

POPE, JAMES D., and HUDE, JEFFREY M., LURGI PSI, 1790 Kirby Parkway, Suite 300, Memphis, TN 38138. **Industrial performance of the mechanical vapor recompression and multiple effect evaporator system: successful operation and significant reduction in steam usage.**

Introduction

During January of 2000, a new evaporator system was started up to meet the needs of a new molasses desugarization (MDS) process installed and started at the same time. The evaporator system described in this paper is part of the MDS facility owned by Crystech, LLC and operated by the American Crystal Sugar Company factory in Hillsboro, North Dakota. Preparation for this startup included time for conceptual development, feasibility studies and evaluations, detailed design, modeling runs, construction, operator and maintenance training and subsequent successful startup. This paper shall document the design and present the results of the operation of the evaporation system selected.

Design Criteria

The MDS process takes soft molasses from the factory sugar end and converts it into at least two streams, product and byproduct. The primary sugar carrier is the product stream, often termed extract. The main non-sugar stream is commonly referred to as raffinate. Intermediate streams must also be evaporated to meet the process needs. The following criteria had to be considered in the design of the molasses desugarization evaporator system:

- Multiple streams to concentrate
- Low initial feed concentrations to the evaporator system
- Recycle of clean process condensate for process usage
- Turn-down capability
- Flexibility to change the number of streams feeding the evaporator
- Year-round operation
- Handling non-sugar stream with precipitation of salts during and after concentration
- Stainless steel material of construction
- Limited excess steam available for new evaporators
- Available electricity

Due to limits on available steam supply at the selected site for the molasses desugarization facility, a single or multiple-effect steam heated evaporator system alone was not an acceptable option. Boiler capacity to meet the process demands could not be met without the addition of new equipment and costly infrastructure upgrades. The use of energy saving techniques, such as mechanical vapor recompression evaporation, was evaluated and ultimately selected to meet the criteria of that situation.

A mechanical vapor recompression evaporator system is similar to a conventional steam heated, single-effect evaporator, except that the vapor released from the boiling solution is compressed in a mechanical compressor. Compression raises the pressure and saturation temperature of the vapor so that it may be returned to the evaporator steam chest to be used as heating steam. The latent heat of the vapor is used to evaporate more water instead of being rejected to cooling

water in a condenser. The compressor provides energy to the vapor that increases in pressure and temperature, thereby recycling the evaporated water into usable steam to meet the evaporative load of the incoming fluid. This reduces the steam needed to meet the evaporative load of the overall system. The energy or driving force for pressure increase is provided through shaft horsepower. The most widely used drive applied to MVR systems is the constant speed motor. For high evaporative loads, like in an MDS facility, these motors may be large sizes requiring high voltage electric supply.

A system of mechanical vapor recompression units for initial concentration where acceptable to remove large quantities of water followed by multiple-effect, steam heated units to finish evaporate the process streams was the solution for the processing and project requirements. Another key factor was that the site had available the necessary electrical capacity to meet the requirements of the MVR system.

MVR Benefits

- Steam and overall energy consumption was significantly reduced compared to conventional steam heated evaporator systems with the same capacity.
- Cooling water requirements were reduced
- The main vacuum pump for the evaporator system was reduced in horsepower, capacity and size compared to one required without use of a MVR
- A vacuum pump for the raffinate MVR was not required due to its operation at pressure above atmospheric pressure
- Higher condensate temperatures provide additional steam savings by providing hot water (95-100 °C) source for use in the process.

MVR Disadvantages

- A higher capital cost than a conventional steam heated system
- A MVR could not be utilized on all streams to be evaporated as part of this process. The extract stream from the MDS process could not be evaporated in a MVR because operation at pressures near one atmosphere would result in boiling temperatures that would be too hot for the heat sensitive high purity and high concentration sugar solutions
- High electrical consumption was required with two 1250 HP motors on the raffinate MVR compressors, and these required a high voltage power supply
- Spares such as turbine blades and bearings were required for inventory.

