Investment Decisions

Using

The *SUGARS* Computer Program

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By:

L. Warner Weiss

Equipment Sales Company
10129 E. Powers Avenue
Englewood, Colorado 80111
(303) 770-2630

INTRODUCTION

One of the many difficult problems that a sugar factory manager faces is deciding how best to employ capital funds to improve his factory. Usually, there are a number of areas that need attention and it can be perplexing to decide which investment will yield the greatest return on invested funds.

Some decisions do not impact the process efficiency of the factory and are selected for investment based on their own merit; e.g., a new factory office. The investment decisions discussed in this paper are process decisions related to improving sugar extraction efficiency and/or reducing energy consumption. These decisions are amenable to analytical analysis by computer modeling of the proposed process improvements and comparing the predictions from modeling with actual operating results.

The *SUGARS* Computer Program is a powerful tool for evaluating sugar processes by modeling to give simulated results under different operating conditions. Many process alternatives that affect the efficiency of a factory can be modeled using *SUGARS* and each alternative can be prioritized based on the expected revenue increase that is predicted by the model. Modeling can also be used to followup on investments to see if the expected results were achieved; and if not, to evaluate what has changed. As is well known, after an investment is made, sometimes it is difficult to ascertain whether the expected results were realized because of changes in another area that may mask the effects; e.g., changes in beet quality from one campaign to the next.

This paper describes a model (base case) of the evaporators and sugar end for a fictitious beet sugar factory, and then gives three examples of changes to this base model with simulated results for each example that are used to predict the revenue gains. Return on investment calculations can then be done following determination of the necessary capital funds to accomplish each change. The predictions shown in this paper are peculiar to the base model only and may not reflect the results that could be expected in an actual factory. Each factory is different and has to be modeled for its own unique characteristics.

BASE MODEL

SUGARS is a comprehensive computer program that is used to evaluate different process concepts and/or equipment efficiencies in a sugar factory[1]. It is completely flexible regarding the type of factory that it can analyze because it uses individual station models that can be arranged in almost any order to fit the process being simulated. It is only necessary that a combination of the basic station models available in *SUGARS* can be used to simulate the performance of actual stations used in the factory. After the model is constructed and all of the data is entered, *SUGARS* will calculate the complete heat and material balance for different process alternatives to assist management with making the best investment and process decisions.

In addition, *SUGARS* does a complete performance evaluation for each centrifugal station in the flow diagram by analyzing the operating results from actual centrifugals used in the factory. The evaluations give valuable information regarding the amount of wash water and massecuite mother liquor that is purged, and sugar crystals that are lost due to melting and passage through the centrifugal screen (undersized crystals).

Simulation results from *SUGARS* can be used as a guide to compare against the actual performance of a factory. Differences between the simulated results and actual operating results can be explored to determine the cause of the difference, which in many cases can lead to a discovery of how the factory could be operated in a more efficient manner.

Figure 1 shows the flow diagram for evaporation and sugar end operations of a fictitious beet sugar factory.

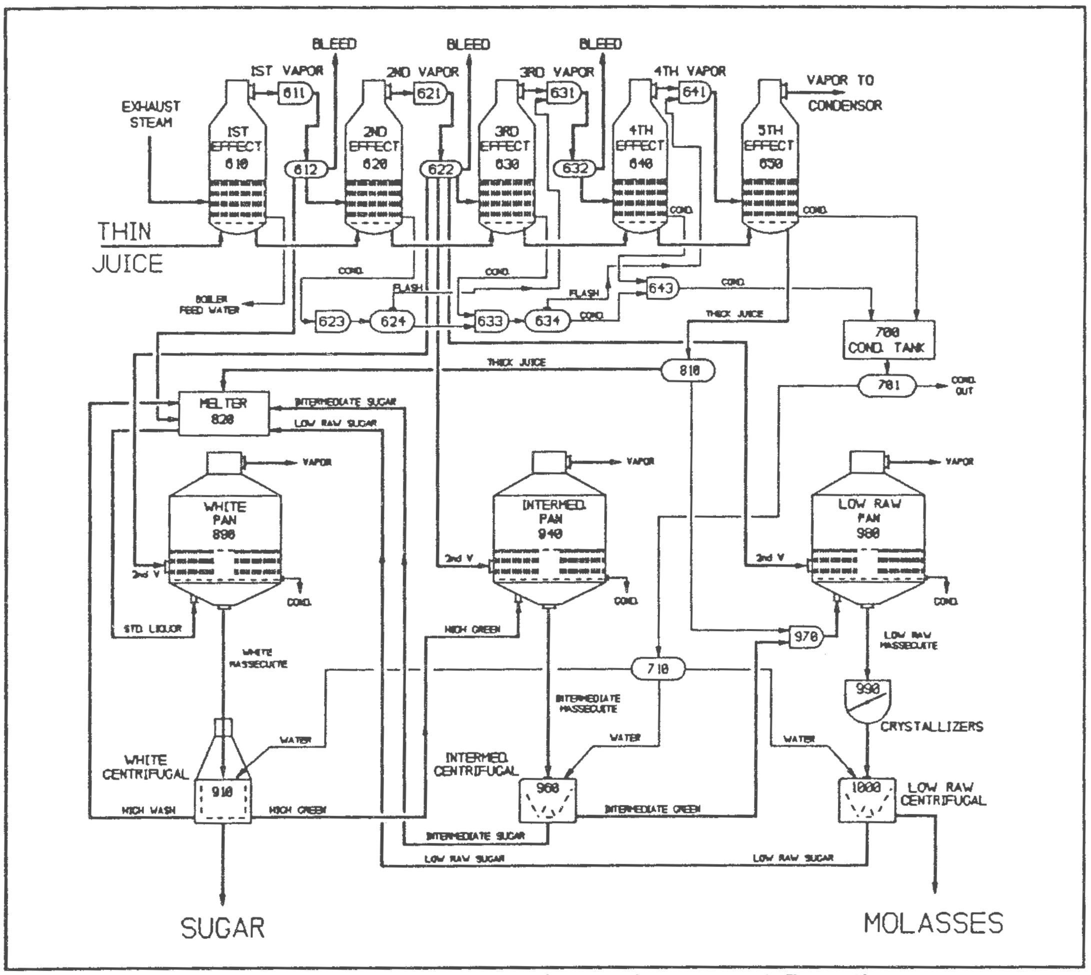

Figure 1. Base model evaporation and sugar end flow diagram.

The flow diagram consists of individual *station models* with interconnecting flows called *internal flows*, and *external flows* that go into the process from outside sources (e.g., steam, dilution water, juice from storage, the main juice flow, etc). Individual station models are mathematical models available in *SUGARS* that are used to simulate actual stations in the process being analyzed.

As shown in Figure 1, there are only two external flows: (1) thin juice and (2) exhaust steam to the 1st effect. All other flows are internal flows that go from station-to-station, or leave the flow diagram (e.g., vapors, condensates, molasses and sugar). Each block with a station number is a station model from *SUGARS* that has been assigned a specific purpose. For example: station no. 890 is a pan station model that is used to process a syrup into a massecuite; station no. 710 is a distributor station model that splits a flow stream into two, or more, output flows (three output flows in this case); station no. 970 is a receiver station model that combines several input flows (two in this case) into one output flow; etc.

SUGARS processes all external flows into the flow diagram using station models to provide output flows from each station. The flows out of each station are processed by other stations until all of the material and heat entering the flow diagram is accounted for by the flows and losses leaving the flow diagram.

The factory shown in Figure 1 has a five effect evaporator station to concentrate thin juice. Condensate is flashed from the 2nd effect to 3rd vapor while the remaining 2nd effect condensate is combined with 3rd effect condensate and flashed to 4th vapor. 1st effect condensate is returned as boiler feed water, and all remaining condensates from the 2nd thru 5th effects are sent to a condensate tank (station no. 700) which supplies centrifugal wash water to the centrifugal stations by way of two distributor stations. Excess condensate from the condensate tank leaves the flow diagram from distributor station no. 701.

Allowance has been made for 1st, 2nd and 3rd vapor bleeds going to other vapor users in the factory that aren't included in the flow diagram. All three pans use 2nd vapor while the melter (station no. 820) uses 1st vapor for injection heating.

Thick juice from the 5th effect (station no. 650) goes to a distributor station (no. 810). The distributor station can send a portion of its flow to the low raw pan (station no. 980) if necessary; however, in the simulations that follow, all of the thick juice goes to the melter.

Both low raw sugar and intermediate sugar go directly to the melter. This is somewhat idealized, since in actual practice, low raw sugar usually can't go directly to the melter because of color problems with the white sugar; however, it is assumed for simplicity that color isn't a problem in any of the examples that follow.

