Investment Decisions

Using

The *SUGARS* Computer Program

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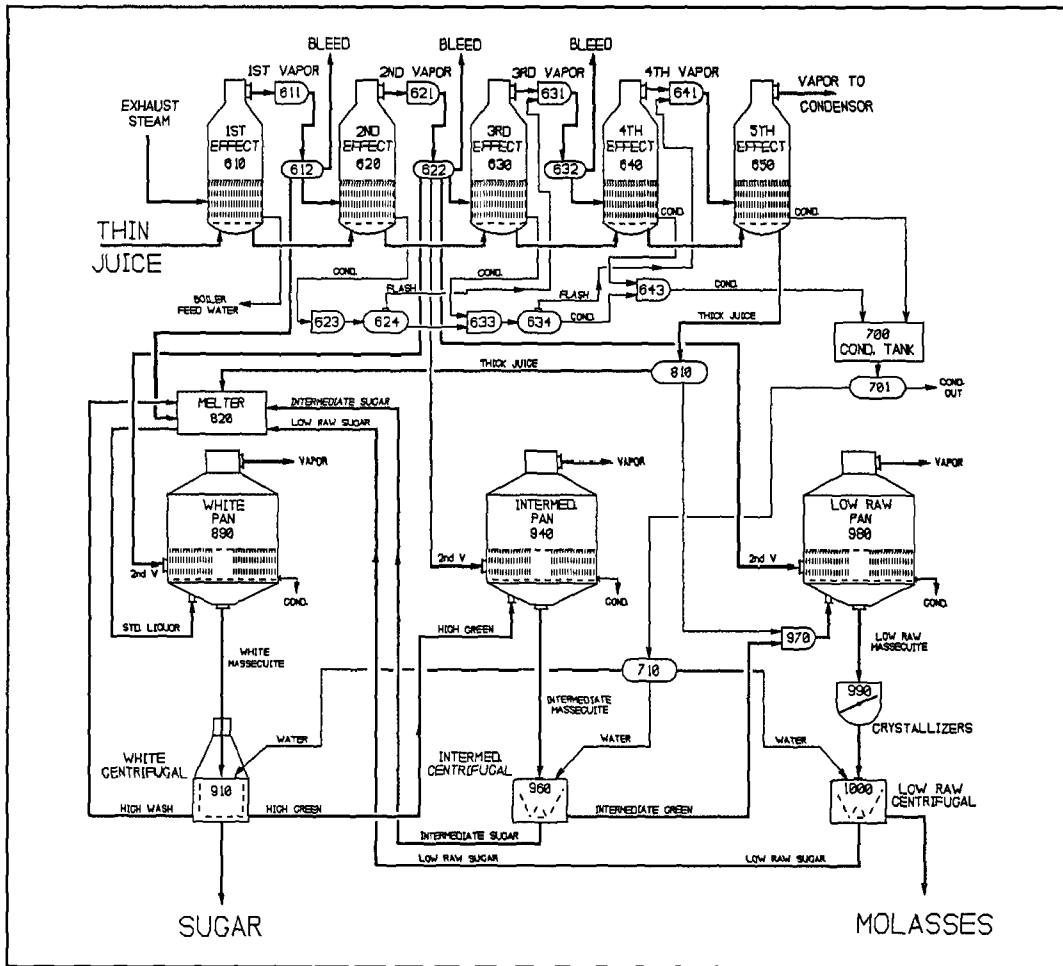
Figure 1. Base model evaporation and sugar end flow diagram.

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

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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.

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                          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*,

and the printout for the base model is shown in Figure 3. As in the material flows printout, the external heating flow into the flow diagram is given first (i.e., EXHAUST STEAM) and the quantity of this flow at **107,837.5** kg/h is underlined and in bold for subsequent reference as changes are made to the base model.

In both printouts, the flow stream characteristics (quantity, %DS, purity, temperature, etc) out of each station are calculated by *SUGARS* based on the input flow to the station and the station characteristics that are specified in the input data.

Most of the headings on the printouts are self-explanatory, except possibly:

%TDM	- Percent Total Dry Matter for the flow stream
%DS	- Percent Dry Substance (solubles)
%ISNS	- Percent Insoluble Solid Non-Sugars
%GAS	- Percent vapor and gases (CO_2 and NH_3) in material flow streams
%GASES	- Percent gases other than vapor in the heating flow streams

Entrainment losses were specified for the evaporator bodies and pans in the base model which is the reason for the trace amounts of liquid and sugar in the vapors from these stations (see heating flows printout, Figure 3 for the vapor and condensate flows). The entrained liquid is assumed to be in equilibrium with the vapor at the same percent dry substance (%DS) and Purity as the material (syrup, juice, etc) leaving the station. Hence, if the vapor is injected into another flow stream, or if it is condensed in a subsequent station, the dry substance and sugar content in the condensate will be considered in the resultant flow stream.