System Selection (Design Criteria and Equipment Selections for Raffinate Evaporation)

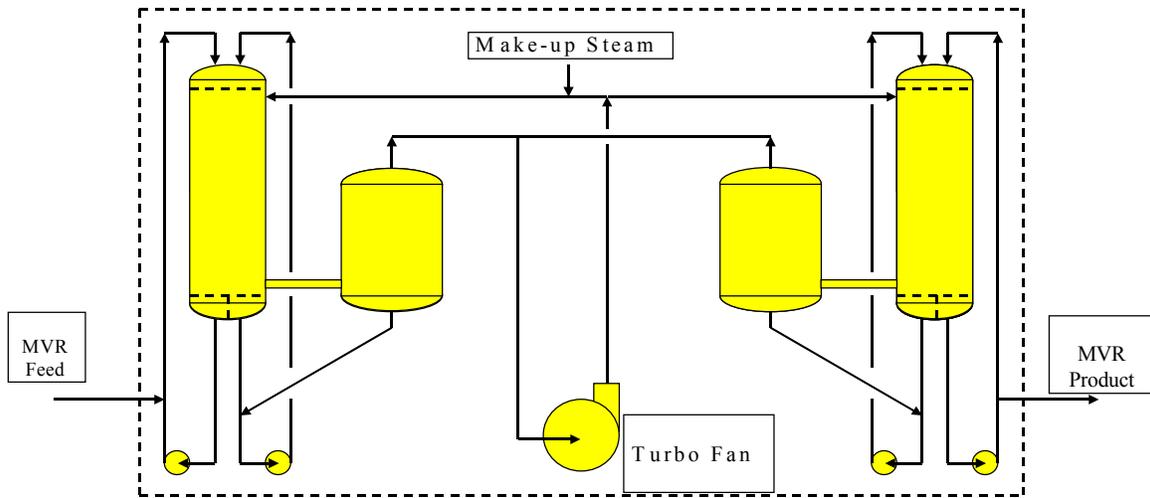
Raffinate is the largest water carrier of all MDS streams, but has the lowest total D.S. compared to the other feed streams. It has the highest total feed rate and a concentration of approximately 4-6%DS. The final concentration required for the raffinate product is approximately 65% to 70%DS. This paper will focus on the system design and operation of the raffinate stream.

Long-tube, vertical type, recirculated evaporators bodies are used with both the MVR and multiple effect evaporators due to their efficiency with high viscosity, high density, and high boiling point elevation liquids. Tube lengths up to 40 feet allow conservation of floor space.

The MVR evaporators are operated under a falling film operation. With these evaporators, the combination of a thin film undergoing boiling and viscous drag at the vapor liquid interface produces high heat transfer coefficients. Boiling is permitted in the tubes and the vapors are drawn off and separated from any entrained liquids in the appropriately named vapor separator attached to the evaporator body by ductwork. By use of the recirculated design on all of the bodies, the ability to turn down the capacity is greatly improved. Reduction in rate will result in reduction of vapors available to be compressed to drive evaporation.

As noted earlier, MVR evaporator limitations restrict the concentration of the raffinate stream to approximately 28% DS. Due to the load involved, more than one system would be required to meet the plant needs. The raffinate was preheated by the process condensate recovered from the MVR steam chests. Each of the raffinate MVR evaporator systems contains multiple steam chests divided up into multiple sections. The raffinate passes through multiple falling film sections prior to leaving the MVR evaporator. The raffinate streams are pumped around the system via recycle pumps. The vapors given off are flashed in the vapor separators that provide a means for any liquid to be returned to the evaporator chest and the vapors to be collected. The evaporated vapors from the raffinate stream are then compressed in the MVR compressor and supplied to the steam chests to give the steam required to heat the raffinate. The process condensate is collected and used for preheat prior to its use in the MDS process. By concentrating the raffinate stream in the MVR evaporator from 4-6% to 28% DS, 90% of the required evaporation will be achieved in the MVR. A typical, general arrangement of an MVR evaporator system with multiple steam chests is as follows.

