Observations on Ion-Exchange Purification of Low-Grade Syrups in Conventional Beet Sugar Refining Methods

E. F. BERRY AND ROBERT S. GADDIE[1]

IN 1947 AN ION exchange pilot plant was installed at the West Jordan, Utah, factory to determine the feasibility of increasing the sugar-end capacity by decreasing the load on the low raw side of the house—this being the holding point at this particular factory. To do this, the new installation was applied to process intermediate green syrup (syrup from high raw centrifugals). It was anticipated that the plant would handle approximately one-fourth of the intermediate green syrup produced. After processing, the effluent syrup from the exchanger was to be returned to the factory thin juice for evaporation to thick liquor.

Equipment

The installation includes a syrup dilution tank at the remelt station, a juice cooler or heat exchanger, a raw syrup storage tank, a sweet water storage tank, a battery of six exchanger tanks arranged to operate in the series, cation-anion, and three tanks for mixing regenerating agents. The pipe and valve sizes are all 2 inches. All tanks and pipe lines on the exchanger, and all acid mixing tanks and pipe lines are rubber lined. Valves are the Hills-McCanna rubber-lined diaphragm type.

One resin recovery trough is installed on the backwash header line leading from the cation tanks to the sewer, and another is on the backwash header from the anion tanks. These are nothing more than riffle boxes where the rate of flow of the backwash water coming from the tanks is decreased, thus enabling the riffles to retain any resins in suspension. Resins are taken from the anion backwash recovery trough periodically and returned to the various anion tanks. There is but little tendency for the cation resins to come over with the backwashing, and as a result it has not been necessary to return any cation resins to the exchanger battery.

The exchanger tanks are each 4 feet in diameter by 12 feet high. The design of the cation and anion tank is identical. The resin beds are 6 feet in depth with a volume of 75 cubic feet each. Resins used are cation C3 and anion A2 produced by Chemical Process Company. Each tank has three small glass-covered rectangular ports to enable the operator to view the inside of the tank to keep the liquid levels adjusted, and to insure that proper juice or regenerant flow is being maintained.

[1]Chemist, West Jordan, Utah, factory and General Chemist, Utah-Idaho Sugar Company, respectively.

Operation

Syrup from the high raw centrifugal machines is diluted to 30 brix at the remelt dilution tank using either sweet water from the exchanger or factory thin juice when there is no exchanger sweet water. The exchanger sweet water furnishes but a small portion of the dilution agent necessary, and the thin juice is used the greater part of the time. The temperature of this diluted juice is then lowered from 60 degrees to 20 degrees Centigrade by pumping through the juice cooler at the head end of the exchanger supply tank.

The juice is then pumped into the exchanger battery where it makes a double pass through two pairs of exchangers operating in series alternating cation, anion before entering the factory thin juice. Thus, four of the tanks in the battery are in use in the juice cycle at one time while the other two are in the regeneration cycle. The juice enters the top of the first cation tank, runs through the bed into the first anion tank and into the sewer until the effluent syrup from the anion tank reaches one brix. This syrup is then turned into the factory thin juice. The finished juice runs until the limiting pH, purity, and color control point is reached. This control point will be discussed later. The only time that the juice makes this single pass is when the exchangers are first started up with all of the tanks completely regenerated and clean. The first pass juice is then turned into the second pair of exchangers and run to the sewer until the effluent from the anion tank reaches one brix. This effluent is then run as finished syrup directly into the factory thin juice. The exchanger syrup increases in brix gradually until it reaches 26 brix while maintaining the requisite purity. The brix of the hourly composite, however, will average 20. When the juice purity lowers until the control point is reached, the juice from the second exchanger pair is turned into the third exchanger pair which is then "sweetened on" in the regular manner and the effluent run into the finished juice line. At this point the resins in the first pair are completely exhausted and the intermediate green syrup is turned off from this pair which is then cut out of the cycle. Water is turned into this pair and the syrup washed from the cation through the anion and into the sweet water tank until the effluent reaches two brix. The beds are individually washed to the sewer, backwashed, drained to bed level, regenerated, washed, and are then ready for the next cycle. At the same time that the incoming syrup is shut off on the number one set, it is turned into the number two cation tank which then becomes the number one in the cycle. This makes the flow of incoming raw syrup continuous and of the finished syrup practically continuous except for the short sweetening-on period.

Differential air pressures are used to force the juice from one tank to another in the operating cycle. The air pressure is carried at 17 pounds per square inch on the first tank, at 15 pounds per square inch on the second, at 10 pounds per square inch on the third tank, and at not more than 5 pounds per square inch on the last tank from which the finished juice is taken.

