), were compressed in a circular die at pressures of 6.47 to 157.5 kg/cm² and characteristics of the compressed material related to the pressure by regression equations which are expected to be useful in the design of bagasse baling and compaction equipment.

Autohydrolysis of bagasse with a view to preparation of cattle feed. Part I. The process

J. L. Magnani, J. Campanari, A. Vallezzi and C. E. V. Rossel. *Bol. Técn. Copersucar*, 1985, (32/85), 58 - 60 *(Portuguese)*.

The structure of bagasse is described and the improvement of its digestibility in the rumen of cattle by prehydrolysis explained. This can be achieved by treatment with steam under pressure and "explosion" by sudden release of pressure. Tests were made using a pilot autoclave at Usina Barra Grande in which 150 kg batches of bagasse were treated with steam at 17 kg/cm² for 5 minutes (out of a 17-minutes cycle). Conditions must be such as to avoid furfural production and some was detected when the treatment exceeded 5 minutes. Composition of the crude and hydrolysed bagasse is tabulated.

Electroaeration of solutions in submerged fermentation of citric acid

N. A. Glushchenko. *Rpt. Grodn. Sel'sko-Khozyaistv. Inst.*, 1987, 7 pp.; through *Ref. Zhurn. AN SSSR (Khim.)*, 1987, (15), Abs. 15 R448.

Conditions are given under which efficient electroaeration of the culture solution is produced in submerged fermentation of citric acid. It is shown that electroaeration has a considerable effect on the basic process parameters: there is a rise in citric acid productivity and in yield on sugar while molasses consumption falls.

Feeding of pressed pulp silage to dairy cows and fattening bulls

P. Lebzien and K. Rohr. *Zuckerrübe*, 1987, **36**, 252 - 254 *(German)*.

Aspects of beet pulp silage covered include: its composition, nutrient content, energy value and digestibility; rate of consumption (particularly favoured by addition of molasses); its effects on rumination and rumen fermentation; recommended quantities for feeding to dairy cows and fat stock; its performance by comparison with other energy rations; and advice on silage preparation.

Production of liquid sugar from Egyptian cane molasses using ion exchange resins

A. M. El-Naggar, M. A. El-Maghraby, A. Abou El-Ela and A. M. El-Sherbeny. *Taiwan Sugar*, 1987, **34**, (3), 12 - 15.

Phosphatation proved the best of four pretreatment methods to reduce the CaO, ash and colour contents in diluted Egyptian cane molasses intended for use as raw material in liquid sugar manufacture. pH affected the results, and pH 4.0 - 4.5 was found to be optimum.After phosphatation, centrifugation and filtration, the molasses was passed through four ion exchange columns containing a strongly acidic cation exchanger, a weakly basic anion exchanger, a mixed bed of a weak cation exchange resin plus a strong anion exchanger, and a combination of strongly acidic cation exchanger, weakly basic anion exchanger and a mixed bed of weakly acidic and strongly basic resin. The resultant straw-coloured liquid sugar had a Brix of 70° and a pH of 7.9 and contained 97.3% invert sugar on Brix, 570 colour units % Bx and 0.095% ash on Brix. Since concentration to 70°Bx caused an increase in colour, a small bed of bone char was used for final polish filtration.

Studies on the fermentative production of L-lysine. Optimization of culture conditions for L-lysine fermentation of cane molasses

Y. T. Liu. *Taiwan Sugar*, 1987, **34**, (3), 25 - 36.

See *I.S.J.*, 1988, **90**, 94.

Possibilities of increasing the yield of baker's yeast from molasses

E. Sobczak. *Przem. Ferm. i Owoc.-Warzyw.*, 1986, **30**, (9), 15 - 17; through *Ref. Zhurn. AN SSSR (Khim.)*, 1987, (16), Abs. 16 R332.

It is noted that in Poland between 1.25 and 1.85 kg of molasses is consumed in the production of 1 kg of baker's yeast. Ways of increasing yeast yield are examined, including optimization of the process, use of suitable yeast cultures and improving control methods.

A new approach to feeding concentrate to dairy cows

J. Harland. *British Sugar Beet Rev.*, 1987, **55**, (3), 34, 36 - 37.

Higher milk yields, milk fat contents and milk protein yields were obtained in trials in which a pelleted product containing 70% molassed beet pulp was fed as a supplement to silage at the same crude protein rate as a conventional starch-based compound. The new product, containing fishmeal, soyabean meal, a mineral/vitamin supplement in addition to the molassed pulp, has the advantage over a starch-based feed of allowing silage digestion to continue throughout the day; the fermentation of starch in the rumen causes the pH to fall and thus adversely affects the action of the cellulolytic bacteria, so that silage is not broken down and the animal stops eating because of a full rumen.

A dried molassed sugar beet-based compound feed for dairy cows

G. Fishwick and G. Hemingway. *British Sugar Beet Rev.*, 1987, **55**, (3), 34 - 35.

Trials showed that the molassed pulp feed described in the previous abstract gave a milk yield and quality comparable to those given by a mixture of two commercial products in the case of lactating cows and ewes; the molassed pulp has the advantage of a virtually constant composition from year to year, whereas the formulation of other commercial products varies.

Advice on (beet) pulp ensilage

J. P. Vandergeten. *Betteravier*, 1987, **21**, (222), 15 *(French)*.

Advice is given on beet pulp ensilage, including the optimum dimensions of the clamp and its preparation.

Pulp drying with superheated steam and mechanical vapour recompression

-. Poumillade, -. Ternynck, B. Marcotte, -. Roche and -. Jollion. *Ind. Alim. Agric.*, 1987, **104**, 647 - 655 *(French)*.

In tests with a prototype dryer at Villenoy sugar factory, the use of superheated steam to dry beet pulp followed by compression of the vapour ensured that the only components leaving the dryer were pulp and condensate (thereby reducing atmospheric pollution) while the energy consumption was reduced from a total of 860 kWh (810 in the form of heat) with conventional drum drying to a maximum of 200 kWh; pulp dry solids was increased from approx. 21% to a value (88%) that remained very constant at a maximum variation of only ±1%. Other advantages included a 5.7% gain in pulp through elimination of losses by burning, greater drying uniformity, a much whiter pulp which was easily pelleted, and a condensate of such quality that it could be used for diffusion without any pretreatment, thus providing a supplementary heat recovery of 2%. Possible future developments and improvements are discussed.

Dynamics of nitrite accumulation during alcoholic fermentation of molasses wort

G. Sobkowicz and E. Oziuba. *Zesz. Nauk R Wroclaw. Technol. Zywn.*, 1986, (4), 81 - 87; through *Ref. Zhurn. AN SSSR (Khim.)*, 1987, (17), Abs. 17 R317.