Typical Multiple Chest MVR Evaporator



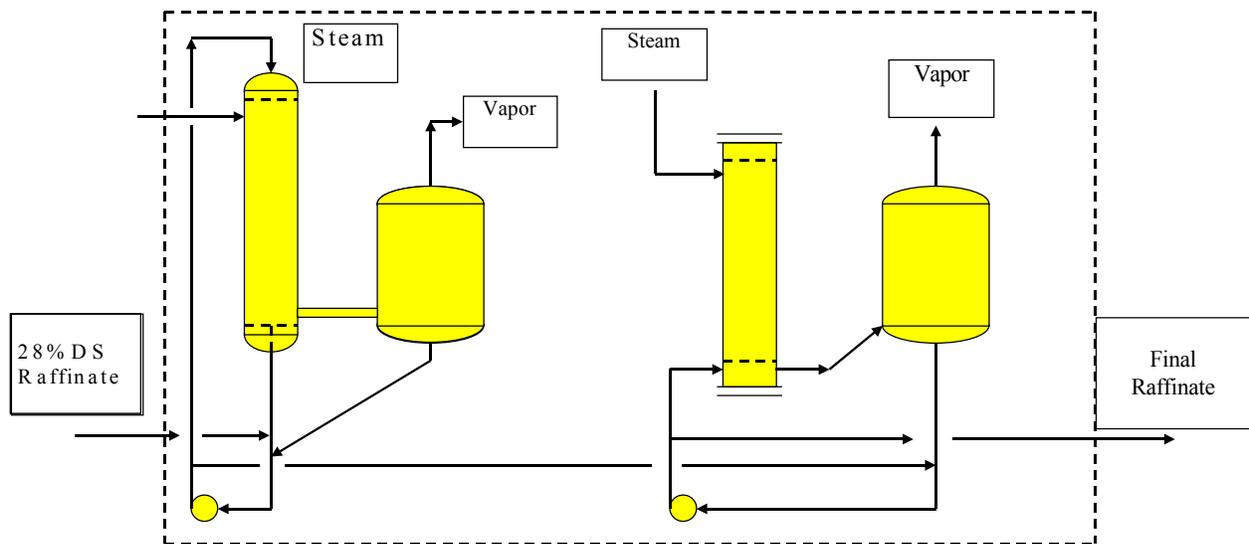
Forced circulation evaporation is used for final concentration of the raffinate stream. Precipitation of salts begins at approximately 42%DS for raffinate. Forced circulation evaporation is achieved in the finisher evaporator by using a high circulation rate through the steam chest tubes. This causes the tubes to be flooded with liquid, suppressing evaporation until the larger surface area of the vapor separation permits flashing of the vapors from the liquid. This method helps prevent excessive fouling or precipitate buildup in the exchanger tubes. The

raffinate finisher is divided into two stages. The first will utilize the falling film evaporation, while the second stage will utilize forced circulation.

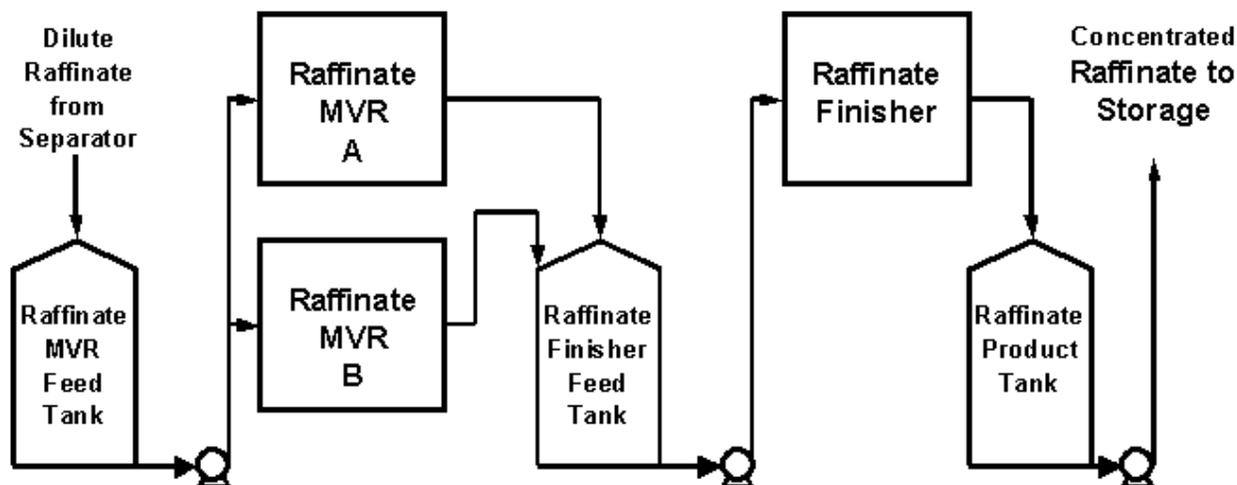
The concentrated raffinate (~28% DS) is collected in a holding tank and then pumped into the first effect of the new plants multiple-effect evaporator train. The raffinate stream is preheated with 35# steam and then sent to the first stage of the first effect. The first effect is heated with steam, and the evaporated raffinate vapors are used in subsequent effects as the driving force for evaporating other streams. The raffinate then passes into the second stage of the first effect evaporator, where it is again heated with steam and its vapors also are used in evaporation downstream in the train. The final concentrated raffinate stream is drawn off and pumped to storage. The steam condensate is used for heat recovery and then utilized in the MDS process or returned to the boiler.