Control

The installation requires one operator per shift and one helper to make up the regenerating sulfuric acid and sodium hydroxide solutions on two shifts only. The regeneration tanks are of sufficient capacity to enable one shift to operate without refilling the tanks. Larger tanks, of course, would eliminate another of these helpers.

Control of the exchanger battery is maintained by keeping the apparent purity of the hourly composite of finished syrup above 90 percent based on the hourly laboratory analysis. Ninety percent was established as the lower control point on purity as this will give a juice of the same average purity as the white pan at the West Jordan factory. Another control used is that of color comparison. Samples of effluent syrup from the exchanger are diluted to the same brix as the regular factory thin juice and a color comparison is made. When the color of this diluted syrup becomes as dark as the color of the factory thin juice at the same brix, a new set of exchangers is cut into the cycle, and the first pair cut out, washed and regenerated. In addition, it has been found that by regulating the flow of the effluent at 15 gallons per minute and using the differential air pressures previously mentioned, one pair of exchangers will last 2 hours and 45 minutes as the number two pair in the cycle delivering finished syrup and another 2 hours and 45 minutes as the number one pair delivering the lower purity single pass syrup.

Thus a pair of exchangers will last $5\frac{1}{2}$ hours before becoming completely exhausted. When spoiled beets are being cut and dark colored syrups result in all the factory juices and syrups, the life of an exchanger pair drops to 2 hours operating as the number two set or for a total time of only 4 hours. Thus, an excellent measure of control has been obtained by keeping the color of the effluent syrup lighter than the color of the thin juice and by taking one pair of exchangers out of the cycle and regenerating every 2 hours and 45 minutes.

Prior to the beginning of the operation a continuous recording conductivity meter was ordered, but up to the time of the writing of this report it had not been received. Information concerning the use of the recording conductivity meter as a control measure in regulating the length of the cycle is not presently available.

Regeneration

The time required for "sweetening off," backwashing, draining to bed level, regenerating, and washing is approximately 2 hours. This allows sufficient time to keep one pair of clean exchangers on hand well in advance of the time needed for continuous operation. Sulfuric acid is used in regenerating the cation beds and sodium hydroxide in regenerating the anion resins. Approximately 315 pounds of sulfuric acid (specific gravity 1,835) made up to an 8-percent solution are used per regeneration. The last half of the sulfuric acid after passing through the cation bed is returned to the

acid mixing tank where the concentration is increased to 8 percent with the addition of new acid. This is then used for regenerating the next cation bed. Two hundred twenty-five pounds of sodium hydroxide are used for each anion regeneration. This is also made up to an 8-percent solution. None of the caustic solution is recovered. These amounts of regenerants were varied until a point was reached which gave maximum capacity for a minimum regenerant.

In addition to the acid regeneration the cation beds are regenerated once each week with 8-percent sodium hydroxide solution to displace the calcium sulfate in the beds and also to remove the proteins and other organic substances introduced into the beds from the juice. Large amounts of calcium sulfate are also precipitated out of the acid solution in the acid tank where the dilute acid returned from the exchanger is increased to 8-percent concentration. It is necessary to flush out this acid make-up tank once weekly to remove the deposits of insoluble calcium sulfate.

Wash water used in washing out syrup, backwashing, and washing out excess regenerants was taken from the condenser leg line of the low raw pan. The average temperature of this condenser water was about 50 degrees centigrade. As the operation got under way, the supply of wash water from this source was found to be insufficient, and it was necessary to supplement this water with water from the well supply. This reduced the average temerature of the wash water to 30 degrees. The only advantage of the warmer water which we could detect was a slight decrease in the time required to wash out the syrup and caustic solutions.

After regeneration both cation and anion beds are individually washed to the sewer until the pH of the cation reaches 2.0. At this point the water from the cation is used in washing the anion. The small amount of acid introduced into the anion bed by this 2.0 pH water apparently has no effect on the capacity of the anion resin. However, we believe it serves to neutralize some of the small amount of residual sodium hydroxide retained by this bed.

Discussion of Results

At the time this paper was written, the ion exchange plant had been in operation continuously for a period of 75 days, and after the first week there has been no measurable decrease in the capacity of the exchanger battery. Some invert sugar is made by the passage of the syrup through the exchanger. The average increase in invert sugar based on dry substance was 1.12 percent and the average temperature of the incoming raw syrup was 23 degrees. The increase in the amount of invert sugar has presented no difficulty in the regular factory control. As a result of the addition of the exchanger effluent to the evaporator supply syrup there has been no increase in the color of the factory thick juice, and no difficulties have been experienced in filtering or working sugar-end products. Determinations on final molasses have averaged about .5 percent of invert on dry substance. No problems have been experienced in Steffenizing this molasses.