Alcoholic fermentation of molasses wort was investigated with addition of nitrates and nitrate-reducing bacteria. It was found that the nitrite concentration in the medium at the end of fermentation was 0.13 mg NO_2/100 cm^3, but bacterial contamination increased this to 0.22 mg/100 cm^3. Addition of nitrates with and without contamination increased the nitrite concentration to 0.4 and 0.26 mg/100 cm^3, respectively. However, these nitrite concentrations had no essential effect on the process.

Technique for producing oxalic acid from sugar by a pressure method

P. Nowak. *Przem. Chem.*, 1986, **65**, (9), 488 - 490; through *Ref. Zhurn. AN SSSR (Khim.)*, 1987, (17), Abs. 17 R328.

Results are given of studies on production of oxalic acid from sugar by a pressure method. The positive effect of pressure on yield was demonstrated. A scheme is presented and the operation of the unit described for production of the acid by a modified sugar oxidation method under pressure. Two variants of the process technology are given.

Acetylated pectic polysaccharides of sugar beet

I. C. M. Dea and J. K. Madden. *Food Hydrocolloids*, 1986, **1**, (1), 71 - 88; through *Ref. Zhurn. AN SSSR (Khim.)*, 1987, (17), Abs. 17 R416.

A method has been developed for isolating acetylated pectic polysaccharides (APP) from sugar beet or beet pulp. The sugars composition was determined in polysaccharide samples after hydrolysis as well as the acetate, methoxy group and protein contents. APP was analysed using nuclear magnetic resonance and chromatography on a column of DEAE-cellulose. Prepared samples of APP contained (%) uronates 44 - 60, neutral sugars 12 - 22, methoxides 4 - 6 and acetyl groups 2 - 9. Rhamnose, arabinose and galactose were found in all polysaccharide samples, covalently bonded with an acid polymer. A large part of the APP was made up of surfactants having foam-forming properties and effective in the formation of emulsions. Using APP, stable emulsions were obtained in oil-water systems containing 10% oil and particles measuring approx. 1 μm. No interrelationship was found between the acetylated pectin fractions as regards their ability to form foams and emulsions and their structures.

Products and by-products from bagasse in GEPLACEA member countries

A. C. Sturion. *STAB*, 1986, **4**, (3), 47 - 51 *(Portuguese)*.

Some of the members of GEPLACEA are world leaders in the utilization of bagasse and the article reviews products and levels of production, etc., in respect of cellulose and paper, fibre and particle board, furfural, animal fodder, carbon and other uses of bagasse.

Synthesis of methyl cellulose from bagasse dissolving pulp

B. García, S. Askienasi and V. García. *ATAC*, 1985, **44**, (6), 20 - 22 *(Spanish)*.

Bagasse dissolving pulp, containing 92% α-cellulose, was converted to alkali cellulose by stirring with NaOH, after which it was added to methyl sulphate at different temperatures and for different times. The product was milled to a white powder and compared with imported methyl cellulose. The results obtained indicate the possibility of producing methyl cellulose from bagasse and thereby saving import costs.

Beet pulp drying in superheated steam under pressure

A. S. Jensen, J. Borreskov, D. K. Dinesen and R. F. Madsen. *Zuckerind.*, 1987, **112**, 886 - 891.

See *I.S.J.*, 1987, **89**, 105.

halved adsorbent usage. This figure is now running at a very satisfactory 5.5% on melt. As a result, only one kiln is required for regeneration of adsorbent. Decolorization from raw liquor to fine liquor averages 75%.

Bone char at Millaquin has traditionally carried a high carbon level which at times has reached 18%. Decarbonizing is practised on a continuing basis to maintain the carbon on char at around 10%.

During the first year of operation with Canesorb, minimal make-up of bone char and Canesorb was allowed while decarbonizing conditions were being established. Make-up rates have since been increased and it is expected that decolorization levels will rise above 75%.

Continuous affination

For many years Millaquin Refinery operated with shut-down periods for annual leave and maintenance in June and December. Year-round operation was introduced in 1984 to service customer requirements better. It was recognised that additional centrifugal capacity would be necessary to allow down-time for maintenance on the existing batch centrifugals.

With knowledge that continuous centrifugals were in use for affination overseas and with our experience of these machines on low-grade duty in the raw sugar factory, the application of a continuous unit for affination in the refinery was investigated. The decision was then taken to install a single continuous centrifugal which would be equipped to process affination and boil-out massecuites sequentially.

Components of various centrifugals were held as spares on site. From these components, we elected to build a "hybrid" centrifugal and decided to mount a 1-metre diameter 34° continuous basket on a batch centrifugal spindle[1].

A mounting boss was designed to attach the basket to the spindle with a flywheel to provide additional mass to the basket. A new monitor case with integral molasses chamber was fabricated to suit the configuration of the continuous basket. The variable speed facility of the Ward-Leonard drive of the batch centrifugal was retained and this feature has proven useful in the handling of a wide range of massecuites.

The "hybrid" machine has run with few problems. Initially, difficulties were experienced in maintaining a crystal layer on the screens. Application of wash water tended to provide lubrication which accelerated the sugar toward the top of the basket. This problem was overcome by the introduction of steam through a perforated pipe onto the basket. A crystal layer was then maintained and wash water requirements were very much lower. Capacity of the centrifugal is given in Table II. The data were obtained at a massecuite level of one metre above the iris feed valve. With increased head, it is possible that capacity could be higher, particularly on affination magma.

Table II. Throughput for various valve openings

Feed valve opening, %	Massecuite throughput, tonnes per hour
20	6.8
40	11.0
60	13.6
80	17.2
100	20.0

Energy consumption for the hybrid centrifugal was very low. For an average massecuite throughput of 12 tonnes per hour, the power requirement is of the order of 7.5 kW. Low crystal breakage is a feature of operation of the hybrid machine. Typical results are shown in Figure 2. While crystal breakage is not of concern for affination and low-grade duty where sugars are melted, handling of white massecuite would require the use of devices to minimize breakage. Most affination work is done at 900 r.p.m. and at this speed breakage is around 8%.

Performance of the hybrid centrifugal has been excellent in terms of mechanical stability, power consumption and affination syrup production. The unit was installed at a very reasonable capital cost.

Refining process (1987)

The refinery process presently practised at Millaquin is shown in Figure 3. Affination, polish filtration and Canesorb have been major innovations which have provided significant economic benefit.

Impurity levels in the main course have dropped since the change to affination. This is reflected in the lower ash levels in white granulated sugar.