Steam heated, multiple-effect evaporators are used for final concentration or as a “finisher” to bring the raffinate up to the final product specification for %DS. The raffinate section of evaporation is integrated with other streams being evaporated, with the vapors from the raffinate used as the driving force for the subsequent effects. These evaporators provide excellent steam savings from the use of vapors in series and results in an economical design, especially with the higher boiling point elevation present with higher concentrations. In addition, the pressure drop across the multiple-effect train results in evaporators operating at vacuum pressures and lower temperatures, which is ideal for heat sensitive streams like extract. Below is a diagram indicating the raffinate portion of the finisher system.

Raffinate Evaporator



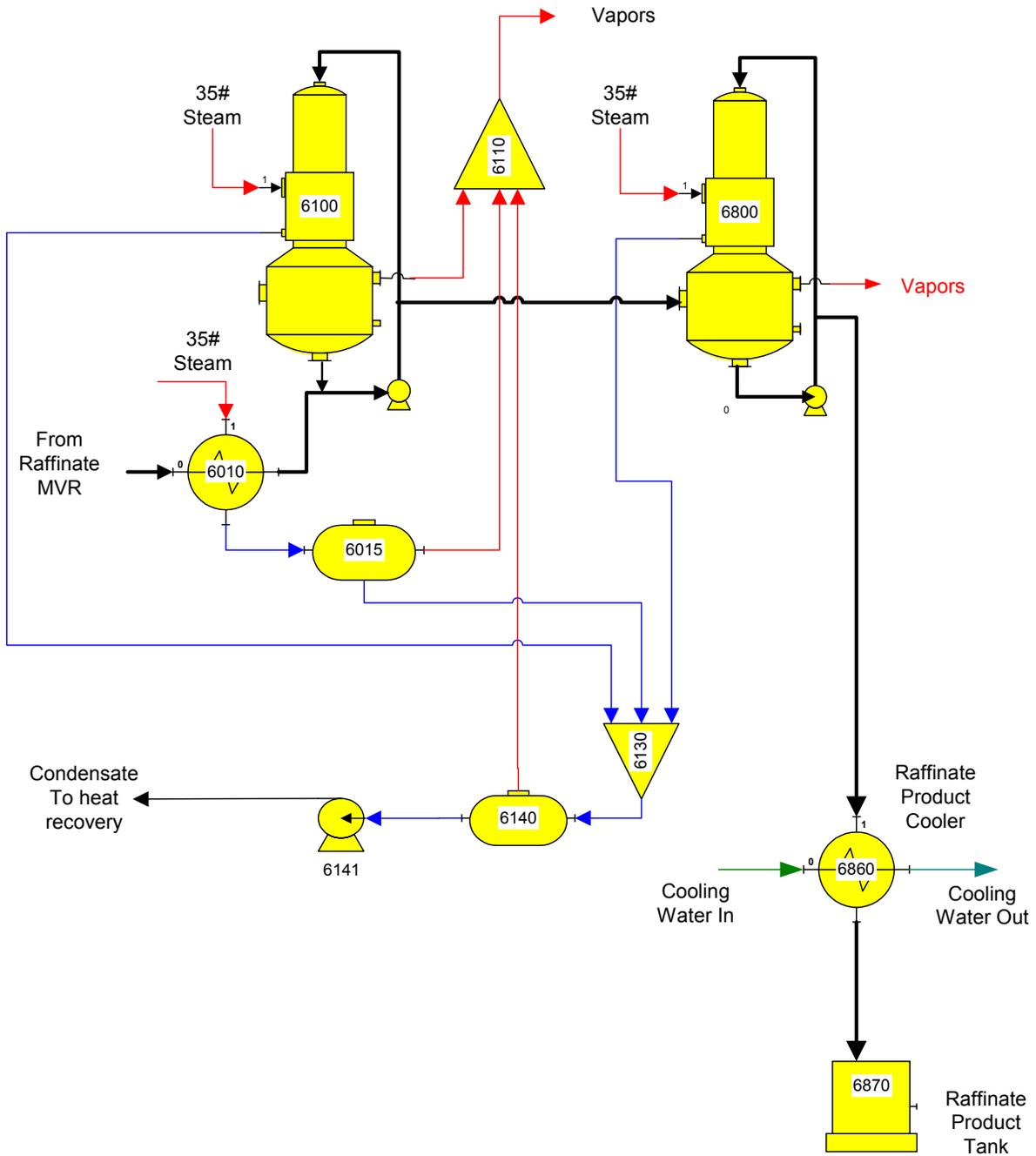
The final design is a combination of the configuration components described. Below is a diagram indicating the final raffinate arrangement.