INTERMEDIATE CRYSTALLIZER

The first modification to the base model is the addition of an intermediate crystallizer to cool the massecuite and give it additional time for crystal growth. Addition of this equipment is shown in Figure 4, where crystallizer station number 950 is placed between the Intermediate Pan (station no. 940) and the Intermediate Centrifugal (station no. 960).

No other changes are made to the flow diagram and all station input parameters remain as they were for the base model. Only an intermediate crystallizer is added to cool the massecuite by 10°C while still keeping the supersaturation of the massecuite the same as it was following its discharge from the intermediate pan; i.e., a 1.10 supersaturation.

Additional crystallization and cooling of the massecuite due to the new intermediate crystallizer will change the balance in the sugar end; and hence, some variations will occur in bleed vapors that are drawn from the evaporator station. As the bleed vapor quantities change, the vapor temperature and pressure for each effect will rise, or fall as the evaporation changes in each effect. *SUGARS* can account for these variations by using heat transfer coefficients and surface areas for each body in the multiple. Optionally, *SUGARS* can hold the vapor temperature, or juice flow out temperature, for each body; however, for the examples in this paper, heat transfer coefficients and surface areas are used for each body in the five effect multiple so that all vapor temperatures can float as the vapor bleeds change.

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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.
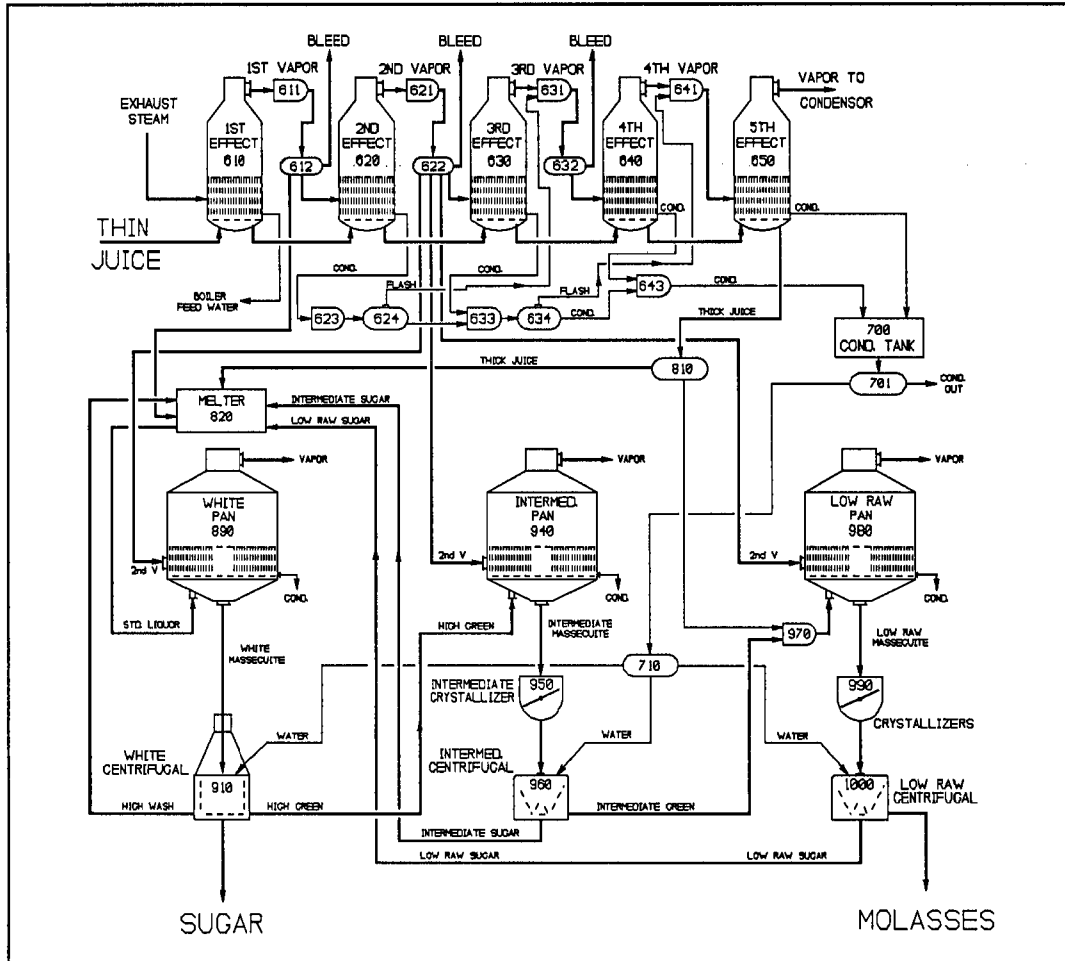