Sweetland filters handle some 300 -

1 Smith & Howard: *Proc. Australian Soc. Sugar Cane Tech.*, 1985, 167 - 171.

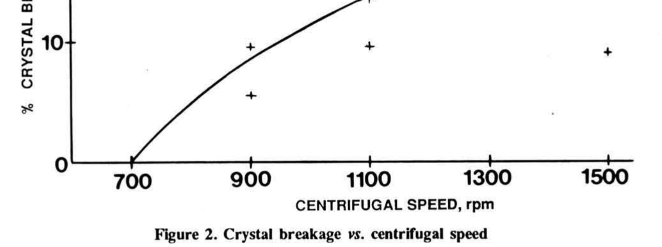


Figure 2. Crystal breakage *vs.* centrifugal speed

400 tonnes of liquor per cycle before cleaning is necessary. Weekly operation for a 600-tonne melt requires only one 30-tonne Canesorb adsorber to be brought on line and one sweetened-off each week.

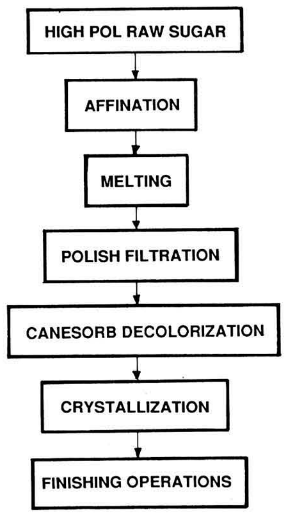


Figure 3. Refining process, 1987

Sweetwater production has been reduced substantially owing to the long cycles associated with polish filtration and the change to Canesorb. The water balance of the factory has improved as a result. Melting of affined and remelt sugars is carried out using hot water when sweetwater is not available.

Adsorbent usage has declined dramatically from 18% char in 1978 to 5.5% using the Canesorb blend. Overall, the work of the char house has been very much reduced.

Plant requirement and manning

Numerous plant changes have occurred in the period covered by this paper. New labour- and cost-efficient plant has been installed for reasonable capital expenditure, some equipment has been upgraded and a great deal of redundant plant has been removed. This has been achieved with the predominant requirement that the process should match the criteria of the input sugar and maximum benefits should be gained therefrom.

Labour cost has been an important consideration throughout the period of change. The number of operators per shift has been reduced from 14 to 4 during this time. The labour reduction has been achieved without redundancy by placement of operators in other sections of the complex as jobs became available.

Summary

The small Millaquin Refinery has faced particular challenges in remaining a viable operation. Its current melt is of the same order as that processed prior to the 1914/18 World War. This paper describes the particular process and operational changes introduced during recent years to reduce operating costs. The emphasis has been on obtaining maximum benefit from the natural advantage of an adjacent raw sugar factory and the use of conventional technology in innovative ways to overall cost benefit.

Liquor decolorization with granular carbon

Experience at two Australian refineries

By P. J. Field
(CSR Limited, Refined Sugars Group, Sydney, Australia)

Introduction

One of the on-going issues for CSR refineries during the 1960's was how best to replace or upgrade the char plant in each refinery. With perhaps one exception, the char houses were characterized by outdated plant, often in poor mechanical condition, with high maintenance costs, unacceptable working conditions, and actual or approaching shortages of melt capacity. Straight replacement of old char plant with new looked too costly. Attempts to resolve this issue led to several responses:


P. J. Field

an increased research effort into the nature of sugar colorants with the ultimate aim of improving the decolorization process,

a search for more cost-effective adsorbents, and

upgrading of affination and recovery stations to allow a shift in refinery colour and impurity loadings from char treatment to recovery crystallization.

The CSR research effort into the nature of colorants has been reported previously[1-3]. That work is still being continued today, but with less urgency. An early outcome of the research effort

Paper presented to the 46th Meeting, Sugar Industry Technologists, 1987.

1 Kennedy & Smith: *Paper presented to 35th Meeting, Sugar Ind. Tech.*, 1970.
2 Paton & Smith: *Proc. Tech. Session Sugar Processing Research*, 1982, 1 - 23.
3 Bardwell *et al.*: *Proc. 44th Meeting Sugar Ind. Tech.*, 1985, 32 - 52.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and the search for more cost-effective adsorbents was the installation of an anion exchange resin decolorizing unit at the CSR Glanville refinery in Adelaide. This plant replaced an old char house as the single decolorizing stage used at that refinery. Experience with that particular plant, and development work associated with brine recycling, have been reported previously at SIT and SPRI meetings[4-6].

Early results from resin decolorization encouraged CSR refinery people to plan for replacement of other char houses with resin units. However, this planning work and the continuing experience in operations at the Adelaide refinery brought out two major difficulties which made the replacement of the char plants with resin decolorization less attractive. These difficulties lay in performance and brine disposal.

Decolorization performance

Despite close attention to operating conditions, long-term decolorization performance stabilized at 75 - 80% for filtered carbonatated liquor to fine liquor, and fine liquor colours were typically in the range of 200 - 250 colour units (measured at pH 9). As a result, the Glanville process did not consistently produce average refined sugar which met the CSR standard (20 ICUMSA colour) when used in conjunction with a three refined massecuite boiling system. When crystal from the second and third boilings was mixed with crystal from the first boiling, there were periods when the combined crystal product exceeded 20 colour.

While this could have been overcome by remelting crystal from the third boiling, or adding an additional decolorizing step, the objective of replacing char with a single, cost-effective decolorization step would not have been achieved. In this context, it was accepted that any replacement decolorization step should be able to achieve 90% colour removal.

Brine disposal

The second difficulty was the strong indication from environmental authorities that, in the future, dark brine regenerant from resin decolorization would not be acceptable for discharge to river systems, owing to its high chloride, colour and BOD levels, or to sewerage systems on account of its chloride content.

Although a satisfactory process for regenerant treatment and reuse was developed[5,6], such an addition to the resin process increased operating costs and made resin less attractive.

As a result, planning for further resin installations was stopped and the char/carbon alternatives considered again. By this time, the Pyrmont char house replacement was becoming urgent, and advice was sought from a number of other refineries and carbon suppliers to add to the existing CSR information base about relative benefits/problems of char and activated carbon decolorization.

Early in 1979, the decision was made to replace one of the two remaining Pyrmont char houses with granular activated carbon in a fixed-bed system during 1981, and to follow this installation with replacement of the second char house in 1983. The project studies clearly showed that granular carbon was likely to achieve the high decolorization level sought, be the most cost effective decolorizer, and meet all environmental requirements.

Pyrmont carbon plant

Construction of the stage I carbon plant began in 1980, with installation of major items complete by mid-1981. Design capacity for Stage I was 4200 tonnes melt per 120-hour week, and the plant was to operate in parallel with the No. 1 char house, while the No. 2 char house was decommissioned. The plant was constructed to meet a CSR process specification by Jord Engineers Australia, with technical advice from Fletcher & Stewart Ltd. CSR retained Calgon Carbon Corporation as technical advisers to assist with process design and commissioning. Provision was made in the design to allow expansion in Stage II to about 8000 tonnes melt capacity. The total cost for Stage I was $Aus 2.4 million.