Computer modeling of the MDS facility using the Sugars® software was completed prior to startup. The model of the entire facility could calculate the overall steam requirements of which the evaporator system is the main contributor. The model also provided the means to optimize the operating conditions of the evaporator train during the design phase given the criteria established during the design.¹ Models of the Raffinate MVR and Raffinate Finisher from the computer model are shown below.

¹ The MDS facility was actually designed for multiple operating scenarios; however, the single set of design parameters that best fit the current operation have been selected for evaluation purposes.

Raffinate Finishers



Startup and Operation

The startup and continuous operation of the evaporator system began in January of 2000. The evaporator was the first major process system to start, since it supplied the process condensate for molasses dilution and separator feed water at the higher temperatures required. Then, the continuous operation of the evaporator handled the dilute product fractions from the startup of

the separator. As the throughput through the separators quickly increased, the evaporator successfully kept up and handled the varying range of evaporative loads.

The modeling runs allowed for a check of the original design parameters and provided a means to review more realistic parameters that would likely be more consistent with the eventual operating numbers. The new facility steam consumption was not allowed to exceed the limits dictated during the design phase. In addition to steam consumption, other parameters were also monitored, including:

- Feed and product flow rates, brix and temperatures,
- Pressure and temperature,
- Boiling point elevation,
- Evaporation capacity,
- Electrical usage,

The following data is a comparison of the design, model and operating parameters from the raffinate system.² The data compiled from operation was taken 12/12/00.

Feed and Product

The most basic feed and product parameters were as follows. All values are based on the normal capacity of the MDS facility of **600 US short tons per day**.

	Raffinate MVR A Feed			Raffinate MVR A Product		
	Flow, lb/hr	Brix	Temp., °C	Flow, lb/hr	Brix	Temp., °C
Design	115,510	5.31	80°	21,905	28	102.2°
Model	115,309	5.31	79.4°	21,891	28	102.2°
Operation	121,373	5.0	79.3°	21,072	28.8	103.7°

	Raffinate MVR B Feed			Raffinate MVR B Product		
	Flow, lb/hr	Brix	Temp., °C	Flow, lb/hr	Brix	Temp., °C
Design	115,510	5.31	80°	21,905	28	102.2°
Model	115,309	5.31	79.4°	21,891	28	102.2°
Operation	120,835	5.0	79.3°	20,978	28.8	103.1°

Due to the arrangement of the MVR to finisher evaporators, the mass out of the MVR does not equal exactly the mass into the finisher during actual operation. As can be seen in the evaporator system flow diagram earlier in this paper, a surge tank is used between the MVR evaporator to the finisher evaporator. This arrangement allows some flexibility in flows and provides some catch-up time for either system. At the time operating data was being collected, the finisher evaporator was running at a slightly higher throughput than the MVRs. In addition, the MVR product temperatures slightly greater than 100°C result in a small amount of flash to take place in this atmospheric surge tank.

² The data compiled from operation was taken 12/12/00.

	Raffinate Finisher Stage 1 Feed			Raffinate Finisher Stage 1 Product		
	Flow, lb/hr	Brix	Temp., °C	Flow, lb/hr	Brix	Temp., °C
Design	43,811	28	102.2°	29,207	42	117.2°
Model	43,782	28	102.2°	29,182	42	113°
Operation	44,503	28.8	90.9°	30,379	42.19	87.9°

	Raffinate Finisher Stage 2 Feed			Raffinate Finisher Stage 2 Product		
	Flow, lb/hr	Brix	Temp., °C	Flow, lb/hr	Brix	Temp., °C
Design	29,207	42	117.2°	17,525	70	93.9°
Model	29,182	42	113°	17,509	70	80.6°
Operation	30,379	42.19	87.9°	18,782	68.24	75.4°

The stage 2 of the finisher is the forced circulation unit. The operating requirement is that the product brix from stage 1 finisher or feed brix to stage 2 forced circulation finisher should never be greater than 42%DS. This limit protects stage 1 from excessive fouling from precipitation.