All external and internal flow streams in a flow diagram analyzed by *SUGARS* have as properties of the flow: pressure, temperature, weight flow, eleven (11) flow components and four (4) coefficients for the solubility coefficient equation. The eleven (11) flow stream components consist of four liquid components (water, sucrose and two non-sugars), four solid components (sucrose crystals, $CaCO_3$, CaO and other) and three gas components (water vapor, CO_2 and NH_3). Three phase flow is considered for each flow stream and material may be transferred between the liquid, solid and gas/vapor phases depending on the features of the station for crystallizing, melting, condensing, vaporizing, evaporating, separating, reacting and/or drying. Numeric values for the properties of flow streams are updated and modified by station models in the flow diagram depending on the characteristics of each station and its input flow.

SUGARS will calculate the quantity and characteristics of all internal flows and output flows leaving the flow diagram, while the external flows must be defined in the input data. Only the characteristics (pressure, temperature and flow components) need to be defined for the steam (*SUGARS* will calculate the flow quantity); whereas, both the characteristics and quantity of the thin juice flow must be input. The external flows and characteristics of each station model are entered using the input screens in *SUGARS*.

The results of the heat and material balance calculations can be displayed on the monitor, sent to a printer, or written to data files. All of the *SUGARS* data files for a factory are in standard text format to allow other computer programs to either write to the data files, or read the calculated results for manipulation and display by another program. A separate data acquisition system with appropriate software can be used to automatically record the necessary factory data and make entries into the *SUGARS* data files.

BEET SUGAR END WITH EVAPORATION

27.01.91 SUGARS V2.50 MATERIAL BALANCE RESULTS (SI UNITS) 17:00:48

%DS, PURITY & %CRYS. ARE ONLY FOR SOLUBLE PORTION OF FLOW

EXTERNAL FLOWS INTO FACTORY

FLOW	NAME	GOES TO	TONS/HR	%TDM	%SUGAR	%DS	PURITY	%CRYS.	TEMP(C)	%ISNS	%GAS
1	THIN JUICE	610	320.00	14.15	12.90	14.15	91.17	0.00	128.0	0.00	0.00

STATION FLOWS WITHIN FACTORY

STATION	NAME	GOES TO	TONS/HR	%TDM	%SUGAR	%DS	PURITY	%CRYS.	TEMP(C)	%ISNS	%GAS
610	1ST EFFECT	620	219.09	20.66	18.84	20.66	91.17	0.00	135.7	0.00	0.00
620	2ND EFFECT	630	129.19	35.04	31.94	35.04	91.17	0.00	124.3	0.00	0.00
630	3RD EFFECT	640	96.45	46.93	42.79	46.93	91.17	0.00	111.5	0.00	0.00
640	4TH EFFECT	650	87.96	51.46	46.92	51.46	91.17	0.00	99.5	0.00	0.00
650	5TH EFFECT	810	73.01	62.00	56.52	62.00	91.17	0.00	70.6	0.00	0.00
810	THICK JUICE DISTRI.										
	OUTPUT #1	820	73.01	62.00	56.52	62.00	91.17	0.00	70.6	0.00	0.00
	OUTPUT #2	970	0.00	0.00	0.00	0.00	0.00	0.00	0.0	0.00	0.00
820	MELTER	890	110.17	69.63	64.80	69.63	93.07	0.00	92.0	0.00	0.00
890	WHITE PAN	910	83.83	91.50	85.16	91.50	93.07	53.11	75.0	0.00	0.00
910	WHITE CENTRIFUGALS										
	GREEN	940	49.28	76.91	66.49	76.91	86.45	0.00	69.5	0.00	0.00
	WASH	820	5.77	73.33	70.25	73.33	95.80	0.00	67.0	0.00	0.00
	SUGAR	OUT	**35.12**	98.46	98.45	98.46	99.99	93.79	64.0	0.00	0.00
940	INTERMEDIATE PAN	960	40.75	93.00	80.40	93.00	86.45	47.46	81.0	0.00	0.00
960	INTERMED. CENTRIF.										
	GREEN	970	23.58	83.69	64.33	83.69	76.87	4.90	75.5	0.00	0.00
	SUGAR	820	18.62	97.56	94.50	97.56	96.86	85.82	72.9	0.00	0.00
970	LOW RAW PAN RECVR.	980	23.58	83.69	64.33	83.69	76.87	4.90	75.5	0.00	0.00
980	LOW RAW PAN	990	21.10	93.50	71.87	93.50	76.87	32.11	85.0	0.00	0.00
990	L.R. CRYSTALLIZERS	1000	21.10	93.50	71.87	93.50	76.87	44.61	54.0	0.00	0.00
1000	LOW RAW CENTRIFUGALS										
	GREEN	OUT	**13.28**	80.51	50.42	80.51	**62.63**	0.00	48.7	0.00	0.00
	SUGAR	820	9.40	96.15	90.10	96.15	93.71	79.43	46.9	0.00	0.00

Figure 2. Base model printout of material flows.

Printed results from *SUGARS* are given in two sections: the first printout is for all of the material flow streams, and the second printout is for all of the heating flow streams in the flow diagram.

Figure 2, shows the printout of the material flows for the base model. External flows into the flow diagram are listed first (e.g., THIN JUICE) and internal flows between stations and leaving the flow diagram are given next. The thin juice flow rate corresponds to a factory of about 6,000 metric tons per day beet slice rate. Numbers in this figure that are underlined and in bold are: total sugar production of **35.12** metric tons per hour, total molasses production of

13.28 metric tons per hour and 62.63 for the molasses purity. These numbers will be referred to later as changes are made to the base model.

BEET SUGAR END WITH EVAPORATION

27.01.91 SUGARS V2.50 VAPOR & CONDENSATE CALCULATION RESULTS (SI UNITS) 17:00:51

EXTERNAL VAPOR & CONDENSATE FLOWS INTO FACTORY

FLOW	NAME	GOES TO	kPa	TEMP(C)	FLOW(kg/h)	%COND.	%TDM	%SUGAR	%VAPOR	%GASES
2	EXHAUST STEAM	610	416.5	145.0	107837.5	0.00	0.00	0.00	100.00	0.00

VAPOR & CONDENSATE FLOWS WITHIN FACTORY

STATION	NAME	GOES TO	kPa	TEMP(C)	FLOW(kg/h)	%COND.	%TDM	%SUGAR	%VAPOR	%GASES
610	1ST EFFECT									
	VAPOR	611	315.0	135.2	100913.5	0.03	0.01	0.01	99.96	0.00
	CONDENSATE	OUT	416.5	145.0	107837.5	100.00	0.00	0.00	0.00	0.00
611	1ST VAPOR RECEIVER	612	315.0	135.2	100913.5	0.03	0.01	0.01	99.96	0.00
612	1ST VAPOR DISTRI.									
	VAPOR #1	OUT	315.0	135.2	10000.0	0.03	0.01	0.01	99.96	0.00
	VAPOR #2	820	315.0	135.2	3361.2	0.03	0.01	0.01	99.96	0.00
	VAPOR #3	620	315.0	135.2	87554.8	0.03	0.01	0.01	99.96	0.00
620	2ND EFFECT									
	VAPOR	621	219.9	123.3	89895.9	0.01	0.01	0.00	99.98	0.00
	CONDENSATE	623	315.0	135.2	87554.8	99.99	0.01	0.01	0.00	0.00
621	2ND VAPOR RECEIVER	622	219.9	123.3	89895.9	0.01	0.01	0.00	99.98	0.00
622	2ND VAPOR DISTRI.									
	VAPOR #1	890	219.9	123.3	26342.5	0.01	0.01	0.00	99.98	0.00
	VAPOR #2	940	219.9	123.3	9877.3	0.01	0.01	0.00	99.98	0.00
	VAPOR #3	980	219.9	123.3	2898.9	0.01	0.01	0.00	99.98	0.00
	VAPOR #4	OUT	219.9	123.3	20000.0	0.01	0.01	0.00	99.98	0.00
	VAPOR #5	630	219.9	123.3	30780.0	0.01	0.01	0.00	99.98	0.00
623	2ND COND. RECEIVER	624	315.0	135.2	87554.8	99.99	0.01	0.01	0.00	0.00
624	2ND COND. FLASH TANK									
	CONDENSATE	633	142.3	109.8	83390.4	99.99	0.01	0.01	0.00	0.00
	VAPOR	631	142.3	109.8	4164.4	0.00	0.00	0.00	100.00	0.00
630	3RD EFFECT									
	VAPOR	631	142.3	109.8	32744.1	0.00	0.00	0.00	99.99	0.00
	CONDENSATE	633	219.9	123.3	30780.0	99.99	0.01	0.00	0.00	0.00
631	3RD VAPOR RECEIVER	632	142.3	109.8	36908.5	0.00	0.00	0.00	99.99	0.00
632	3RD VAPOR DISTRI.									
	VAPOR #1	OUT	142.3	109.8	30000.0	0.00	0.00	0.00	99.99	0.00
	VAPOR #2	640	142.3	109.8	6908.5	0.00	0.00	0.00	99.99	0.00
633	3RD COND. RECEIVER	634	142.3	113.4	114170.4	99.99	0.01	0.01	0.00	0.00
634	3RD COND. FLASH TANK									
	CONDENSATE	643	92.8	97.6	110825.7	99.99	0.01	0.01	0.00	0.00
	VAPOR	641	92.9	97.6	3344.7	0.00	0.00	0.00	100.00	0.00
640	4TH EFFECT									
	VAPOR	641	92.9	97.6	8490.1	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	643	142.3	109.8	6908.5	100.00	0.00	0.00	0.00	0.00
641	4TH VAPOR RECEIVER	650	92.9	97.6	11834.7	0.00	0.00	0.00	100.00	0.00
643	4TH COND. RECEIVER	700	92.8	98.3	117734.2	99.99	0.01	0.01	0.00	0.00
650	5TH EFFECT									
	VAPOR	OUT	28.5	68.0	14947.3	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	700	92.9	97.6	11834.7	100.00	0.00	0.00	0.00	0.00
700	COND. TANK	701	92.8	98.3	129569.0	99.99	0.01	0.01	0.00	0.00
701	CONDENSATE DISTRI.									
	CONDENSATE #1	710	92.8	98.3	9362.9	99.99	0.01	0.01	0.00	0.00
	CONDENSATE #2	OUT	92.8	98.3	120206.1	99.99	0.01	0.01	0.00	0.00
710	CENT. H2O DISTRI.									
	CONDENSATE #1	910	92.8	98.3	6336.9	99.99	0.01	0.01	0.00	0.00
	CONDENSATE #2	960	92.8	98.3	1447.9	99.99	0.01	0.01	0.00	0.00
	CONDENSATE #3	1000	92.8	98.3	1578.1	99.99	0.01	0.01	0.00	0.00
890	WHITE PAN									
	VAPOR	OUT	25.5	65.4	26336.6	0.00	0.01	0.00	99.99	0.00
	CONDENSATE	OUT	219.9	123.3	26342.5	99.99	0.01	0.00	0.00	0.00
940	INTERMEDIATE PAN									
	VAPOR	OUT	28.6	68.0	8528.1	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	OUT	219.9	123.3	9877.3	99.99	0.01	0.00	0.00	0.00
980	LOW RAW PAN									
	VAPOR	OUT	30.0	69.1	2474.6	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	OUT	219.9	123.3	2898.9	99.99	0.01	0.00	0.00	0.00