Plant description

The stage I plant comprised:

five stainless steel adsorber columns of 2.75 metres diameter and working volume 71 cubic metres of carbon. Design flow rate through the columns was 22 kl/hour, with pressure drops of about 200 kPa over each column, and a maximum column working pressure of 680 kPa,

a six-hearth kiln of 2.6 metres internal diameter and fired with natural gas on the third, fourth, fifth and sixth hearths. An after-burner was fitted to remove any off-odours from the kiln. Design capacity was 0.5 tonnes dry carbon per hour and all kiln controls were based on a microprocessor control system,

tanks to hold spent and reactivated carbon, and carbon transport water,

piping for liquor and water flows and carbon transport,

structural steelwork needed to make the plant a "stand-alone" unit, outside the refinery process buildings, and

a separate 1.35-metres diameter adsorber column for golden syrup decolorization.

A simplified flow sheet of the plant is shown in Figure 1. The existing refinery process – affination, carbonatation and filtration – was retained ahead of the decolorization stage.

Plant operation

In operation, four columns ran liquor while the fifth column was on stand-by – sweetening-on or -off, filling or emptying carbon. The liquor columns were set up as two pairs directly coupled and running in series. Filtered carbonatated raw liquor (brown liquor) was pumped over the first column ("lead" column) of a pair to produce intermediate liquor. The intermediate liquor then flowed over the second column ("trail" column) to produce fine liquor for pan feed. After filling with carbon, a column would

4 Cunneen & Hawkins: *Paper presented to 31st Meeting, Sugar Ind. Tech.*, 1972.
5 Delaney *et al.*: *Paper presented to 37th Meeting, Sugar Ind. Tech.*, 1978.
6 Delaney: *Proc. Tech. Session Sugar Processing Research*, 1980, 207 - 211.

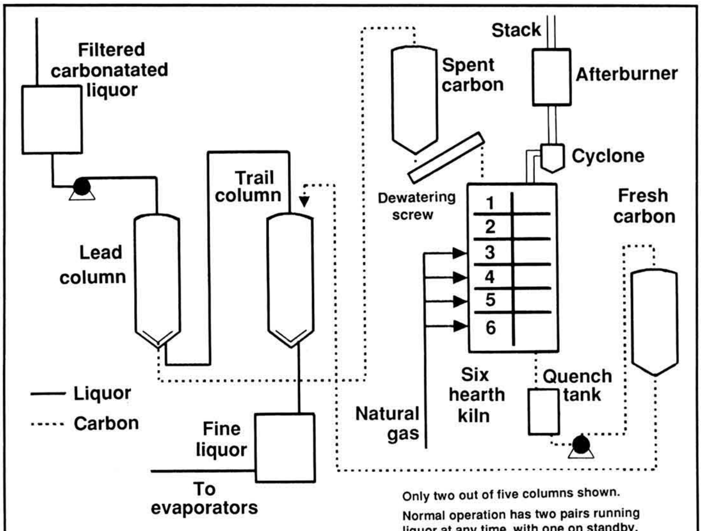

Figure 1. Pyrmont refinery simplified carbon plant flowsheet (Stage 1)

spend about 200 hours in the trail position, and then be changed to the lead position for a further 200 hours. At the end of this time the column would be sweetened-off with about 2 - 2.5 bed volumes of water, and the carbon transferred to the spent carbon tank ready for reactivation.

Target overall decolorization was 90% from raw (brown) liquor to fine liquor, with about 80% decolorization in lead columns. For typical average feed liquor colours, this decolorization performance would produce a fine liquor for pan feed at less than 150 colour (measured at pH 9). The kiln design throughput of 0.5 tonnes dry carbon/hour was intended to ensure 0.8 tonnes of carbon burnt per 100 tonnes liquor solids treated. The adsorbent used was Calgon "Cane-Cal".

Plant performance - Stage I

While plant performance during commissioning was generally satisfactory, production operation were soon affected by a number of serious difficulties which reduced throughput and overall decolorization. An estimate of overall plant performance can be made from the process data set out in Table I.

The operating difficulties which led to these results included:

(a) *kiln frequently out of use*, caused by hearth refractory failures, burner control system faults, and carbon feed blockages. On one occasion, serious overheating of the kiln occurred owing to a damper control failure and an inadequate response by operators and supervisors. This last incident led to a 10-week period out of action in order to rebuild the kiln. We believe that this incident was the main cause of some subsequent hearth failures and a recent failure of the kiln centre shaft.

(b) *Reduced liquor flow rates* (and throughput), caused by high column pressure drops.

As a result of these operating problems, the planned Stage II expansion was postponed, and a program of investigation and remedial work initiated. The investigation work showed that:

the hearth failures were caused by too-rapid heating during kiln start-ups, and large temperature differentials over hearths. The critical nature of the rate of temperature rise during start-up had been recognised at the design stage, but the form of control installed proved to be inadequate for an operating plant,

there were frequent burner flame failures, which appeared to be related to an "on-off" operating pattern, and pilot flame failures,

the weekend stand-by system, when the kiln was kept at a supposedly safe (for the refractory) temperature during shutdowns, was made ineffective by the flame failures, which led to the kiln cooling down below the safe temperature over weekends,

unsatisfactory kiln operation caused increased levels of carbon fines which settled as a layer close to the tops of

Table I. Decolorization performance of the Stage I plant

	Design basis	Year ended March 1983	Year ended March 1984
Decolorization, %			
Lead columns	80	71	69
Trail columns	50	43	45
Overall	90	84	83
Fine liquor colour*	<150	176	206
Fine liquor pH	>7.0	7.6	7.5
Liquor flow rates, kl/hour	22	18.2	18.8
Carbon used on liquor solids, %	0.8	0.33	0.42
Kiln throughput, tonnes/hour	0.50	0.23	0.26
Carbon apparent density	0.50	0.50	0.53
Molasses number	240 minimum	270	249
Iodine number	1000	832	Discontinued
Particle size, mm	1.0	0.97	1.00
Abrasion number	75 minimum	83	79
Actual kiln hours as % available	-	68	72
Number of hearth failures	0	3	2
Carbon loss, % of carbon reactivated	<5	9.8	4.1

* Liquor colour measured at pH 9

adsorber columns. The carbon fines then acted as a filter for the relatively low levels of carbonate particles carried over in filtered liquor. The combined carbon/carbonate fines caused high column pressure drops and reduced liquor flow rates. Hard lumps formed from the fines also caused blockages at the outlet of the spent carbon tank. High column pressure drops were also associated with air pockets in the carbon beds. This was traced to the liquor piping configuration which allowed the trail columns to partially empty out at low flow rates, with the vacuum relief valves on the columns then opening and admitting air. When flow was re-established, air remained in the carbon bed and was not vented despite an automatic air bleed system installed at the top of each column.