Figure 4. Evaporation and sugar end with intermediate crystallizer.

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

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effect is reduced; and hence, the difference in temperature between syrup leaving, and heat-

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                    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    12.62  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
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Figure 5. Base model with intermediate crystallizer - material flows.

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                    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 **$3,894** 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 **$673** 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 **$55** 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 **$379,920** 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

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(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.
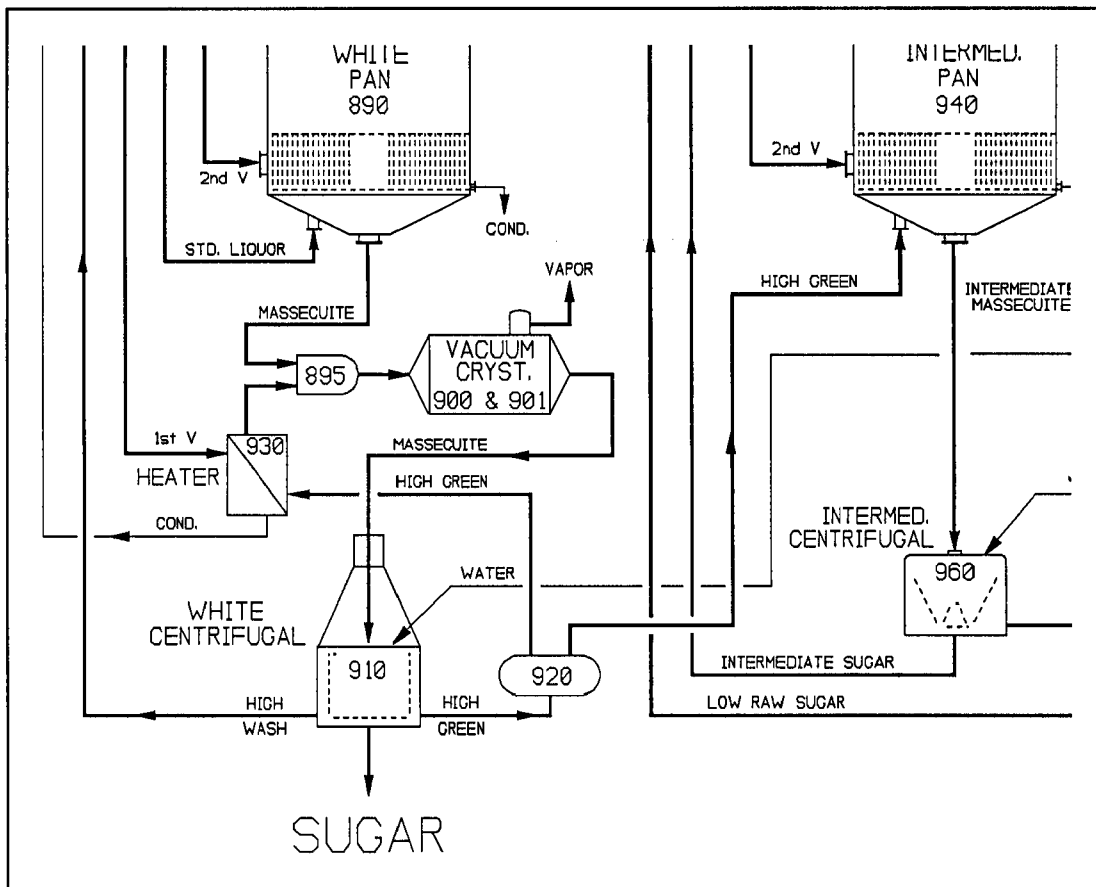

Figure 7. Vacuum crystallization installation.