Figure 3. Base model printout of heating flows.

Printout of the heating flows (i.e., steam, vapor, condensate, etc) is given next by *SUGARS*, The results from *SUGARS* show the changes to the sugar extraction efficiency for the sugar end and to the vapor consumption for the melter, pans and evaporator bodies while considering: the lower temperature of the flow streams due to the heat lost in the intermediate crystallizer, the heat of crystallization due to sucrose crystal growth, the specific heats of the flow streams and the quantity of the flows.

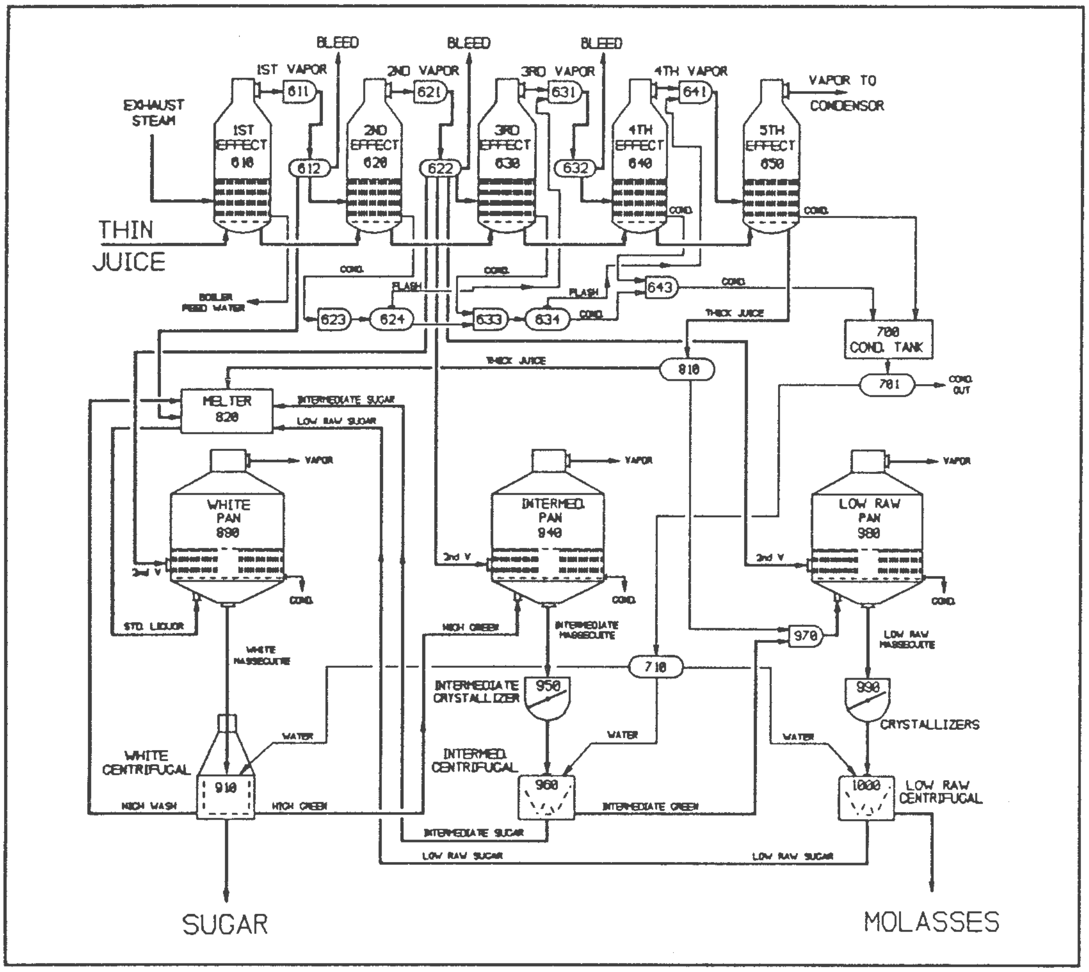

Figure 4. Evaporation and sugar end with intermediate crystallizer.

Printouts for both the material and heating flows are shown in Figures 5 and 6, respectively. Portions of each printout aren't shown because the changes are minor and not significant to the discussion.

As shown in the material balance printout, the sugar production increases to 35.44 metric tons per hour, while the molasses quantity decreases to 12.82 metric tons per hour and the purity drops to 61.49 (a 1.14 point drop). Exhaust steam consumption increases to 108,183 kilograms per hour. White pan vapor consumption (refer to the condensate flow out quantity) stays almost constant; while, the vapor consumption of the intermediate and low raw pans increases. The largest increase occurs in the low raw pan vapor needs, because the syrup feed to this pan is 8.3°C cooler than it was in the base model. Also, note that because more evaporation is done in the second effect, the amount of evaporation needed in the fifth effect is reduced; and hence, the difference in temperature between syrup leaving, and heat-

BEET SUGAR END WITH EVAPORATION & INTERMED. CRYS.

27.01.91 SUGARS V2.50 MATERIAL BALANCE RESULTS (SI UNITS) 17:02:33

%DS, PURITY & %CRYS. ARE ONLY FOR SOLUBLE PORTION OF FLOW

EXTERNAL FLOWS INTO FACTORY

FLOW	NAME	GOES TO	TONS/HR	%TDM	%SUGAR	%DS	PURITY	%CRYS.	TEMP(C)	%ISNS	%GAS
1	THIN JUICE	610	320.00	14.15	12.90	14.15	91.17	0.00	128.0	0.00	0.00

STATION FLOWS WITHIN FACTORY

STATION	NAME	GOES TO	TONS/HR	%TDM	%SUGAR	%DS	PURITY	%CRYS.	TEMP(C)	%ISNS	%GAS
⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮
890	WHITE PAN	910	84.49	91.50	85.21	91.50	93.13	53.17	75.0	0.00	0.00
910	WHITE CENTRIFUGALS										
	GREEN	940	49.62	76.89	66.54	76.89	86.54	0.00	69.5	0.00	0.00
	WASH	820	5.82	73.34	70.28	73.34	95.84	0.00	67.0	0.00	0.00
	SUGAR	OUT	35.44	98.46	98.45	98.46	99.99	93.80	64.0	0.00	0.00
940	INTERMEDIATE PAN	950	41.02	93.00	80.48	93.00	86.54	47.60	81.0	0.00	0.00
950	INTERMED. CRYSTALLZR	960	41.02	93.00	80.48	93.00	86.54	51.52	71.0	0.00	0.00
960	INTERMED. CENTRIF.										
	GREEN	970	22.53	82.81	62.55	82.81	75.53	6.12	67.2	0.00	0.00
	SUGAR	820	19.95	97.71	94.85	97.71	97.08	87.54	64.0	0.00	0.00
970	LOW RAW PAN RECVR.	980	22.53	82.81	62.55	82.81	75.53	6.12	67.2	0.00	0.00
980	LOW RAW PAN	990	19.95	93.50	70.62	93.50	75.53	29.91	85.0	0.00	0.00
990	L.R. CRYSTALLIZERS	1000	19.95	93.50	70.62	93.50	75.53	42.71	54.0	0.00	0.00
1000	LOW RAW CENTRIFUGALS										
	GREEN	OUT	12.82	80.91	49.75	80.91	61.49	0.00	48.7	0.00	0.00
	SUGAR	820	8.63	96.03	89.43	96.03	93.13	78.28	46.9	0.00	0.00

Figure 5. Base model with intermediate crystallizer - material flows.