With kiln performance clearly the critical element, remedial work to overcome these operating difficulties included:

general upgrading of operating fault alarms, including over-temperature alarms for each hearth, and

changes to the burner control system including installation of individual gas pressure regulators for each burner pilot flame, and new combustion air valves which increased the possible burner turn-down ratio from 2.3:1 to 6:1. The new valves allowed a linear air:gas ratio through the full fire range.

a feedback loop for gas control was also added (gas follows combustion air) to compensate for changes in furnace pressure:

burner control logic was changed to positively limit heating rates to about 35° - 45°C/hr,

the kiln was maintained at operating temperature at weekends, rather than at lower stand-by temperatures (about 600°C). Maintaining weekend temperatures incurred a small gas penalty equivalent to 18 MJ/tonne melt but removed the weekly temperature cycling at start-up and shut-down. It also provided a significant capacity increase for the kiln through the increased operating time available with no warm-up time,

a change was made from bricked hearths to hearths cast *in situ,* and

in addition, the columns were strengthened mechanically with external stiffening rings to stand full vacuum. This allowed removal of the vacuum relief valves. Liquor piping was repositioned to ensure the columns ran full at all times (except for drain-out sequences).

All of these changes added a further $Aus130,000 to the original plant cost. However their effect was to make a substantial improvement in the kiln and column performance and reliability. For example, between January 1982 and March 1985, fourteen hearths had failed, whereas between March 1985 and June 1986, no hearths failed, although two older hearths had to be rebuilt as a result of previous damage.

Consistent kiln operation reduced the carbon fines content so that carbonate mud was no longer trapped in the columns, and column flow rates and pressure drops approached design standards. Decolorization started to improve. As a result, design work on Stage II was recommenced. The design refinements were also incorporated in the New Farm refinery carbon plant, for which planning and installation were already underway.

Pyrmont Stage II plant

Approval to proceed with Stage II plant was obtained in March 1985, and the plant was commissioned in June 1986.

The original concept for Stage II was for duplication of the columns and kiln within the existing Stage I steel framework. Besides the high cost for duplication, such a plant would have given a total melt capacity of 8400 tonnes/week, higher than necessary to meet market forecasts.

As a result, a lower-cost design concept was developed. This concept involved columns operating at higher flow rates than the Stage I plant, and increasing the kiln capacity. A simplified flowsheet of the Stage II plant is shown as Figure 2.

Design concept and plant description

The Stage II plant comprised:

three new adsorber columns, similar to the existing columns, giving a total of eight columns for liquor decolorization. Columns operated as 3 lead and 3 trail columns, with two on stand-by,

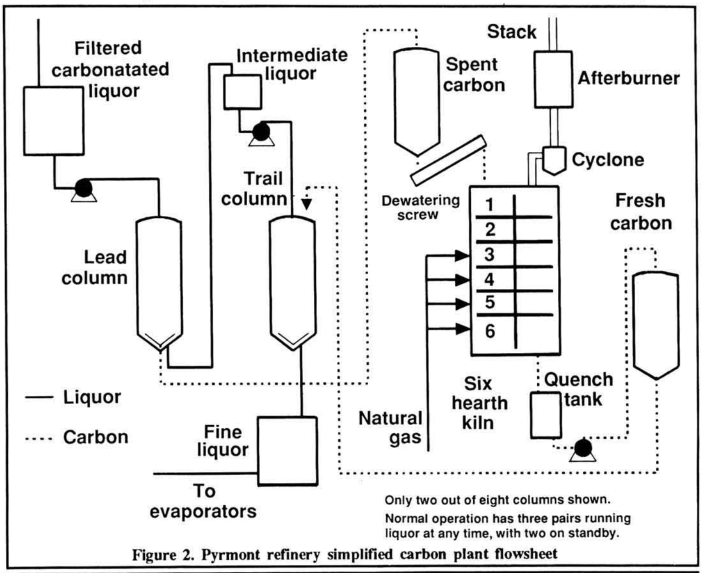

Figure 2. Pyrmont refinery simplified carbon plant flowsheet

up-rating the kiln from 0.5 tonnes carbon/hour to 0.7 tonnes/hour by installing larger burners in Nos. 4 and 5 hearths,

installation of an intermediate liquor tank between lead and trail columns so as to remove column pressure drop limitations, and allow increased liquor flow-rates (22 kl/hour up to 26 kl/hour average) at higher density,

control system changes to simplify operation and better match the full plant to the supervisory control system then being installed, and

some old cast iron liquor tankage replaced with stainless steel.

The extended plant replaced the remaining No. 1 Char house and was to provide benefits including:

four operators would become surplus,

all of the old char plant could be demolished, opening the way to improved factory housekeeping and further redevelopment of the site, and

an increase in decolorization capacity up to about 7000 tonnes melt capacity per week, substantial energy savings, and improved decolorization compared with the old char plant.

The plant was completed essentially on time and within the budget cost of $Aus1.8 million. Commissioning proceeded smoothly. While relatively low in cost, the design concept presented some risks:

kiln breakdowns at previous levels would limit refinery operations,

up-grading of the kiln to 0.7 tonnes/hr by increasing heat input might mean that gas velocities in the kiln would be too high, leading to fluidization of carbon particles in the hearth drop-holes, and carbon emission from the stack, and

increasing liquor flow rates from 22 to 26 kl/hour, and a reduction in the design contact time from 6 hours to about 5 hours, might impair decolorization performance.

All of these aspects were studied, and there was agreement that the risks were acceptable in the light of either operating experience or trial results. In particular, kiln performance and availability had improved to the point where reliance on a single kiln was acceptable, with the proviso that an extra column (the eighth column) was installed to allow storage of one column full of fresh carbon ahead of the filling schedule. This column would provide cover for the most likely kiln breakdown – failure and replacement of a hearth.

Performance

Recent plant performance is summarized in Table II. While performance of the Stage II plant has been influenced by addition of new carbon into the new columns, it seems clear that the design standards will continue to be achieved when the long-term operating conditions are reached. Since the plant was commissioned, the planned manning reductions have been achieved, and the old char plant demolished. With the carbon plant occupying an area about one-sixth of the area previously occupied by the char plant a large area of land is now available for further development of the refinery.