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
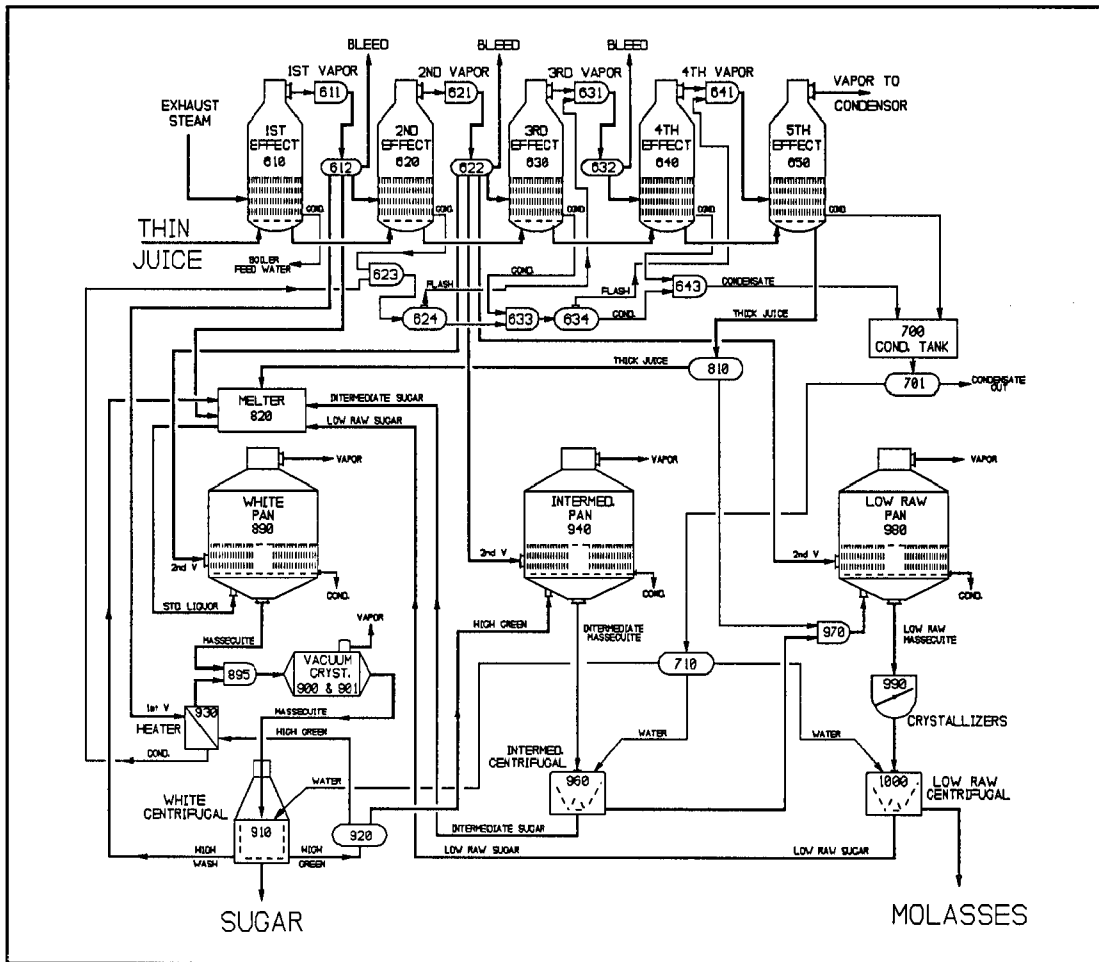
material balance printout given in Figure 9.

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    36.12  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    11.88  81.84  48.21  81.84  58.91  0.00    48.7  0.00  0.00
         SUGAR           820     7.03  95.71  87.61  95.71  91.54 76.13    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 **$12,168** 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 **$2,050** 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 **$40** 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)	+40
Net gain per day	$10,158

The net daily gain in revenues due to the addition of a vacuum crystallizer would amount to **$1,218,960** 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 **$3,540**, or **$424,800** over a 120 day campaign. The necessity for raising the