BEET SUGAR END WITH EVAPORATION & INTERMED. CRYS.

27.01.91 SUGARS V2.50 VAPOR & CONDENSATE CALCULATION RESULTS (SI UNITS) 17:02:35

EXTERNAL VAPOR & CONDENSATE FLOWS INTO FACTORY

FLOW	NAME	GOES TO	kPa	TEMP(C)	FLOW(kg/h)	%COND.	%TDM	%SUGAR	%VAPOR	%GASES
2	EXHAUST STEAM	610	416.4	145.0	108183.4	0.00	0.00	0.00	100.00	0.00

VAPOR & CONDENSATE FLOWS WITHIN FACTORY

STATION	NAME	GOES TO	kPa	TEMP(C)	FLOW(kg/h)	%COND.	%TDM	%SUGAR	%VAPOR	%GASES
⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮
650	5TH EFFECT									
	VAPOR	OUT	29.4	68.7	14657.6	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	700	93.4	97.8	11605.9	100.00	0.00	0.00	0.00	0.00
⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮	⋮
890	WHITE PAN									
	VAPOR	OUT	25.5	65.5	26345.2	0.00	0.01	0.00	99.99	0.00
	CONDENSATE	OUT	219.4	123.2	26326.3	99.99	0.01	0.00	0.00	0.00
940	INTERMEDIATE PAN									
	VAPOR	OUT	28.6	68.0	8598.6	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	OUT	219.4	123.2	9955.7	99.99	0.01	0.00	0.00	0.00
980	LOW RAW PAN									
	VAPOR	OUT	29.7	68.9	2574.7	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	OUT	219.4	123.2	3250.4	99.99	0.01	0.00	0.00	0.00

Figure 6. Base model with intermediate crystallizer - heating flows.

ing vapor entering, the fifth effect is smaller to accomplish the necessary heat exchange between the heating vapor and syrup flow streams which results in a slightly higher temperature for the syrup and vapor leaving the last effect.

The increase in sugar production due to the addition of an intermediate crystallizer is .32 metric tons per hour. At the current U.S. price for sugar of $23.00 per (cwt) bag (or, $507 per metric ton), the net revenue gain from increased sugar production is <u>$3,894</u> per day.

Molasses production decreases by .46 metric tons per hour. Using a current U.S. price for molasses of $55.00 per U.S. ton (or, $61 per metric ton), the net revenue loss from molasses sales amounts to <u>$673</u> per day.

Steam consumption increases by 346 kgs/hour, and assuming the cost of steam is $3.00 per 1,000 pounds (or, $6.60 per metric ton), the additional expenses for steam will amount to <u>$55</u> per day. Thus, the net revenue benefit from adding an intermediate crystallizer to the base model is:

Additional sugar revenues (+.32 MTPH)	$3,894
Less molasses decrease (-.46 MTPH)	-673
Less additional steam costs (+346 kgs/hour)	-55
Net gain per day	$3,166

The above net daily gain in revenues due to a new intermediate crystallizer would amount to <u>$379,920</u> for a 120 day campaign if no other changes were made to the process and the operating conditions remained the same as in the base model.

VACUUM CRYSTALLIZER

A vacuum crystallizer can be used to extract additional sucrose from the mother liquor in a massecuite by giving sucrose crystals additional growing time while adding syrup to the massecuite and flashing water vapor out of the crystallizer to control the dry substance.

Figure 7 shows the installation of a vacuum crystallizer on white massecuite in the base model flow diagram. This arrangement uses a *SUGARS* blender station model (station no. 895) to blend high green with white massecuite. The high green syrup is heated in a heat exchanger (station no. 930) before it is blended with the massecuite, and the combined flow is sent to a crystallizer under vacuum where water from the massecuite and syrup flashes and leaves as vapor. Two *SUGARS* station models are used to model the vacuum crystallizer: a flash station model (station no. 900) and a crystallizer station model (station no. 901). The flash station reduces the temperature of the massecuite to 60°C, and the crystallizer station controls the supersaturation of the massecuite leaving the vacuum crystallizer, which in this example is held to 1.05 supersaturation (white massecuite supersaturation is 1.10 after it leaves the pan).

Figure 8 shows the complete flow diagram with the use of 1st vapor (condensate could be used instead) to heat the high green to 85°C in a heat exchanger before it is blended with white massecuite. The quantity of high green is controlled by the blender station model (station no. 895) which is set to a specified purity for the flow leaving the blender. For this example, the purity is specified to be 90.70 to give a high green flow that is a ratio of approximately 20% to the white massecuite flow.

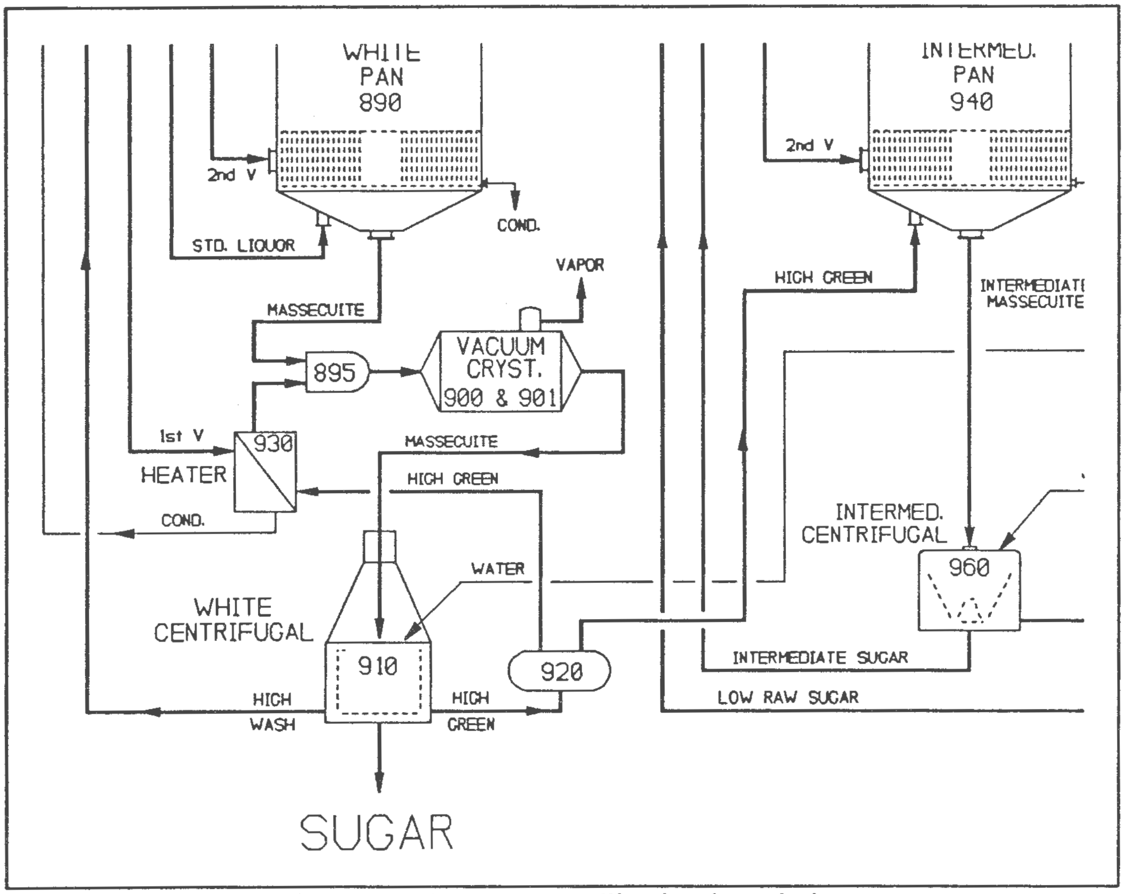

Figure 7. Vacuum crystallization installation.

High wash, or standard liquor, could be used in place of high green for the syrup that is blended with the white massecuite. If high wash is used, it would be possible to eliminate the recycle of high wash to the melter[2]. Each of the possible blend flows has its advantages, and disadvantages; however, for simplicity in making a comparison to the base model, high green was chosen for this example. Separate simulations could be done using *SUGARS* for each of the other possibilities.