Table II. Decolorization performance of the Stage II Plant

	Design basis	Typical data year ended March 1987
Decolorization %		
Lead columns	80	81
Trail columns	50	57
Overall	90	92
Fine liquor colour*	<150	98
Fine liquor pH	>7.0	7.8
Liquor flow rate, kl/hour	26 kl/hr maximum	21.6**
Carbon used, % on liquor solids	0.8	0.7
Carbon apparent density	0.50	0.51
Molasses number	240 minimum	275
Particle size, mm	1.0	0.98
Abrasion number	75 minimum	73
Carbon loss, % of carbon reactivated	<5.0	2.4
Gas usage, MJ/tonne carbon used	-	10,050
Afterburner	-	Not required

* Measured at pH 9
** Average rate required to meet melt – actual rates achieved in excess of design

A further benefit from the carbon decolorization process has been the opportunity to market a lower-cost liquid sugar produced by evaporation of fine liquor.

New Farm carbon plant

Construction of the New Farm carbon plant at Brisbane commenced in January 1984 and commissioning began in May 1985. The design concept was similar to the Pyrmont Stage I design, but with detail changes intended to avoid the operating problems experienced in the Pyrmont plant. Installed cost for the plant was $Aus 3.2 million.

Plant description

The New Farm plant comprised:

five stainless steel adsorber columns of 2.5 metres diameter, each holding about 28 tonnes of carbon and with a maximum working pressure of 680 kPa,

one golden syrup column,

a 6-hearth kiln, gas-fired, with a design capacity of 0.5 tonnes/hour,

spent and fresh carbon tanks,

associated tankage and piping for water, liquor flows and carbon transport,

steelwork to support the plant as a stand-alone unit, outside the main process buildings, and

a Bailey Network 90 process control system for liquor column and kiln control.

The adsorbent selected for the Brisbane plant was Mitsubishi Chemicals "Cane-Diahope".

Plant operation and performance

The operating system is similar to that used for the Pyrmont Stage I plant, with four columns on liquor at any time.

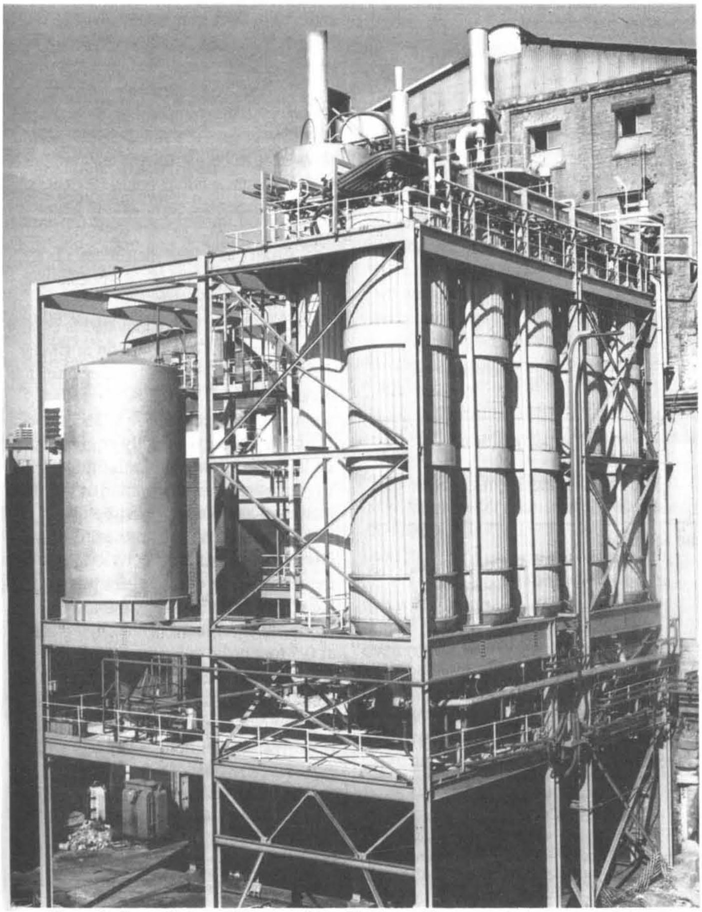

Figure 3. General view of Pyrmont Stage II Carbon Plant, showing adsorber columns and fresh carbon tank

The columns are set up as two pairs of columns in series. With a design capacity of 0.5 tonnes of carbon/hour, the kiln has substantial overcapacity for maintaining a carbon usage of 0.8% on liquor solids at New Farm's current melt rate of about 450 tonnes/day. As with Pyrmont, the granular carbon decolorization step follows conventional affination, carbonatation, and pressure filtration of carbonatated liquor. However, unlike Pyrmont, New Farm incorporates a "polish" liquor filtration of carbonatated raw (brown) liquor immediately prior to the carbon columns. Recent performance is summarized below in Table III.

In summary the New Farm plant is performing consistently, and close to the design specification. The expected benefits in operating cost savings have also been achieved.

Kiln study

One consequence of the Pyrmont operating difficulties in 1982/1985 was that planned investigational work had to be abandoned. However, with a successful commissioning period completed at New Farm, it was appropriate to begin looking at some of the kiln operating parameters with the aim of improving reactivation and reducing energy costs. It was also thought important to establish a reference base for operating parameters on which future development work at both Pyrmont and New Farm might be based, and used for subsequent performance comparisons. The study was essentially seen as empirical and with direct measurements wherever possible.

Particular areas of interest were:

carbon residence time in the kiln,

alternative operating regimes such as operation with maximum carbon feed rate and kiln shut-down periods *versus* low rate and continuous operation,

whether weekend "stand-by" at operating temperatures similar to Pyrmount practice was appropriate, and

the effect of steam injection on the bottom hearth.

In all of these studies, changes in carbon apparent bulk density, molasses number and iodine number were used to assess the quality of reactivation. Two feed rates were used: 0.5 tonnes dry spent carbon/hour (the design maximum) and about 0.3 tonnes/hour, the feed rate roughly corresponding to continuous kiln operation during normal refinery operations, with weekends on stand-by.

Prior to commencing this work, it was necessary to develop safe, representative sampling procedures for spent and reactivated carbon, together with methods for accurate kiln feed rate measurements, and adjustment of burner air/gas ratios.

While the work at New Farm is not complete, some preliminary data is presented for interest.

Kiln residence time

The kiln residence time was determined at a rabble arm rotation speed which gave what appeared to be the best carbon profile on each hearth. This occurred when the hearth was completely covered in a full "saw-tooth" pattern (see Figure 4 below).