Since the exchanger effluent has been added directly to the regular factory thin juice during the entire campaign, the only method of obtaining performance figures is to make comparisons with previous years. Such comparisons are shown in table 1 where the weighted averages for the previous 5- and 10-year periods are compared with results obtained this year.

Table 1.

	% Sugar in beets	Purity beets	Tons sliced	Low raw mass. % in beets	Increase net extraction	Decrease molasses % beets
10-year average	15.814	84.52	74,474	9.7	4.04	1.10
5-year average	15.742	84.36	90,303	9.9	4.70	1.22
1947	15.050	84.84	91.550	7.7		

From this table it may be noted that while the average purities were approximately the same for the three periods, the net extraction was 4.04 percent greater this year than the 10-year previous period and 4.70 percent better than the previous 5-year period. Similarly, there was a decrease in molasses production of 1.10 percent on beets on the 10-year period and 1.22 percent on beets for the 5-year period. Also, it is interesting to observe the decrease on low raw massecuite percentage on beets from 9.7 and 9.9 to 7.7. This constitutes a decrease of approximately 25 percent in the amount of low raw massecuite to be processed.

To better illustrate the increase in net extraction resulting from the ion exchangers, a comparison between the relative purity of beets sliced and the net extraction for the past 15 years is shown graphically in figure 1.
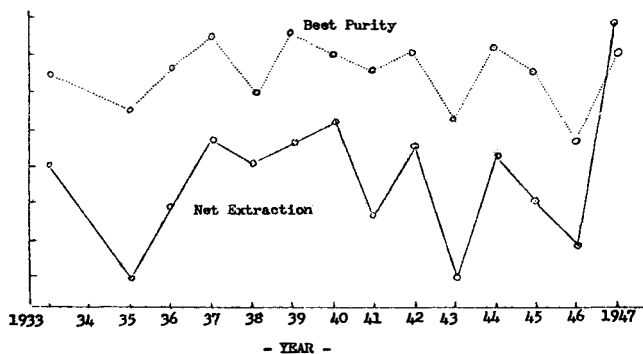

Figure 1.

It may be noted here that the net extraction practically parallels the beet purity curve and that it is approximately the same distance below it until 1947 where the two curves cross and the extraction becomes greater than the average beet purity.

The estimated costs and net returns resulting from the operation of the exchanger pilot plant are presented in table 2. The operating data in table 1 are used in these computations.

The net profit shown in table 2 does not include amortization of the initial cost of the exchangers, maintenance, resin losses, power, or fuel costs. The exchangers require very little maintenance and consequently these costs will be low. A total of only five pumps are required to pump all of the acid, caustic, and juice solutions, and of these only two are continuously in operation. Thus the power costs will be small. Inasmuch as the intermediate syrup is diluted mostly with thin juice, little additional fuel above that regularly required to evaporate the thin juice is needed to raise the exchanger syrup to the seventy brix thick juice level. A small amount of exchanger sweet water is used in diluting the raw syrup to thirty brix and also some dilution occurs in passage of the syrup through the beds. This will introduce a slight error when fuel costs are eliminated from the cost statement.

Table 2.

Operating costs per 100 tons of beets

Three exchanger operators and two helpers	$ 3.90
Regenerants (H_2SO_4 and NaOH)	9.90
Total operating costs	**$13.80**

Net costs and returns per 100 tons beets based on previous ten years

Operating costs	$13.80
Molasses not recovered at $40.00 per ton	44.00
Total costs	**$57.80**
Credit on increased extraction sugar at 6.5 cents per pound	79.04
Net profits per 100 tons beets	**$21.24**

Net costs and returns per 100 tons beets based on previous 5 years

Operating costs	$13.80
Molasses not recovered at $40.00 per ton	48.80
Total costs	**$62.60**
Credit on increased extraction sugar at 6.5 cents per pound	91.96
Net profit per 100 tons beets	**$29.36**

In addition to the advantages enumerated in the above discussion, it has been found that the processing of a portion of the syrup which would normally go to make up the low raw pan has resulted in a much better control of the purities on the sugar end, particularly on the raw side. At times it is inevitable that purities of intermediate syrup will be too high to obtain a minimum loss of sugar in molasses. At such times most of this high-purity syrup can be processed through the exchangers until the regular sugar-end control has brought the intermediate syrup again to the requisite purity, thus saving the extra sugar which would be lost through the processing of one or two high-purity crystallizers. Conversely, during certain periods in the latter part of campaign, it is not uncommon for the purity of the West Jordan beets to drop to less than 80 percent. When this occurs, it

is very difficult to maintain sugar-end purities, especially on the white side, at high enough levels. Under such conditions, the ion exchange plant, still processing intermediate green syrup, may be used to great advantage.