Also, using *SUGARS*, it is very easy to evaluate changes in the quantity of high green (or another syrup) that is blended with the white massecuite and its effect on sugar production and steam consumption, or to evaluate changes in the blend syrup temperature and vacuum pressure for the vacuum crystallizer.

One of the advantages of a vacuum crystallizer is the reduction in syrup quantity that is processed by the pans; and consequently, the sugar end can handle additional capacity when a vacuum crystallizer is installed. Also, the sucrose crystal yield will increase in the white massecuite when a vacuum crystallizer is installed. Both of these effects are shown in the material balance printout given in Figure 9.

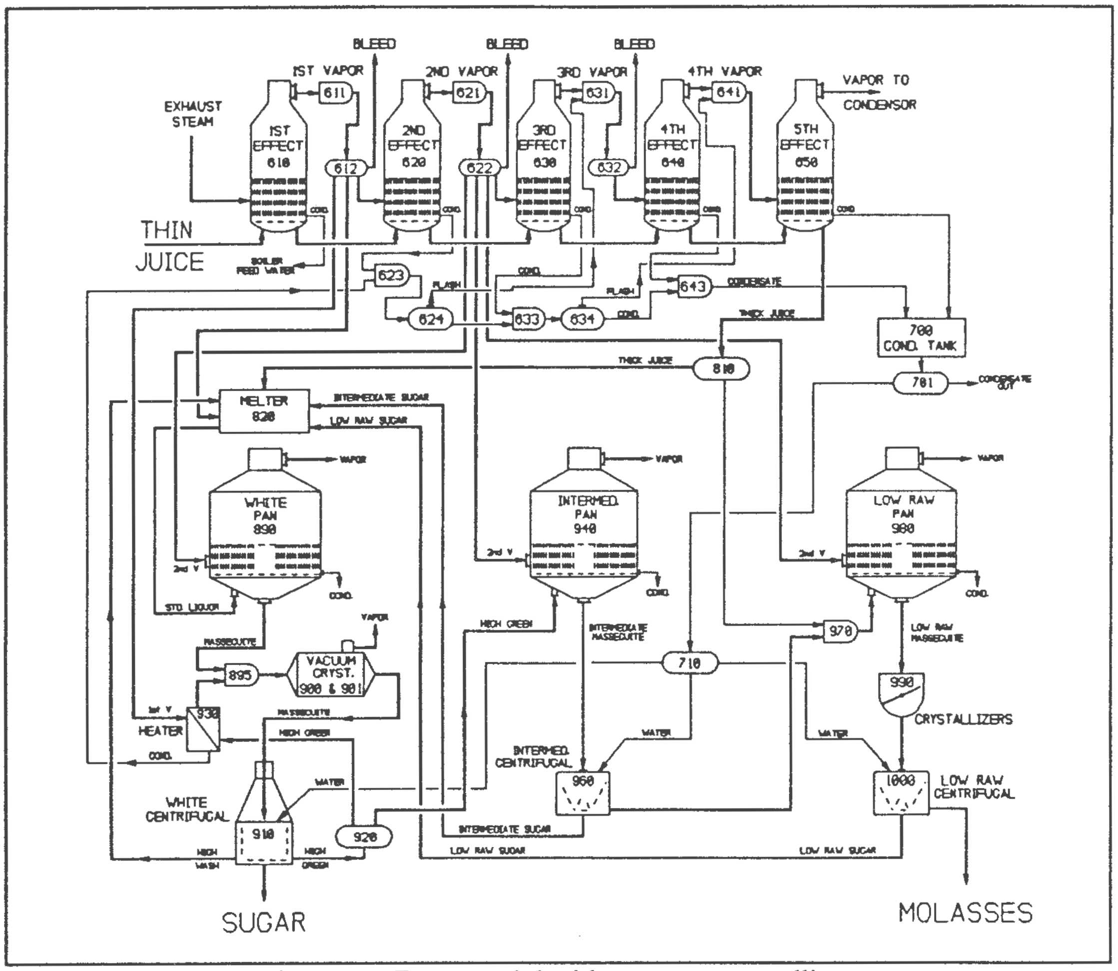

Figure 8. Base model with vacuum crystallizer.

As shown in Figure 9, the quantity of crystals increases by about 17.7% between the white pan and vacuum crystallizer massecuites; however, the increase in crystallization between the base case and vacuum crystallizer crystal quantities is really only about 2.79% which is approximately the same as the increase in sugar production from the white centrifugals. All of the pan (and melter) loadings decrease due to the vacuum crystallizer: the white pan decreases by approximately 10.6%, the intermediate pan decreases by 24.5%, and the low raw pan decreases by 16.6%. In contrast, white centrifugal loading increases by about 6%. Thus, one of the major advantages of a vacuum crystallizer is the additional capacity that can be handled by the sugar end, but it can be realized only if there is sufficient centrifugal capacity to handle the additional quantity of white massecuite.

Sugar extraction increases with a vacuum crystallizer and molasses purity decreases. Figure 9 shows an increase of 1.0 metric ton per hour of sugar with a decrease of 3.72 points in the molasses purity when compared to the base case. In the vacuum crystallizer model shown, the low raw pan purity may be too low for acceptable operation, and it may be necessary to raise this purity by using high green syrup. If this were done, the results of the simulations shown in Figures 9 and 10 would change and the additional sugar extraction of 1.0 metric tons per hour would diminish to a smaller amount.

BEET SUGAR END WITH EVAPORATION & VACUUM CRYST.

27.01.91 SUGARS V2.50 MATERIAL BALANCE RESULTS (SI UNITS) 17:04:30

%DS, PURITY & %CRYS. ARE ONLY FOR SOLUBLE PORTION OF FLOW

EXTERNAL FLOWS INTO FACTORY

FLOW	NAME	GOES TO	TONS/HR	%TDM	%SUGAR	%DS	PURITY	%CRYS.	TEMP(C)	%ISNS	%GAS
1	THIN JUICE	610	320.00	14.15	12.90	14.15	91.17	0.00	128.0	0.00	0.00

STATION FLOWS WITHIN FACTORY

STATION	NAME	GOES TO	TONS/HR	%TDM	%SUGAR	%DS	PURITY	%CRYS.	TEMP(C)	%ISNS	%GAS
•	•	•	•	•	•	•	•	•	•	•	•
•	•	•	•	•	•	•	•	•	•	•	•
•	•	•	•	•	•	•	•	•	•	•	•
820	MELTER	890	101.74	67.37	62.08	67.37	92.14	0.00	92.0	0.00	0.00
890	WHITE PAN	895	74.91	91.50	84.31	91.50	92.14	51.91	75.0	0.00	0.00
895	VACUUM CRYS.-BLENDER	900	90.16	88.72	80.47	88.72	90.70	41.04	76.7	0.00	0.00
900	VACUUM CRYST.-FLASH	901	88.83	90.05	81.68	90.05	90.70	41.66	60.0	0.00	0.00
901	VACUUM CRYST.-CRYST	910	88.83	90.05	81.68	90.05	90.70	51.52	60.0	0.00	0.00
910	WHITE CENTRIFUGALS										
	GREEN	920	53.36	75.08	61.61	75.08	82.06	0.00	58.2	0.00	0.00
	WASH	820	6.07	72.39	68.29	72.39	94.34	0.00	60.2	0.00	0.00
	SUGAR	OUT	<u>36.12</u>	98.41	98.40	98.41	99.99	94.21	51.9	0.00	0.00
920	HIGH GREEN DISTRI.										
	OUTPUT #1	930	15.25	75.08	61.61	75.08	82.06	0.00	58.2	0.00	0.00
	OUTPUT #2	940	38.11	75.08	61.61	75.08	82.06	0.00	58.2	0.00	0.00
930	HIGH GREEN HEATER	895	15.25	75.08	61.61	75.08	82.06	0.00	85.0	0.00	0.00
940	INTERMEDIATE PAN	960	30.77	93.00	76.32	93.00	82.06	40.47	81.0	0.00	0.00
960	INTERMED. CENTRIF.										
	GREEN	970	19.35	85.00	61.41	85.00	72.26	3.47	75.7	0.00	0.00
	SUGAR	820	12.51	97.26	92.70	97.26	95.32	82.34	72.9	0.00	0.00
970	LOW RAW PAN RECVR.	980	19.35	85.00	61.41	85.00	72.26	3.47	75.7	0.00	0.00
980	LOW RAW PAN	990	17.59	93.50	67.56	93.50	72.26	24.48	85.0	0.00	0.00
990	L.R. CRYSTALLIZERS	1000	17.59	93.50	67.56	93.50	72.26	38.03	54.0	0.00	0.00
1000	LOW RAW CENTRIFUGALS										
	GREEN	OUT	<u>**11.88**</u>	81.84	48.21	81.84	<u>**58.91**</u>	0.00	48.7	0.00	0.00
	SUGAR	820	7.03	95.71	87.61	95.71	91.54	75.13	47.0	0.00	0.00

Figure 9. Base model with vacuum crystallizer - material flows.

