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

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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.

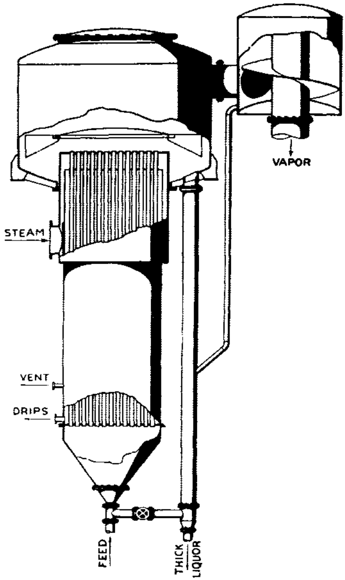

Figure 1.—Long tube vertical evaporator.

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.

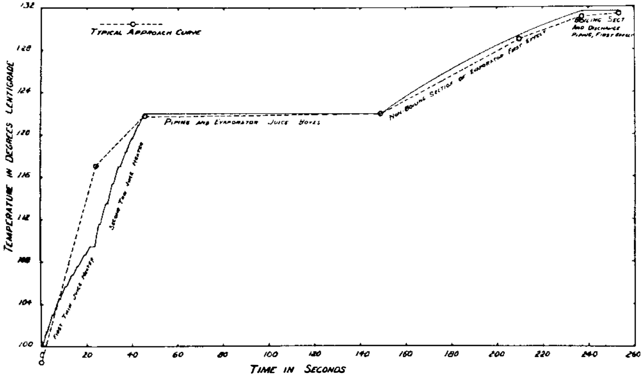

Figure 2.—Solid line, calculated time-temperature curve for thin juice for evaporator first effect and preheaters. Dotted line, experimental time temperature curve showing the approach to the calculated curve.

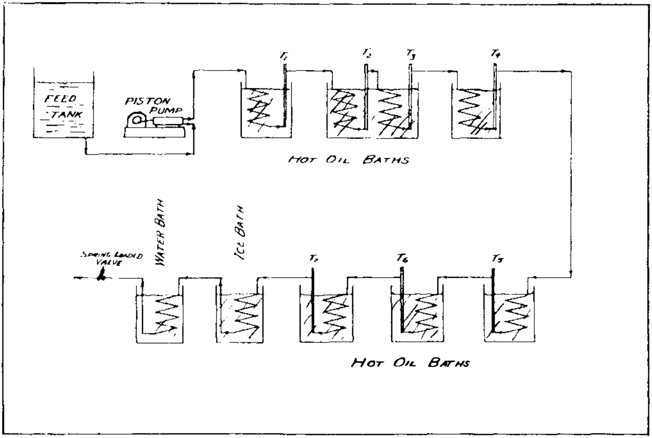

Figure 3.—Experimental apparatus for obtaining time-temperature curves.

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.

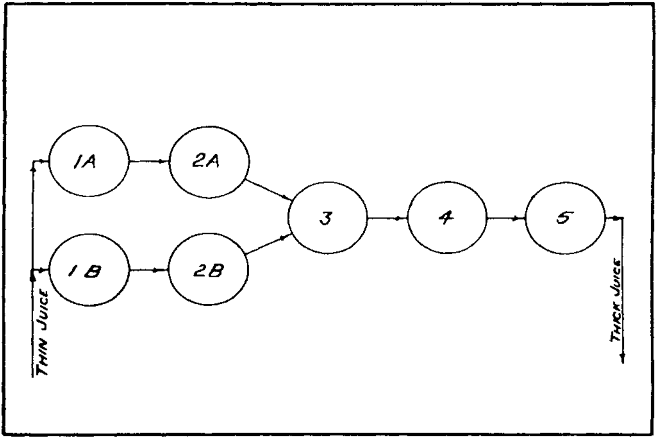

Figure 4.—Flow diagram for Woodland evaporators.

-3-

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

July 1	Sugar Solution boiled water pH about 7	Polarization,A	12.26	12.26	12.27	12.28	Not
		" 6	0.02	0.03	0.04	0.03	sign.
		%Invert on Poln	0.00	0.01	0.01	0.01	/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 90.3)

1800 Tons Beets 40T Excess Mol
18 Percent Sugar - Total Mol = 151.0T

Item	Station	Sp Ht.	Temp °C Ent	Temp °C Leave	Temp Diff/ °C	Temp Diff/ °F	% 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 Heat Vapor (Exhaust 60°SH)								952	931	945	962	981
1	Calorizators-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	14600					
	Wet Pulp and Water	1.00		60		108	200.0	21600					
	Total Heat							36200					
	Total Heat Input							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	5.4	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	35	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 =			992		12.53	12430		13.38			
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			3.59		
	Total Vapors % On Beets								.65	18.20	33.82	16.55	2.97
	Total BTU									16910	32860	15930	2910
	Water Evaporated In Evaporators						137.07		68.30				
	Steam to Granulators								.50				
	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 212 F., L = 970								75.30				
	Lbs. of Steam per Ton of Beets F & A 212°F								1506				
	Lbs. of Steam per Bag of Sugar F & A 212°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								461				
	BTU per Bag of Sugar								480,000				
	Oil Percent On Beets-(18,500 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
$U\Delta T$, 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.