Summary

The ion exchange pilot plant at the West Jordan factory processing intermediate green syrup has worked successfully in supplementing the regular refining methods, and the installation has been limited in its function only by the pilot-plant size. Inasmuch as the West Jordan factory is equipped with a pulp drier and supplies molasses for one of the Steffen's factories in addition, it is not desired to eliminate molasses production entirely. However, with a larger plant, molasses production could be controlled from the demand by varying the quantities of syrup processed through the ion exchange plant.

An increase in capacity of 25 percent on the low raw side of the factory was realized with the installation. A substantial increase in extraction with a subsequent decrease in the amount of molasses produced as compared with averages for the previous 10 and 5 years at this factory was obtained. No detrimental effects in the normal processing of the mixture of this ion exchange effluent and the regular factory juices were observed. Sugar-end control of crystallizer purities was simplified. Higher white pan purities on juices from deteriorated beets were obtained.

High Pressure Evaporation

FRED F. COONS[1]

At THE PRESENT time there are only a few long tube vertical evaporator installations in this country in the sugar industry, one of which is located at Woodland, California. It is hoped, because of their novelty, that there will be interest in this dissertation, which will attempt to describe the various phases of high pressure evaporation in a beet sugar factory.

General Description

Figure 1 shows a general cross section of a long tube vertical evaporator. Steam is introduced at the top of the tube bundle through an annular baffle and flows down and parallel to the outside of the tubes. Condensate and non-condensable gases are removed from the bottom just above the tube sheet. Juice enters the evaporator at the bottom and boils as it flows upward through the tubes. As the liquor evaporates large volumes of vapor are formed which cause vapor and juice to issue from the tubes at high

velocities. One might say that the action in a long tube vertical evaporator closely resembles that in a common household coffee percolator. The mixture of vapor and juice impinges on an umbrella-shaped deflector plate as it leaves the tubes causing the liquor to become separated from the vapors. Vapors pass above the deflector plate through a vapor outlet to an external centrifugal entrainment separator where further juice and vapor separation takes place. Juice after hitting the deflector plate is forced downward to an annular space around the tube bundle to which a juice outlet line is connected. A vapor-liquor seal is maintained in this line by a ball-float device that maintains a level in a small chamber. The juice is then throttled to the next effect.
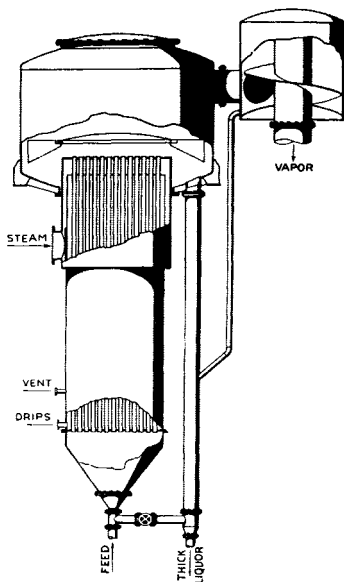

Figure 1.—Long tube vertical evaporator.

Tube sizes and composition are varied depending on the installation and the material to be evaporated. Tube sizes range from 1 to 2 inches in diameter and are from 16 to 24 feet long. Tubes are made from iron, copper, brass or any special alloy as the use demands, and are rolled into the tube sheets.

In the sugar industry steam and juice flow in parallel so that the more concentrated sugar juices are subjected to lower temperatures. First effect steam pressures are usually maintained between 25 and 50 pounds, and last effect vacuum is kept between 20 and 28 inches of Mercury.

Sugar Destruction

When thin juice is exposed to high temperatures for prolonged periods of time it will invert, decompose and caramelize, forming dark-colored juices. However, in the long tube evaporator the retention time is so short that no serious degradation occurs.

Figure 2 shows a calculated time-temperature curve for thin juice as it passes through the thin juice heater and evaporator first effect in the Woodland installation. Many assumptions were made in this estimation; thus the results are only approximations. Experimental work has been conducted, subjecting thin juice to temperatures for the same times as those in the time-temperature curve and the results showed little or no degradation. Figure 3 shows a diagram of the apparatus used. Oil baths were used to simulate the temperatures of the juice for the given periods of time. The resulting time-temperature curve is seen in figure 2 showing the experimental approach to the calculated curve. The experimental data are tabulated in table 1. Slight increases in color of the thin juice were noticed which were due to caramelization, but again, were not significant.
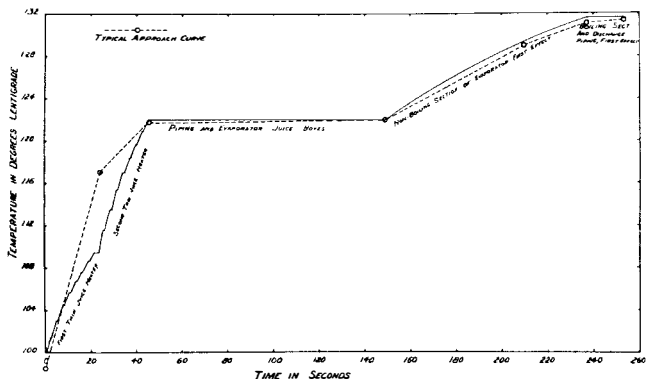