Steam consumption decreases with a vacuum crystallizer because of the reduced loading of massecuite through all of the pans and the smaller amount of recycle through the melter. Figure 10 shows the heating flows printout with the exhaust steam consumption underlined and in bold letters. Comparing this printout with the printout for the base case shows a small decrease in exhaust steam consumption of about 253 kgs/hour. Also, note that the vapor consumption for the intermediate and low raw pans is reduced (compare the condensate flows out of each pan), but the vapor to the white pan increases because the quantity of water evaporated in the white pan is higher (by almost .5 metric tons per hour) which is due to the dry substance of the standard liquor being lower than the reduction in massecuite quantity.

The increase in sugar production due to the addition of a vacuum crystallizer is 1.0 metric tons per hour. At the current U.S. price for sugar of $23.00 per (cwt) bag (or, $507 per metric ton), the net revenue gain from increased sugar production is <u>**$12,168**</u> per day.

Molasses production decreases by 1.4 metric tons per hour. Using a current U.S. price of $55 per U.S. ton (or, $61 per metric ton) for molasses, the net revenue loss from molasses sales amounts to <u>**$2,050**</u> per day.

Steam consumption decreases by 253 kgs/hour, and assuming steam costs $3.00 per 1,000 pounds (or, $6.60 per metric ton), the steam savings would amount to <u>**$40**</u> per day when compared to the base model.

BEET SUGAR END WITH EVAPORATION & VACUUM CRYST.

27.01.91 SUGARS V2.50 VAPOR & CONDENSATE CALCULATION RESULTS (SI UNITS) 17:04:32

EXTERNAL VAPOR & CONDENSATE FLOWS INTO FACTORY

FLOW	NAME	GOES TO	kPa	TEMP(C)	FLOW(kg/h)	%COND.	%TDM	%SUGAR	%VAPOR	%GASES
2	EXHAUST STEAM	610	416.5	145.0	107584.9	0.00	0.00	0.00	100.00	0.00

VAPOR & CONDENSATE FLOWS WITHIN FACTORY

STATION	NAME	GOES TO	kPa	TEMP(C)	FLOW(kg/h)	%COND.	%TDM	%SUGAR	%VAPOR	%GASES
610	1ST EFFECT									
	VAPOR	611	315.2	135.2	100656.7	0.03	0.01	0.01	99.96	0.00
	CONDENSATE	OUT	416.5	145.0	107584.9	100.00	0.00	0.00	0.00	0.00
611	1ST VAPOR RECEIVER	612	315.2	135.2	100656.7	0.03	0.01	0.01	99.96	0.00
612	1ST VAPOR DISTRI.									
	VAPOR #1	OUT	315.2	135.2	10000.0	0.03	0.01	0.01	99.96	0.00
	VAPOR #2	820	315.2	135.2	3133.3	0.03	0.01	0.01	99.96	0.00
	VAPOR #3	930	315.2	135.2	502.1	0.03	0.01	0.01	99.96	0.00
	VAPOR #4	620	315.2	135.2	87021.4	0.03	0.01	0.01	99.96	0.00
•	•	•	•	•	•	•	•	•	•	•
•	•	•	•	•	•	•	•	•	•	•
•	•	•	•	•	•	•	•	•	•	•
890	WHITE PAN									
	VAPOR	OUT	25.2	65.2	26830.2	0.00	0.01	0.00	99.99	0.00
	CONDENSATE	OUT	220.5	123.4	27208.5	99.99	0.01	0.00	0.00	0.00
900	VACUUM CRYST.-FLASH	OUT	12.9	50.8	1333.1	0.00	0.00	0.00	100.00	0.00
930	HIGH GREEN HEATER	623	315.2	135.2	502.1	99.99	0.01	0.01	0.00	0.00
940	INTERMEDIATE PAN									
	VAPOR	OUT	27.3	67.0	7345.7	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	OUT	220.5	123.4	9108.6	99.99	0.01	0.00	0.00	0.00
980	LOW RAW PAN									
	VAPOR	OUT	29.1	68.5	1760.3	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	OUT	220.5	123.4	2094.1	99.99	0.01	0.00	0.00	0.00

Figure 10. Base model with vacuum crystallizer - heating flows.

Thus, the net revenue benefit from adding a vacuum crystallizer to the base model is:

Additional sugar revenues (+1.0 MTPH)	$12,168
Less molasses decrease (-1.4 MTPH)	-2,050
Plus steam savings (-253 kgs/hour)	<u>+40</u>
Net gain per day	$10,158

The net daily gain in revenues due to the addition of a vacuum crystallizer would amount to <u>**$1,218,960**</u> for a 120 day campaign if no other changes were made to the process, and the operating conditions remained the same as in the base model with no consideration given to the electrical cost for operating the vacuum pump and stirrer in the crystallizer.

However, before this revenue increase is used to justify an investment in a vacuum crystallizer, one caveat that should be noted is that the purity of the low raw pan may need to be raised. If the low raw pan purity is raised, using high green, to the same value as for the addition of the intermediate crystallizer example (i.e., 75.53 purity), the additional sugar revenues will change to (+.34 MTPH) $4,137, the molasses decrease will change to (-.46 MTPH) $673, and the steam savings will amount to (-477 kgs/hour) +$76 for a total net daily revenue gain of <u>**$3,540**</u>, or <u>**$424,800**</u> over a 120 day campaign. The necessity for raising the low raw pan purity will depend to a large degree on the purity of the thin juice; and hence, the advantages of vacuum crystallization are more favorable with high thin juice purities so that very little, if any, high green is required to raise the purity of the low raw pan.

MOLASSES DESUGARIZATION

Significant additional sugar production can be achieved by using a chromatographic separation system for desugaring molasses. Sugar in molasses is extracted in a separator station and recycled back to the sugar end, while the raffinate, or by-product flow is concentrated and sold as a feed, or fertilizer.

Figure 11 shows the installation of a molasses desugarization system in the base model flow diagram. Molasses from the low raw centrifugal is sent to a *SUGARS* blender station model (station no. 1100) where it is diluted. The diluted molasses is then separated into extract and raffinate flow streams by a separator station model (station no. 1110) using elution water. Raffinate, which is very dilute and low in sugar, is concentrated in a multiple effect evaporator station that uses a thermocompressor (station no. 1155) to reduce the steam consumption of the multiple. For simplicity, condensate flashing isn't considered for this multiple and its operation is assumed to be similar to a normal thin juice evaporator. The concentrated raffinate can be mixed with pulp, or sold as a by-product. Extract from the separator is sent back to the main evaporator station of the factory where it is mixed with thin juice and concentrated to thick juice for crystallization. Purities of the thin juice and extract flow streams are similar so that only a small decrease occurs in the purity; however, the higher concentration of the extract raises the dry substance of the juice flow to the evaporator station and helps to reduce the evaporation load.

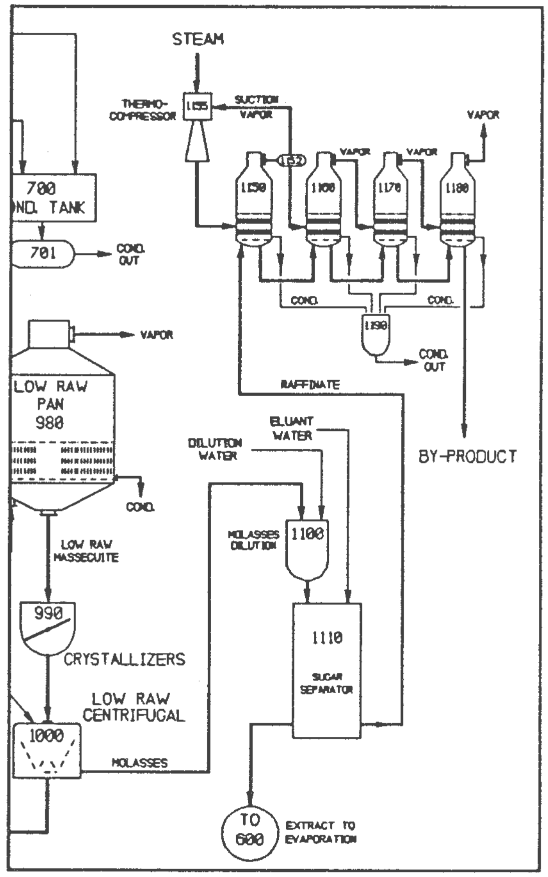

Figure 11. Molasses desugarization.