Pressure and Temperature

The operating pressures and temperatures in the finisher evaporator had the most surprising deviation from the design parameters. The pressures and temperatures are important because of the use of vapors from the raffinate finishers to evaporate other streams and because of the use of the process condensate as molasses dilution and separator feed water. Due to this use, process condensate is preferred as close to 85°C as possible to reduce steam consumption for heating.

A measurement of pressure is made in each vapor separator of the evaporator corresponding to the pressure that boiling occurs and the pressure of the vapor. A measurement of temperature is made of the boiling liquid in the circulation at each steam chest. This temperature is the boiling point of the liquid or raffinate solution at the pressure measured. The pressure (P in psia) and temperature (T in °C) were as follows.

	Raffinate MVR A					
	Feed	Preheat	Pass 1	Pass 2	Pass 3	Pass 4
Design P	---	---	14.6 psia	14.6 psia	14.7 psia	14.8 psia
Design T	80°C	99.4°C	100°C	100.6°C	101.1°C	102.2°C
Model P	---	---	14.7 psia	15.0 psia	15.2 psia	15.8 psia
Model T	79.4°C	99.4°C	100°C	100.6°C	101.1°C	102.2°C
Oper. P	---	---	15.1 psia	13.2 psia	13.2 psia	15.1 psia
Oper. T	79.3°C	96.9°C	100.4°C	100.7°C	102°C	103.7°C

Raffinate MVR B						
	Feed	Preheat	Pass 1	Pass 2	Pass 3	Pass 4
Design P	---	---	14.6 psia	14.6 psia	14.7 psia	14.8 psia
Design T	80°C	99.4°C	100°C	100.6°C	101.1°C	102.2°C
Model P	---	---	14.7 psia	15.0 psia	15.2 psia	15.8 psia
Model T	79.4°C	99.4°C	100°C	100.6°C	101.1°C	102.2°C
Oper. P	---	---	15.3 psia	14.8 psia	14.8 psia	15.3 psia
Oper. T	79.3°C	97.1°C	100.7°C	101.3°C	101.9°C	103.1°C

The pressures given above for the raffinate MVRs are for the vapor from boiling going to the suction of the mechanical compressor. The discharge of the mechanical compressor had a design pressure of approximately 18.2 psia, while the operating data gave a compressor discharge pressure of 17.8 to 18.2 psia for the raffinate applications.

Raffinate Finisher				
	Feed	Preheat	1 st Effect, Stage 1	1 st Effect, Stage 2
Design Pressure	---	---	22.8 psia	6.0 psia
Design Temp.	102.2°C	116.7°C	117.2°C	93.9°C
Model Pressure	---	---	22.8 psia	6.0 psia
Model Temp.	102.2°C	116.7°C	113°C	80.6°C
Oper. Pressure	---	---	7.4 psia	3.75 psia
Oper. Temp.	90.9°C	100.5°C	87.9°C	75.4°C

Boiling Point Elevation (BPE)

The temperature and pressure operating data allowed for an opportunity to evaluate the actual boiling point elevation of the streams being evaporated. A comparison of boiling point elevation values might also provide some explanation for the change in operating pressures encountered. However, because of the change in operating pressure, a direct comparison between design and operating boiling point values is not possible since pressure influences the boiling point elevation value. Therefore, a typical formula (Bubnik, page 170, 322/20) for determining boiling point elevation has been used to try to bring design and operating to a common basis. The boiling point elevation formula includes pressure that boiling occurs, solution purity, and DS or Brix of the solution. In addition, for these comparisons, the boiling point elevation was considered significant only at higher concentrations; therefore, the values for the raffinate finisher are considered only.