Figure 2.—Solid line, calculated time-temperature curve for thin juice for evaporator first effect and preheaters. Dotted line, experimental time temperature curve showing the approach to the calculated curve.
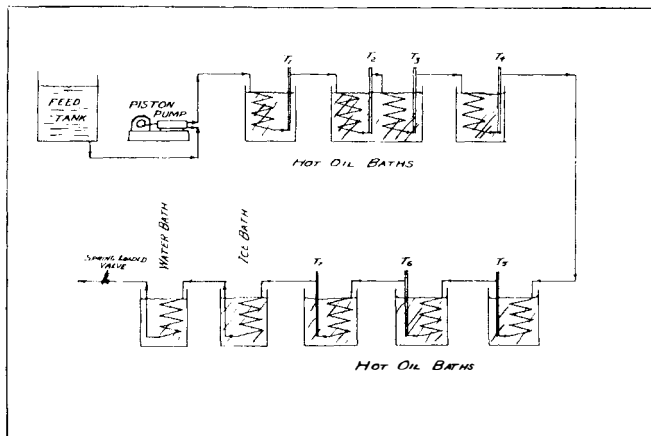
Figure 3.— Experimental apparatus for obtaining time-temperature curves.

Description of the Woodland Installation

Previous statements in this paper have been concerned with long tube vertical evaporators in general, we will now confine ourselves to the installation in Woodland.

At Woodland we have a five effect evaporator manufactured by the Swenson Evaporator Company. Four of the effects are the long tube vertical type and the fifth is a calandria type. The first and second effects each consist of two identical bodies operated in parallel. The remaining effects are single bodies. Evaporator bodies used in the first, second and third effects are of the same size, each having 950·1½ inch O.D. 14 B.W.G. tubes 18 feet long. The fourth effect is somewhat smaller containing only 398 tubes. The calandria fifth effect has 1,120·2½ inch O.D. 12 B.W.G. tubes 4½ feet long. This gives the first effect 12,000 square feet; the second effect 12,000 square feet; the third effect 6,000 square feet; the fourth effect 2,500 square feet; and the fifth effect 3,200 square feet heating surface.

Tube Corrosion

Originally the tubes were made from iron throughout the evaporator, but serious corrosion and pitting of the iron tubes were experienced. Iron tubes had an average life of about 3 years, with many failing much before this time, causing many expensive interruptions in service. After testing

various tubes, the first and second effects are now completely tubed with brass. Some of the brass tubes have been in operation 5 years with very little corrosion. Eventually the entire evaporator will be tubed with brass when our present stock of iron tubes is depleted. The first two effects were considered the most important as any interruption in their service would seriously reduce the capacity of the evaporators.

Steam and Juice Flow System

During operation approximately half of the thin juice is fed to evaporator 1A and half to 1B. Product from 1A feeds 2A and 1B feeds 2B. Products from the second effects are combined to feed the remaining effects in series. See figure 4. Vapors from both first bodies are joined in a common entrainment separator and then are divided to each second body. The same system is used in the second effect.

When one body is out of operation for boil-out, vapors are withdrawn through a 6-inch header to a special boil-out condenser. When one of the first bodies is out of service, some exhaust steam is by-passed to first vapors to make up the deficiency. Similarly some of the feed is led directly to the second effect in order to maintain the capacity.

Vapor Heating

One of the most important advantages of the long tube vertical evaporators is in the overall steam economy resulting from complete utilization of vapors for process heating and evaporation. Table 2 shows an outline of the vapor distribution as used in the Woodland factory. First vapors are used in:

    White pans
    Thin juice heaters

Second vapors are used in:

    Intermediate pans
    Second-carbonation heater
    Thin juice boiler
    Thin juice heater
    High and low melters (open injection)
    Cold waste heater (open injection)
    Saccharate milk heater (open injection)

Third vapors are used in:

    Battery
    Raw juice heater
    Second-carbonation filter heater
    Standard liquor heater
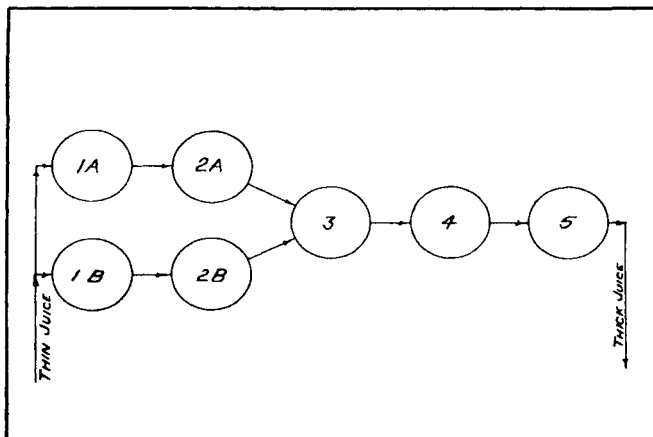   Raw pans

Fourth vapors are used in the raw juice heater.

Figure 4.—Flow diagram for Woodland evaporators.

			-3-	Orig	1	Test No. 2	3	Change
June 11	Sugar Solution unboiled water pH ca-4	Polarization,A " σ % Invert on Poln		13.27 0.03 0.02	13.26 0.05 0.08	13.27 0.01 0.08	13.24 0.02 0.08	Not sign. ∱0.06
June 12	Sugar Solution unboiled water pH ca-4	PolarizationA " σ % Invert on Poln		12.36 0.03 0.02	12.32 0.00 0.12	12.32 0.02 0.13	12.33 0.00 0.13	Not sign. ∱0.11
June 13	Thin Juice	Polarization,A " σ % Invert on Poln		11.54 0.04 0.23	11.50 0.04 0.24	11.52 0.05 0.25	11.54 0.05 0.25	Not sign. ∱0.02
June 17	Sugar Solution unboiled water pH ca.4	Polarization,A " σ % Invert on Poln		12.44 0.03 0.02	12.39 0.05 0.08	12.38 0.02 0.09	12.41 0.04 0.09	Not sign. ∱0.07
June 18	Thin Juice	Polarization,A " σ % Invert on Poln		10.14 0.05 0.26	10.10 0.07 0.26	10.11 0.04 0.26	10.12 0.06 0.26	Not sign. 0.00
June 20	Thin Juice	Polarization,A " σ %Invert on Poln		10.12 0.03 0.28	10.09 0.04 0.32	10.06 0.02 0.32	10.06 0.02 0.32	Not sign. ∱0.04
June 21	Sugar Solution boiled water pH about 7	Polarization,A " σ % Invert on Poln		11.08 0.03 0.02	11.05 0.02 0.04	11.09 0.04 0.05	11.10 0.04 0.05	Not sign. ∱0.03

July 1	Sugar Solution boiled water pH about 7	Polarization,A " %Invert on Poln	12.26 0.02 0.00	12.26 0.03 0.01	12.27 0.04 0.01	12.28 0.03 0.01	Not sign. /0.01

SUMMARY OF AVERAGE RESULTS

	Sugar Sol'n pH 4	Sugar Sol'n pH 7	Thin Juice
Polarization	No change	No change	No change
Increase in invert on pol'n	0.08	0.02	0.02

Table 1. Experimental results for sugar destruction by heat.

At times it has been necessary to by-pass second to third vapors in order to maintain circulation in the raw pans and maintain sufficient heat for proper battery operation. This procedure is not recommended unless it is absolutely necessary because it cuts down the overall steam economy of the factory.

Normal Operating Procedure

The usual method of operation, assuming a constant rate of slice, is to keep the exhaust steam pressure at the maximum allowable figure considering the mechanical and the steam load. This runs between 40 and 45 pounds at Woodland. Live steam is exhausted from about 300 pounds to the exhaust steam pressure through the turbine and is kept at the desired figure by bleeding live steam through a pressure reducing valve to the exhaust steam line. This is possible because the steam load is greater than the mechanical load. When the evaporator becomes so dirty that the mechanical load exceeds the steam load, steam is blown to the roof by a pressure relief valve at 46 to 47 pounds. Generally this condition is not allowed to be reached.

If the evaporators are extremely clean and the operator wishes to reduce the capacity, he first closes down on the amount of water flowing to the jet condenser which reduces the vacuum causing a decrease in the overall temperature drop and thereby reducing the capacity. If further reduction is required, fourth vapors are backed up by throttling the valve to the fifth effect. This reduces the temperature drop across the first effects where most of the evaporation takes place. If still further reduction is necessary the exhaust steam pressure is then reduced.