The flow diagram shown in Figure 11 is a simplified diagram that doesn't consider any pretreatment of the molasses using sodium hydroxide (NaOH), hydrochloric acid (HCl), or other chemicals[3], or softening of the thin juice before evaporation. Adding pretreating, or softening equipment and the related flow streams, would not have a significant effect on the heat and material balance done by *SUGARS*; however, using recirculation flow streams from the separator station for diluting the molasses could reduce the quantity of molasses dilution water[4]. Also, not considered in this model is the change in sucrose solubility coefficients due to the recycle of extract from the separator, and the subsequent effect the change in solubility coefficients would have on crystallization. Most likely this would have a positive effect because of a reduction in melassigenic compounds due to the separator; i.e., the *SUGARS* simulation should be conservative regarding sugar extraction when a separator is installed. In actual practice, non-separable non-sugars (e.g., amines) would probably prevent processing all of the molasses in the separator and recycling the extract back to the thin juice. Instead, a portion of the molasses could be sent to storage and would serve as a blow down to prevent a build up of these non-sugars, or instead the extract could be concentrated and stored the same as thick juice to be processed at a later time. The flow diagram could be made more complex by adding other station models to consider all of the process requirements; however, without further solubility data and for the purpose of evaluating the potential revenue gain, installation of only the additional equipment shown in Figure 11 should suffice.

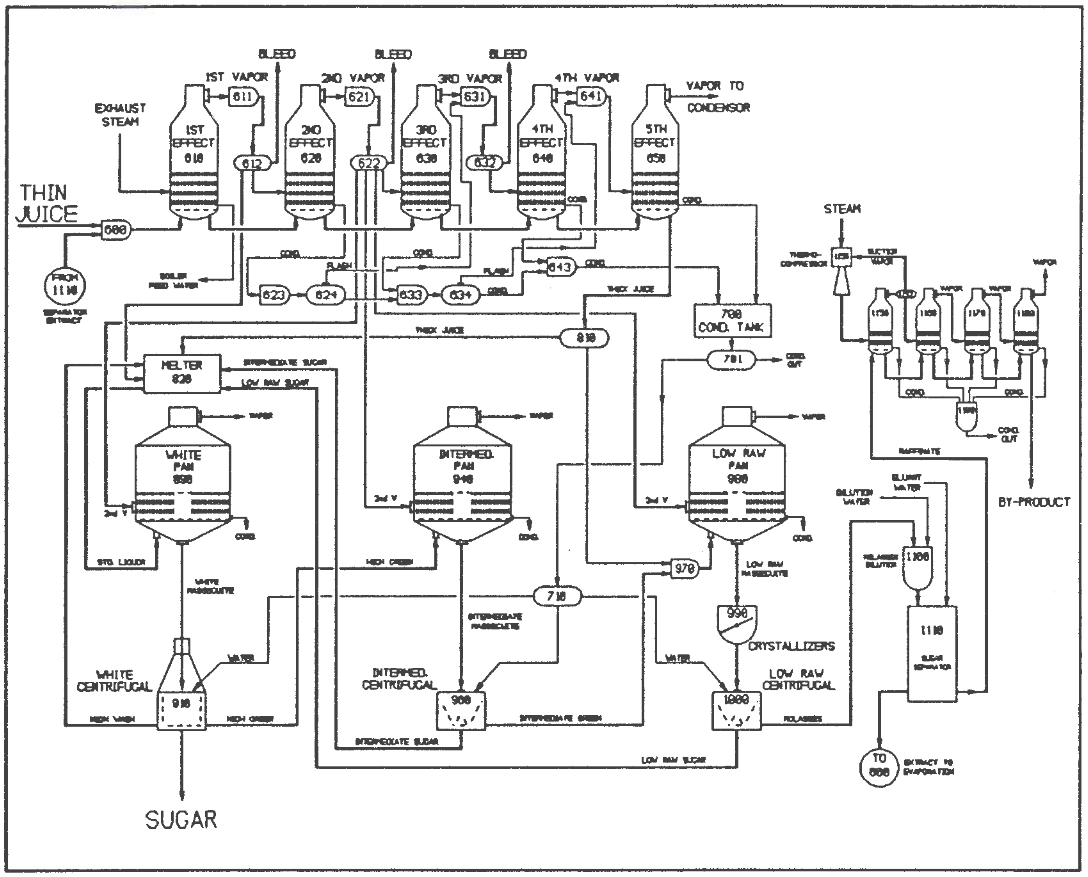

Figure 12. Base model with molasses desugarization.

The complete flow diagram for the base model with molasses desugarization is shown in Figure 12. No changes were made to the base model other than the addition of the molasses desugarization equipment and a receiver station (no. 600) to receive extract flow from the separator and combine it with thin juice to feed the evaporator station. Performance data for each station in the base model is also unchanged.

Nominal values of 90% sucrose recovery and 82.5% non-sugar elimination for the molasses to extract separation were used for the performance of the sucrose separator station model (no. 1100). Also, the molasses is diluted to 61.70% dry substance before it is sent to the separator and 75% of the water in the diluted molasses is passed to the extract. Elution water is proportioned into the separator at approximately 7.5 times the weight quantity of sugar in the molasses and 18.6% of the eluant is passed to the extract flow stream. All remaining quantities and components are passed to the raffinate flow stream leaving the separator station.

BEET SUGAR END W/ EVAPS. & MOLASSES DESUGARIZATION

27.01.91 SUGARS V2.50 MATERIAL BALANCE RESULTS (SI UNITS) 17:06:29

%DS, PURITY & %CRYS. ARE ONLY FOR SOLUBLE PORTION OF FLOW

EXTERNAL FLOWS INTO FACTORY

FLOW	NAME	GOES TO	TONS/HR	%TDM	%SUGAR	%DS	PURITY	%CRYS.	TEMP(C)	%ISNS	%GAS
1	THIN JUICE	600	320.00	14.15	12.90	14.15	91.17	0.00	128.0	0.00	0.00
3	DILUTION WATER	1100	4.89	0.00	0.00	0.00	0.00	0.00	90.0	0.00	0.00

STATION FLOWS WITHIN FACTORY

STATION	NAME	GOES TO	TONS/HR	%TDM	%SUGAR	%DS	PURITY	%CRYS.	TEMP(C)	%ISNS	%GAS
600	T.J./EXTRACT RECVR.	610	345.13	15.45	14.05	15.45	90.91	0.00	124.3	0.00	0.00
610	1ST EFFECT	620	237.84	22.42	20.38	22.42	90.91	0.00	135.0	0.00	0.00
620	2ND EFFECT	630	141.82	37.59	34.17	37.59	90.91	0.00	122.8	0.00	0.00
630	3RD EFFECT	640	109.72	48.59	44.17	48.59	90.91	0.00	110.2	0.00	0.00
640	4TH EFFECT	650	101.32	52.61	47.83	52.61	90.91	0.00	98.5	0.00	0.00
650	5TH EFFECT	810	85.98	62.00	56.36	62.00	90.91	0.00	69.7	0.00	0.00
810	THICK JUICE DISTRI.										
	OUTPUT #1	820	85.98	62.00	56.36	62.00	90.91	0.00	69.7	0.00	0.00
	OUTPUT #2	970	0.00	0.00	0.00	0.00	0.00	0.00	0.0	0.00	0.00
820	MELTER	890	129.75	69.55	64.58	69.55	92.86	0.00	92.0	0.00	0.00
890	WHITE PAN	910	98.62	91.50	84.97	91.50	92.86	52.84	75.0	0.00	0.00
910	WHITE CENTRIFUGALS										
	GREEN	940	58.18	76.99	66.30	76.99	86.11	0.00	69.4	0.00	0.00
	WASH	820	6.78	73.29	70.11	73.29	95.67	0.00	66.7	0.00	0.00
	SUGAR	OUT	**41.11**	98.45	98.44	98.45	99.99	93.76	63.9	0.00	0.00
940	INTERMEDIATE PAN	960	48.17	93.00	80.08	93.00	86.11	46.92	81.0	0.00	0.00
960	INTERMED. CENTRIF.										
	GREEN	970	28.05	83.80	64.08	83.80	76.47	4.86	75.4	0.00	0.00
	SUGAR	820	21.83	97.54	94.38	97.54	96.76	85.58	72.9	0.00	0.00
970	LOW RAW PAN RECVR.	980	28.05	83.80	64.08	83.80	76.47	4.86	75.4	0.00	0.00
980	LOW RAW PAN	990	25.14	93.50	71.50	93.50	76.47	31.45	85.0	0.00	0.00
990	L.R. CRYSTALLIZERS	1000	25.14	93.50	71.50	93.50	76.47	44.05	54.0	0.00	0.00
1000	LOW RAW CENTRIFUGALS										
	GREEN	1100	15.92	80.63	50.22	80.63	62.28	0.00	48.6	0.00	0.00
	SUGAR	820	11.10	96.12	89.90	96.12	93.54	79.10	46.8	0.00	0.00
1100	MOLASSES DILUTION	1110	20.81	61.70	38.43	61.70	62.28	0.00	63.2	0.00	0.00
1110	SUGAR SEPARATOR										
	EXTRACT	600	25.13	32.01	28.64	32.01	89.46	0.00	72.0	0.00	0.00
	RESIDUE	1150	55.40	8.65	1.44	8.65	16.68	0.00	78.7	0.00	0.00
1150	B.P. EFFECT #1	1160	38.43	12.48	2.08	12.48	16.68	0.00	128.0	0.00	0.00
1160	B.P. EFFECT #2	1170	28.52	16.81	2.80	16.81	16.68	0.00	110.1	0.00	0.00
1170	B.P. EFFECT #3	1180	18.16	26.40	4.40	26.40	16.68	0.00	95.1	0.00	0.00
1180	B.P. EFFECT #4	OUT	7.38	65.00	10.84	65.00	16.68	0.00	67.8	0.00	0.00

Figure 13. Base model with molasses desugarization - material flows.