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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.
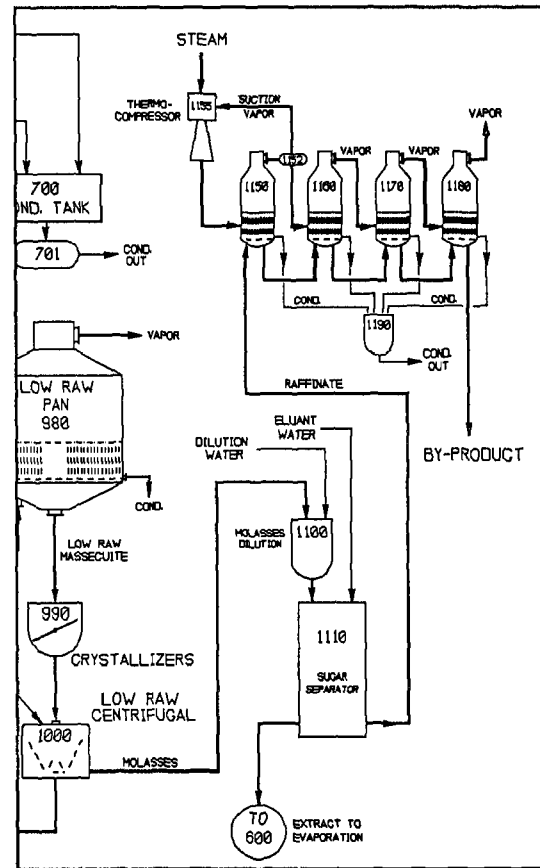

Figure 11. Molasses desugarization.

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

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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.
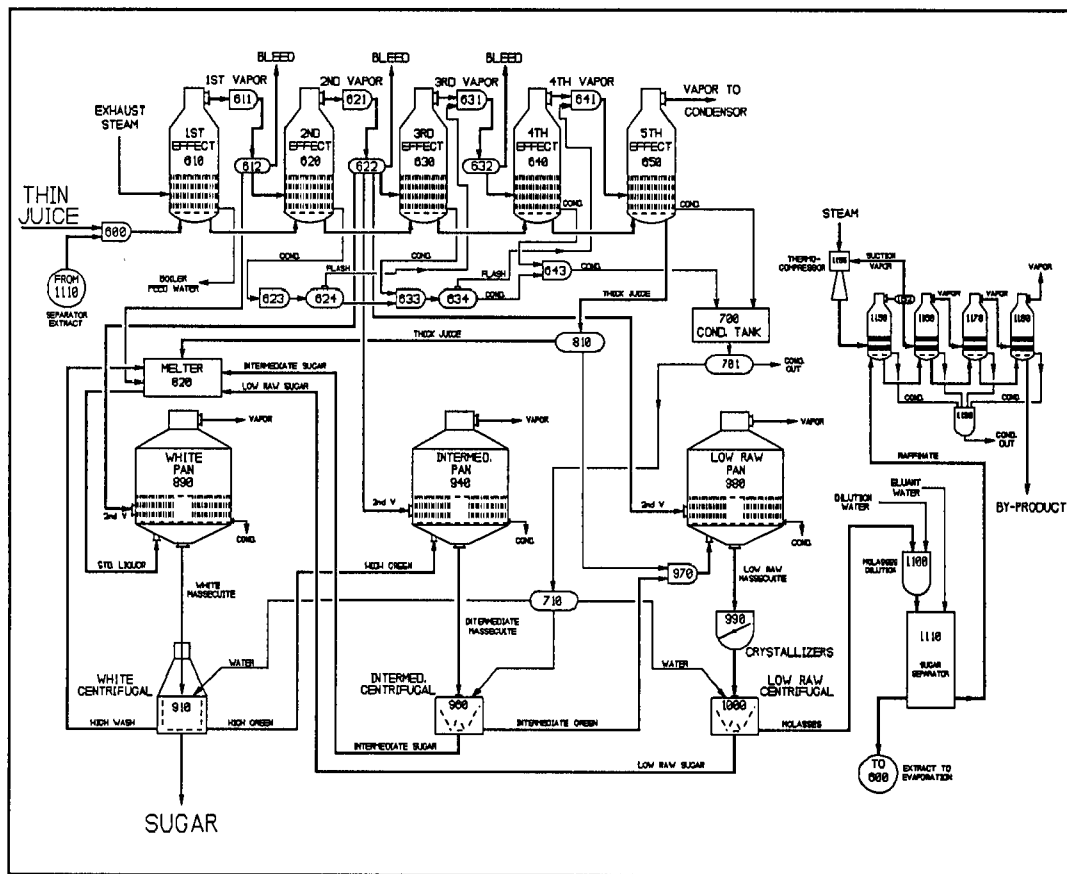
Figure 12. Base model with molasses desugarization.

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

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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.

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                    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    116718.3    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     14534.4    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    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      7464.5   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
      .            .           .       .       .         .          .       .       .      .       .
      .            .           .       .       .         .          .       .       .      .       .
      .            .           .       .       .         .          .       .       .      .       .
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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.

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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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