THEORETICAL HEAT BALANCE
Steffen House - Excess Molasses
(Saccharate Purity 90.5)

1800 Tons Beets &OT Excess Mol
18 Percent Sugar - Total Mol = 151.0T

Item	Station	Sp Ht.	Temp °C Ent	Temp °C Leaves	Temp Dif °C	% On Beets	BTU per 100# Beets	Exh St.	1st Vap	2nd Vap	3rd Vap	4th Vap
	Steam Pressure #/□ Ga							40	30	17	5	8"Hg.
	Steam Temp °F							287	274	254	227	197
	Steam Temp °C							141.3	134.3	123.3	108.3	917
	Latent Temp Vapor (Exhaust 60°SH)							952	931	945	962	991
1	Colorimeters-Cossettes Entering	1.00	30			54	100.0	5400					
	Water Entering	1.00	50			90	242.0	21800					
	Total Heat							27200					
	Diffusion Juice Leaving	.95		60		108	142.0	14400					
	Wet Pulp and Water	1.00		60		108	160.0	21600					
	Total Heat							36200					
	Total Heat Chgs							9000				9.35	
2	Raw Juice Heater	.95	60	72	12	21.6	142.0	2910					2.97
	Raw Juice Heater	.95	72	85	13	23.4	142.0	3160				3.28	
3	Second Carbonation Heater	.95	80	92	12	21.6	174.55	3580			3.79		
4	Second Filter Heater	.95	89	92	3	3.4	174.55	890			.93		
5	Thin Juice Boiler	.95	88	101	13	23.4	174.55	3880		4.11			
6	Thin Juice Heater	.95	96	111	15	27	174.55	4480		4.75			
	Thin Juice Heater	.95	111	126	15	27	174.55	4480	4.82				
7	High and Low Melter (Open Inj)	.60	15	85	50	90	51.49	2780		2.78			
8	Std. Liquor Heater	.60	85	95	10	18	51.49	560				.58	
9	White Pan Evap'n		Latent Heat =				12.53	12430	11.36				
10	Inter Pan Evap'n		Latent Heat =		992		4.22	4190		4.44			
11	Raw Pan Evap'n		Latent Heat =		1018		2.28	2320			2.41		
12	Receiving Tanks		300° 1 1/2'O.D.Cu. Tubing					600	.65				
13	Cold Waste Heating (Open Inj)	1.00	18	90	72	129.6	79.8	10360			10.36		
14	Saccharate Milk Ht (Open Inj)	1.00	40	85	45	81	72.0	3590			3.92		
	Total Vapors % On Beets							.65	18.20	33.82	18.55	5.97	
	Total BTU							16910	32860	15930	2910	

Water Evaporated In Evaporators 137.07   68.30
Steam to Granulators .... .90
Steam To Beet Cutters 1.25
Steam To Turbines (Electrical and Mechanical Power) 6.60
Total Steam % On Beets = 254°F = 17# Ga,L = 945 .... 77.30
Total Steam % On Beets F A A 212 P₁ L = 970 .... 75.30
Lbs. of Steam per Ton of Beets F A A 2129°F 1506
Lbs. of Steam per Bag of Sugar F A A 2129°F 406
Gas = 1040 BTU - 82% Boiler Eff. (Cu.Ft. per Ton Beets) 1710
Gas per Bag of Sugar 96.8% Produced of Total Entered 460
BTU per Bag of Sugar 480,000
Oil Percent On Beets-(18,560 BTU Oil. 86% Boiler Eff.) 3.86

Table 2. Theoretical heat balance. Steffen house, showing vapor utilization.

Heat Balance

In normal operation according to the theoretical heat balance no steam reaches the last effect. Only when an evaporator body is out of service or the capacity of the vapor demand is reduced do we have any heat reaching the fifth effect. In table 3 we have a heat balance of our evaporators made from data taken from 120 days of the last campaign. Note that steam is reaching the last effect in a quantity greater than is available from the fourth effect. This indicates that there is considerable by-pass of vapors and probably errors in the data. This heat balance does not include the additional heat gained from condensate flash, but this is assumed to be approximately equal to the amount of heat lost due to radiation and conduction.

The actual heat transfer coefficients in the first two effects are considerably less than the design coefficients which fact further accounts for the vapor pressures being lower than anticipated.

HEAT BALANCE, HOUSE EVAPORATORS

120 DAYS 1947 CAMPAIGN, SPRECKELS SUGAR CO., WOODLAND

2100 T., 620 GPM AT 250°F, 13.0 RDS, 87 APC, VACUUM, 21.3", 155.5°F

ASSUME HEAT LOSS TO BE EQUAL TO HEAT GAINED BY
CONDENSATE FLASH.