Figure 13 shows the material flows printout with the white sugar production underlined and in bold. Comparing this quantity of sugar production with the base model reveals a whopping 5.99 metric tons per hour of additional sugar. As expected, loadings on all of the equipment in the sugar end increase appreciably while the purities are somewhat lower but similar to those in the base model. The quantity of thick juice and standard liquor increases by 17.8%, and loadings on all of the other equipment in the sugar end increase by 17% to 20%; hence, one of the significant costs of adding a molasses desugarization system might be the investment that is needed to increase the capacity of the sugar end if the separator is operated in unison with the normal beet campaign.

As mentioned earlier, blending extract flow with thin juice before it goes to the evaporator station will change the solubility characteristics of the sugar end crystallization. Because a larger amount of melassigenic non-sugars are removed as compared to the sugar recovered in the separator, it is expected that the solubility will not be worse than in the base model; and instead, most likely it will be better. Hence, the sugar extraction efficiency may actually be better than shown in Figure 13.

BEET SUGAR END W/ EVAPS. & MOLASSES DESUGARIZATION

27.01.91 SUGARS V2.50 VAPOR & CONDENSATE CALCULATION RESULTS (SI UNITS) 17:06:32

EXTERNAL VAPOR & CONDENSATE FLOWS INTO FACTORY

FLOW	NAME	GOES TO	kPa	TEMP(C)	FLOW(kg/h)	%COND.	%TDM	%SUGAR	%VAPOR	%GASES
2	EXHAUST STEAM	610	416.5	145.0	**<u>116718.3</u>**	0.00	0.00	0.00	100.00	0.00
4	ELUTION WATER	1110	101.3	80.0	59722.0	100.00	0.00	0.00	0.00	0.00
5	150# STEAM	1155	1125.0	185.0	**<u>14534.4</u>**	0.00	0.00	0.00	100.00	0.00

VAPOR & CONDENSATE FLOWS WITHIN FACTORY

STATION	NAME	GOES TO	kPa	TEMP(C)	FLOW(kg/h)	%COND.	%TDM	%SUGAR	%VAPOR	%GASES
•	•	•	•	•	•	•	•	•	•	•
•	•	•	•	•	•	•	•	•	•	•
•	•	•	•	•	•	•	•	•	•	•
650	5TH EFFECT									
	VAPOR	OUT	27.3	67.0	15339.0	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	700	89.1	96.5	11739.6	100.00	0.00	0.00	0.00	0.00
700	COND. TANK	701	89.0	97.1	133650.8	99.99	0.01	0.01	0.00	0.00
701	CONDENSATE DISTRI.									
	CONDENSATE #1	710	89.0	97.1	11046.2	99.99	0.01	0.01	0.00	0.00
	CONDENSATE #2	OUT	89.0	97.1	122604.6	99.99	0.01	0.01	0.00	0.00
710	CENT. H2O DISTRI.									
	CONDENSATE #1	910	89.0	97.1	7454.5	99.99	0.01	0.01	0.00	0.00
	CONDENSATE #2	960	89.0	97.1	1711.3	99.99	0.01	0.01	0.00	0.00
	CONDENSATE #3	1000	89.0	97.1	1880.4	99.99	0.01	0.01	0.00	0.00
890	WHITE PAN									
	VAPOR	OUT	25.4	65.4	31133.7	0.00	0.01	0.00	99.99	0.00
	CONDENSATE	OUT	208.8	121.6	31096.5	99.99	0.01	0.00	0.00	0.00
940	INTERMEDIATE PAN									
	VAPOR	OUT	28.4	67.9	10016.1	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	OUT	208.8	121.6	11594.0	99.99	0.01	0.00	0.00	0.00
980	LOW RAW PAN									
	VAPOR	OUT	29.9	69.0	2911.9	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	OUT	208.8	121.6	3411.7	99.99	0.01	0.00	0.00	0.00
1150	B.P. EFFECT #1									
	VAPOR	1152	254.4	128.0	16972.1	0.00	0.00	0.00	100.00	0.00
	CONDENSATE	1190	416.5	145.0	22528.4	100.00	0.00	0.00	0.00	0.00
1152	VAPOR DISTRIBUTOR									
	VAPOR #1	1155	254.4	128.0	7993.9	0.00	0.00	0.00	100.00	0.00
	VAPOR #2	1160	254.4	128.0	8978.1	0.00	0.00	0.00	100.00	0.00
1155	THERMO-COMPRESSOR	1150	416.5	153.0	22528.4	0.00	0.00	0.00	100.00	0.00
•	•	•	•	•	•	•	•	•	•	•
•	•	•	•	•	•	•	•	•	•	•
•	•	•	•	•	•	•	•	•	•	•

Figure 14. Base model with molasses desugarization - heating flows.

Steam consumption increases with the installation of a molasses desugarization system. The quantity of exhaust steam to the main multiple effect, and the quantity of 150 pound steam to the by-product evaporator station are underlined and in bold in Figure 14. Suction vapor flow to the thermocompressor on the by-product evaporator is 7,993.9 kgs/hour (set Figure 14, station no. 1152, Vapor #1) based on an entrainment ratio of .55 for the thermocompressor for a pressure ratio of approximately 3.7 between the motive steam and suction steam pressures.

The increase in sugar production due to the addition of molasses desugarization is 5.99 metric tons per hour. At the current U.S. price of $23.00 per (cwt) bag (or, $507 per metric ton), the net revenue gain from this increase is **$72,886** per day.

Molasses production decreases to zero, or the loss of molasses production amounts to 13.28 metric tons per hour. Using a current U.S. price of $55 per U.S. ton (or, $61 per metric ton) for molasses, the net revenue loss from molasses sales is **$19,442** per day. However, the value of by-product is estimated to be approximately 80% of molasses on a dry weight basis. Thus, the by-product value is approximately $35.50 per U.S. ton (or, about $39 per metric ton) when considering molasses at 80.50%DS and by-product at 65%DS. This results in a by-product value of **$6,910** per day to partially off-set the loss in molasses sales.

Steam consumption increases by 8,881 kgs/hour for exhaust and 14,534 kgs/hour for 150 pound steam; hence, the total steam increase is 23,415 kgs/hour. Assuming both exhaust and 150 pound steam cost $3.00 per 1,000 pounds (or, $6.60 per metric ton), the additional steam cost is **$3,710** per day.

The chromatographic separator incurs additional expenses for replacement of the resins used in the separator. Modern resins have a long life; however, replacement costs are high. For this example, dilution and elution water costs, other chemical (i.e., NaOH, HCl, etc) costs and separator resin replacement costs are outside the scope of this paper and will need to be evaluated on a case-by-case basis. Thus, the net revenue benefit from adding molasses desugarization to the base model is:

Additional sugar revenues (+5.99 MTPH)	$72,886
Less molasses decrease (-13.28 MTPH)	-19,442
Plus by-product revenues (+7.38 MTPH)	+6,910
Less additional steam costs (23.4 MTPH)	-3,709
Net gain per day	$56,645

The above net daily gain in revenues from the installation of a molasses desugarization system would amount to approximately **$6,797,400** for a 120 day campaign when the costs of other chemicals, dilution and elution water, and resin replacement are excluded.

SUMMARY

The *SUGARS* computer program is a flexible program for modeling sugar factory processes and equipment. Its uses and features are discussed and examples are given of an approximately 6,000 metric ton per day three boiling beet sugar end with five effect evaporator. The examples shown are additions to the sugar end of: (1) an intermediate crystallizer; (2) a vacuum crystallizer on white massecuite; and (3) a molasses desugarization system.

The results from each example are discussed and it is shown that: (1) addition of an intermediate crystallizer can result in a daily revenue gain of more than $3,000; (2) addition of a vacuum crystallizer on white massecuite can result in increased daily revenues of more than $10,000 without any high green being used to raise the low raw pan purity, but only

about $3,500 per day if high green is used to raise the low raw pan purity; and, (3) Addition of molasses desugarization can result in daily revenues increasing by more than $56,000 when excluding the costs of additional water, chemicals for molasses pretreatment, and separator column resin replacement. Other effects to the process are discussed for each of the examples using the results from computer modeling by *SUGARS* to evaluate their impact on the sugar end.

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