	Raffinate Finisher	
	1 st Effect, Stage 1	1 st Effect, Stage 2
<i>Brix</i>	42.19	68.24
<i>Pressure, psia</i>	7.4 psia	3.75 psia
<i>Purity</i>	14	14
BPE from Formula, °C	3.9°C	10.9°C
BPE from Operating Data, °C	6.1°C	9.7°C

Based on this comparison of boiling point elevation values estimated from a formula versus calculated from operating data, the original estimates of BPE values for raffinate were reasonable. In addition, the formula would also likely have the same success with other MDS streams. The referenced formula provides an acceptable estimate for boiling point elevation of sugar solutions of varying purity.

Evaporative Capacity

The MDS evaporator system was designed for a specific evaporative capacity, and this evaporation requirement is being satisfied. As discussed earlier in this paper, the raffinate has the largest evaporative load requirement of any of the MDS separator products, which is why it has been given such a high level of attention. In addition, the raffinate evaporation makes the best use of the MVR evaporator technology since almost 90% of the raffinate water removal takes place in the MVR units. Evaporative loads are given below.

	Design	Operating
Raffinate MVR Evaporative Load:	187,209 lb/hr	200,158 lb/hr
Raffinate Finisher Evaporative Load:	26,286 lb/hr	25,721 lb/hr
Total Raffinate Evaporative Load:	213,495 lb/hr	223,426 lb/hr
Total Evaporative Load (All streams):	379,522 lb/hr	417,767 lb/hr

Steam and Electrical Consumption

The MDS evaporator system successfully combines MVR evaporators with finisher evaporators. There is a significant trade-off between steam consumption reduction and higher electrical power requirements. The total of three MVR evaporator compressors required a maximum design power of 2,536 kilowatts. However, during normal operation, 80% of the maximum kilowatts was required.

The MDS evaporator system resulted in outstanding reductions in steam consumption. This was achieved with both the use of MVR evaporators and the finisher evaporator's multiple effect configuration for efficient use of the steam that is consumed. Evaporators have been used for MDS facilities without utilizing MVR systems. However, experience has shown that for similar MDS facilities and capacities, steam rates of approximately 85,000 to 100,000 lb/hr are required. These earlier installations used only multiple effect evaporator systems.

To date, steam consumptions for the evaporator system at this new MDS facility have ranged from approximately 40,000 to 50,000 lb/hr. This operating data for current steam rates is also in good comparison to the original design steam rate of 58,000 lb/hr. Therefore, the use of the MVR units has cut the steam requirements in half for an MDS facility.

Using the evaporative loads presented earlier and simplified relationships of steam rates to evaporation rates, a quick evaluation can be of how many effects it would take in a multiple effect evaporator system to achieve the same kinds of steam usages. Using the simplified ratio for number of effects in a multiple effect evaporator system would result in the following.

Number of Effects	Ratio (Steam / Evap. Load)	Steam Required, lb/hr
1	1/1	417,767
2	1/2	208,884
3	1/3	139,256
4	1/4	104,442
5	1/5	83,553
6	1/6	69,628
7	1/7	59,681
8	1/8	52,221
9	1/9	46,419
10	1/10	41,777

In this simplistic evaluation it would require ten effects for a multiple effect evaporator system to meet the same evaporative load as the MDS facility evaporator discussed in this paper.

Conclusion

The application of MVR evaporators requires an in depth analysis of the nature and volume of the evaporative loads as well as an understanding of the configuration alternatives available. Based on the system requirements, specifications, and limitations described, a system of mechanical vapor recompression units for initial concentration to remove large quantities of water followed by multiple-effect, steam heated units to finish evaporate the process streams is an excellent solution for the MDS application presented. The operating data presented in this paper demonstrate the excellent steam saving advantages of MVR evaporators and the ability of MVR equipment to operate successfully in a molasses desugarization process for the sugar beet industry. It would also be reasonable to conclude that MVR evaporators would have similar success in other evaporation applications in the sugar beet processing factory.

Acknowledgements

1. The evaporator system described in this paper is part of the MDS facility owned by Crystech, LLC and operated as part of the American Crystal Sugar Company factory in Hillsboro, ND. The data presented is used by permission from the American Crystal Sugar Company, Moorhead, MN.
2. The evaporator system described in this paper was supplied by the Dedert Corporation. The evaporator design information presented is used by permission from the Dedert Corporation, 20000 Governors Drive, Olympia Fields, IL.

References

Bubnik, Z., et. al., Sugar Technologists Manual, Chemical and Physical Data for Sugar Manufacturers and Users, 8th edition, Bartens, 1995.