EFFECT	I	II	III	IV	V
CHEST PRESSURE	40.2#	22.9#	10.1#	2.1#	9.1"
TEMP. OF STEAM, °F.	286.9	263.5	239.6	218.8	194.0
HEAT TO CHEST-BTU/HR x 10⁻³	121,954	72,560	19,905	6,520	9,653
TEMP. OF LIQUOR OUT, °F.	264.5	241.6	221.5	197.0	158.9
AMOUNT OF FLASH, °F.	-14.5	22.9	20.1	24.5	38.1
JUICE TO EVAP. #/HR x 10⁻³	325.8	200.3	119.6	96.78	88.20
SPECIFIC HEAT	0.95	0.93	0.85	0.80	0.77
HEAT IN FLASH BTU/HR x 10⁻³	-4,486	4,266	2,043	1,897	2,588
HEAT AVAILABLE FOR EVAP. BTU/HR. x 10⁻³	117,468	76,826	21,948	8,417	12,241
WATER EVAP. #/HR x 10⁻³	125.5	80.7	22.72	8.58	12.18
JUICE FROM EVAP. #/HR x 10⁻³	200.3	119.6	96.78	88.20	76.02
R.D.S. OUT	21.2	35.4	43.7	48.0	55.7
HEAT OFF TO PROCESS BY DIFF. BTU/HR x 10⁻³	44,908	56,921	15,421	-1,236	——
HEAT OFF TO PROCESS THEORETICAL, BTU/HR x 10⁻³	29,600	57,500	18,000	5,090	——
ΔT, TEMP. DIFF. °F.	22.4	21.9	18.1	21.8	35.1
HEATING AREA, SQ. FT.	12,000	12,000	6,000	2,500	3,200
UΔT, BTU/HR/SQ.FT.	10,163	6,047	3,317	2,608	3,017
U, HEAT TRANS. COEFF. BTU/HR/SQ.FT./°F	454	276	183	120	86
ASSUMED B.P.R. AND PRESSURE LOSS, °F.	1.0	2.0	2.7	3.0	3.4
G.P.M. PER TUBE	0.336	0.194	0.207	0.407	——
BOILOUTS	7	8	6	4	4

Table 3.—Heat balance, Woodland evaporators for 1947 campaign.

Factors Affecting Heat Transfer

Why the low heat transfer coefficients? This may be due to several different causes. One might be the failure to completely remove the non-condensable gases. The general agreement between evaporator experts is that non-condensable gases are swept along with the steam and can be most effectively removed at the end of their path. This is the point in the

long tube evaporator where all of the vapors have been condensed and is located at the bottom of the steam chest. Many operators feel that the non-condensables should be removed at the top of the steam chest.

The second reason, and the more likely, is the very rapid scaling rate which may be caused to a certain extent by improper design. Note that in table 3 the calculated flow rate in GPM per tube is approximately 1/3, 1/10, 1/10, 2/5 in the first four effects, respectively. In some recent pilot evaporator experiments, which are however not conclusive, it was shown that the evaporator tubes scale at an accelerated rate at the low flow rates, with 1/3 GPM per tube being about the lowest rate that can be handled with the normal scaling rate. To further aggravate the situation there is the possibility of the channeling of the flow to certain tubes leaving the others in comparative dryness. The main reason for increased scaling at these low flow rates is that the juice has an opportunity to "stew" and reach higher temperatures and higher local concentration which in turn causes scaling. In some pilot evaporator studies we have obtained as much as three times the heat transfer coefficients by operating at higher flow rates with very slight scaling. Values of 700, 800, 700, 450 were not uncommon in the first four effects, respectively, after many hours of operation. Note that the factory averaged only 500, 300, 200, and 100. However, channeling is no problem in our pilot evaporator because of its size. One way to increase the flow rate per tube and yet maintain the same overall heating area is to decrease the number of tubes and increase their length.

Scale and Scale Removal

The scale found in our evaporator tubes is similar to that found in any beet sugar evaporator consisting mainly of calcium oxalate, carbonate and sulfate. The exact composition of course, is dependent on the nature of the beets being handled.

Scale in the long tube evaporator cannot be readily removed by mechanical methods and thus chemical procedures are used. Based on experience the following method has been found satisfactory in most instances: After removal from service the evaporator body is filled with a 4-percent caustic soda and soda-ash solution and is boiled for 3 or 4 hours allowing the solution to circulate up through the tubes and down the discharge line and so on. It is then followed by a water rinse to wash out the remaining traces of caustic. Now it is boiled an additional 3 hours on a 3-percent inhibited hydrochloric acid solution. After the final rinse the evaporator tubes are generally clean.

Summary

Summarizing our experiences with the long tube vertical evaporator we have found them to be entirely satisfactory in spite of their lower than anticipated heat transfer coefficients because of their great overall saving in steam. No doubt with further experimental work the reasons and the remedies for these discrepancies can be found and a better design can be made.