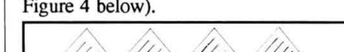

Figure 4

Table III. New Farm decolorization performance

	Design	Actual year ended March 1987
Decolorization, %		
Lead columns	80	81
Trail columns	50	40
Overall	90	90
Fine liquor colour*	<150	134
Fine liquor pH	>7	7.3
Carbon used, % on liquor solids	0.8	0.7
Carbon apparent density	0.50	0.52
Kiln throughput, tonnes carbon/hour	0.50 maximum	0.48
Molasses number	240 minimum	253
Iodine number	1000	820
Particle size, mm	1.0	NA
Abrasion number	75 minimum	90
Carbon loss, % of carbon reactivated	<5	3.5
Gas usage, MJ/tonne carbon reactivated including after-burner	-	19,680

* Measured at pH 9

Residence time was estimated directly using carbon tagged with lithium chloride. Results at two feed rates are shown in Table IV.

Table IV. Residence time estimates

Feed rate, tonnes dry spent carbon/hour	Rabble arm speed, rpm	Residence time, minutes
0.31	1.22	78
0.59	1.33	62

From Figure 5, which shows more detailed results, it is clear that carbon flow rate through the multiple hearth kiln is far from ideal, with substantial by-passing and back-mixing.

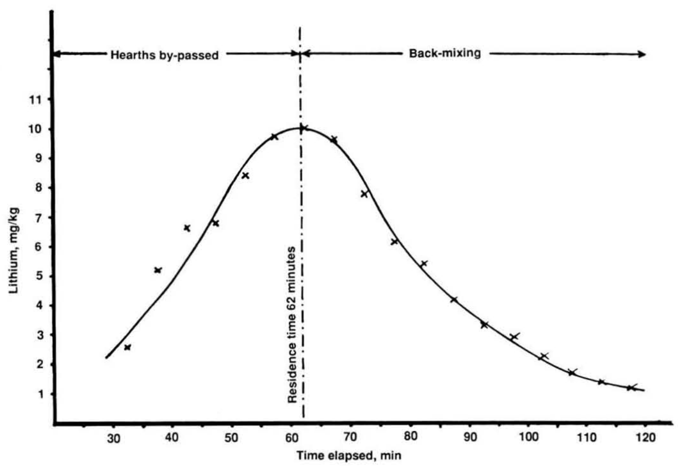

Figure 5. G.A.C. kiln residence time at 1.33 rpm rabble arm speed and 0.50 tph feed rate

Alternative operating regimes

The initial operating regime for the New Farm kiln was with 0.5 tonnes/hour dry spent carbon feed rate, the after burner at high fire - gas valve 100% open, approximately 650°C at optimum air/gas ratio, and the kiln at full stand-by (operating temperature) between reactivation runs. The temperature profile was: Hearth 3 - 700°C, 4 - 800°C, 5 - 930°C, and 6 - 950°C.

In deciding alternative operating regimes, an after-burner temperature of 300°C was selected since observation showed it to be the lowest temperature at which no stack odour could be detected. Other regimes are summarized below in Table V.

From gas usage measurements for these operating regimes, it appears that, for the New Farm kiln, gas consumption per tonne of carbon reactivated is comparable to Pyrmont consumption at similar feed rates and without the after-burner at about 10500 MJ/tonne. A controlled warm up from cold requires about 33000 MJ of gas energy, roughly equivalent to 24 hours gas consumption at full stand-by, and afterburner operation at 300°C increased gas consumption by about 25% over the "no afterburner" usage. Given the relationship between the requirement for carbon reactivation and New Farm kiln capacity, the lowest energy cost regime is given by operation at 0.5 tonnes/hr, with the kiln shut down between each regeneration run which lasts about 60 hours. New Farm has now adopted this least-cost regime. However, as mentioned earlier, Pyrmont experience indicated that such an operating regime contributed to hearth failures owing to the effects of the heating/cooling cycle on the kiln refractory. It is believed that the closely-controlled heating and cooling rates achieved at New Farm will minimize hearth failures.

Effect of steam injection

The effect of steam injection to the bottom hearth was tested for bottom hearth temperatures of 900°C, 925°C, and 950°C, and feed rates of 0.3 and 0.5 tonnes of carbon per hour. While the relationship between residence time at the maximum hearth temperature, actual bottom hearth temperature, and water concentration in hearth gasses is not clear, it appears that operation at high hearth temperatures is more effective than increasing steam injection rate. Further work is needed to understand

Table V. Comparative kiln operating costs – Test regimes

Case	1	2	3	4	5	6	7
Feed rate, tonnes/hr	0.50	0.50	0.50	0.50	0.50	0.30	0.30
After-burner	High fire up to 650°C	300°C	300°C	Off	Off	300°C	300°C
Full stand-by temperatures	Yes	Yes	No	Yes	No	Yes	No

better the gasification reaction on the bottom hearth.

In addition, limited testing of the relationship between apparent bulk density and iodine and molasses numbers indicated no apparent relationship between iodine number and apparent bulk density changes, while there were indications of a possible relation between molasses number and apparent bulk density changes.

Conclusion and general comments

Although the operating problems associated with the Stage I Pyrmont plant initially cast doubts on the decision to use granular carbon as the sole decolorizer, performance with Pyrmont Stage II and New Farm has shown that the design target of 90% decolorization at 0.8% carbon usage can be achieved.

While it is still early days, present experience suggests that use of an intermediate liquor tank, with mixing of lead column product, does not appear to have a significant effect on decolorization. Certainly, the objectives of reducing operating costs, upgrading plant and working conditions have all been achieved. Environmental requirements, in terms of kiln stack odours or particle emission, have also been met. However, there remain some areas of concern where further investigation is needed. Two such areas are inversion over carbon, with resultant loss of product yield, and the relevance of molasses number in evaluating carbon types and reactivation performance.

Under normal operations, reducing sugar levels in fine liquor are typically 0.02% above those of filtered liquor at Pyrmont, and even more than this at New Farm. After weekend shutdowns, reducing sugar levels of over 1% have been measured in the first runnings from the columns. Recirculation of liquor at about half normal flow rate during shutdowns has been found partially to overcome this reducing sugar gain. Besides the inversion issue, fine liquor pH levels at New Farm are typically 0.5 pH units lower than for Pyrmont. It is not clear at this stage whether these effects are related to differences in carbon types or in operating procedures between the two refineries. One significant difference in operating procedures is that polish filtration at New Farm removes additional fine carbonate particles which are not removed at Pyrmont.

The other issue is the relevance of molasses numbers in assessing carbon quality and reactivation. Is there a real performance difference between carbons of say 200 molasses number and 270? We would like to know.

Finally, both Pyrmont and New Farm had char houses in use for close to 100 years. Many sugar technologists spent large amounts of time there finding out a whole range of interesting and useful information. Char work became part of the refinery "culture". With new carbon plants at these refineries, and the strong possibility of other CSR refineries changing to carbon later, it seems there will be a whole generation of new technologists who will never know the joys and disciplines of working in a char house. Whether that is a good or bad thing is a matter of opinion.

Acknowledgement

The author wishes to acknowledge the contributions made to development and operation of the decolorization plants by the technical staff at CSR Pyrmont Refinery, including Messrs. H. P. Benecke and C. D. MacKenzie, and technical staff at CSR New Farm Refinery, including Dr. D. Mackintosh and Mr. B. R. Willis.

British Society of Sugar Cane Technologists

The 1988 Annual General Meeting and Spring Technical Meeting of the BSSCT were held on April 7 at the Royal Commonwealth Society in London. In the absence through ill-health of the President, Dr. R. A. Yates, Vice-President Simon Winn took the chair. The meeting approved the minutes of the 1987 AGM and of the Extraordinary General Meeting of October 1987, as well as the accounts for the year ended December 31, 1987. Matters arising included the Mike Bennett Award; a short statement was issued describing the basis on which the award would be made and that for 1987 was presented to Dr. M. Workman in respect of his paper on cane cultivation improvement in Swaziland, presented at the Autumn 1987 meeting. A visitor from the USA – Dr. Margaret Clarke, of Sugar Processing Research Inc. – was welcomed and it was mentioned that other overseas visitors who were in London at the time of meetings would be welcomed. The date of the Autumn meeting was announced as October 7, 1988.

After lunch, the technical meeting included four presentations, including one on "Shipping bulk sugar – a problem of stability" by James Somner and John Pearson; "Savings in cane transport" by Harold Poole; "Progress in irrigation" by M. Kay (read by Bob Davey of Tate & Lyle) and "Computer simulation of a pan house" by Mike Donovan. It is expected that some if not all of these will become available as written papers and that it will be possible to publish either the texts or summaries in this Journal. The availability of printed texts, preferably at the meetings, will be a factor in the selection of the Mike Bennett Award winner, and it seems likely that this will encourage their preparation and distribution.

Make a date .....

When?	Who?	Where?
May 19, 1988	Research Society of Japan Sugar Refineries Technologists, c/o Japan Sugar Refiners Association, 5 - 7 Sanbancho, Chiyoda-ku, Tokyo 102, Japan.	Kobe Convention Centre
June 6/9, 1988	South African Sugar Technologists Association, c/o S.A.S.A. Experiment Station, Mount Edgecombe, Natal, South Africa 4300.	Marine Parade Holiday Inn, Durban
June 8/9, 1988	Sociedade dos Técnicos Açucareiros e Alcooleiros do Brasil, C.P. 532, Piracicaba, SP, Brazil 13400. (Environment Seminar)	Riberão Preto, SP.
June 15/17, 1988	American Society of Sugar Cane Technologists, Knapp Hall, Louisiana State University, Baton Rouge, LA 70803, U.S.A.	Holiday Inn - Surfside, Clearwater, Florida.
August 1988	Philippines Sugar Technologists, Room 308, Doña Salud Building, San Juan Street, Bacolod City, Philippines.	
August 1988	Sociedade dos Técnicos Açucareiros e Alcooleiros do Brasil, C.P. 532, Piracicaba, SP, Brazil 13400. (Administrative/Information Seminar)	Aguas de São Pedro, SP.
September 1988	Barbados Sugar Technologists' Association, Hythe, Welches, Christ Church, Barbados, W.I.	Barbados
September 1988	Sociedade dos Técnicos Açucareiros e Alcooleiros do Brasil, C.P. 532, Piracicaba, SP, Brazil 13400. (Human Resources Seminar)	Barra Bonita, SP
September 14/16, 1988	Sociedad Colombiana de Técnicos de la Caña de Azúcar, Apartado 4448, Calle 58 Norte No. 3N-15, Cali, Colombia.	Cali.
October 1988	ARTAS Réunion Society of Agricultural & Sugar Technologists, c/o C.E.R.F., B.P. 315, 97490 Ste.-Clotilde, Réunion.	Réunion.
October 1988	Sociedade dos Técnicos Açucareiros e Alcooleiros do Brasil, C.P. 532, Piracicaba, SP, Brazil 13400. (Sugar cane by-products Seminar)	Piracicaba, SP
October 7, 1988	British Society of Sugar Cane Technologists, 55 Liddon Road, Bromley, Kent BR1 2SR, England.	Royal Commonwealth Society, London.
November 1988	Sociedade dos Técnicos Açucareiros e Alcooleiros do Brasil, C.P. 532, Piracicaba, SP, Brazil; 13400. (Maintenance Seminar)	E.S.A. 'Luiz de Queiroz", Piracicaba, SP.
November 3/4, 1988	Jamaican Association of Sugar Technologists, c/o Sugar Industry Research Institute, Kendal Road, Mandeville P.O., Jamaica, W.I.	Jamaica
February 26/March 2, 1989	American Society of Sugar Beet Technologists, 2301 Research Boulevard, Suite 107, Fort Collins, CO 80526, U.S.A.	Hyatt Regency Hotel, New Orleans, LA.
September 23/29, 1989	International Society of Sugar Cane Technologists, c/o STAB, C.P. 532, Piracicaba, SP, Brazil 13400.	São Paulo, Brazil.

Zambia sugar expansion[1]

Zambia's sugar production in the 1988/89 crop year is forecast to rise to 150,000 tonnes, of which between 30,000 and 40,000 tonnes will be for export, according to a spokesman of Zambia Sugar Co. Ltd. This output, nearly all of which would be white sugar, compares with around 130,000 tonnes in the 1987/88 crop year, when exports would be around 20,000 tonnes. Export sales in 1986/87 had reached a record 56,351 tonnes, but this was because Zambia had built up surplus stocks. Better yields and use of production capacity are intended to raise annual production to more than 160,000 tonnes over the next few years. Exports go currently to Zaire, Rwanda, Brundi and Tanzania but the company has also received inquiries from Uganda and has done a small amount of business with Angola and Namibia. Zambia is strongly lobbying the EEC to grant it an annual export quota. Like Papua-New Guinea, Zambia has been proposed for inclusion on the sugar protocol between the Community and the ACP countries[5]; initially a supply quota of zero is proposed but both countries will be able to claim a supply quota as soon as one becomes available within the total quantity of about 1.3 million tonnes, white value.

Denmark sugar exports, 1987[2]

Denmark exported 4,090 tonnes of raw sugar in 1987 and 294,976 tonnes of white sugar, giving a total of 324,716 tonnes, raw value. Principal destinations for the white sugar were Norway (81,384 tonnes) India (41,200 tonnes), the UK (40,496 tonnes) Israel (28,133 tonnes), Tunisia (15,554 tonnes) Iran (14,000 tonnes), Syria (13,200 tonnes), Peru (12,600 tonnes) and Iceland (10,369 tonnes).

1 *Reuter Sugar Newsletter*, January 13, 1988.
2 F. O. Licht, *Int. Sugar Rpt.*, 1988, **120**, S93.

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