Energy, Exergy analysis and comparison of conventional multiple effect evaporation and pressure driven membrane distillation plus evaporation process in Indian sugar industry.

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Abstract
Sugarcane juice concentration by evaporation is an energy intensive process in conventional sugar factories. In the present scenario since energy conservation and economy is of prime importance, low energy consuming processes such as membrane technologies save large amounts of energy due to the absence of change in phase. The present work elaborates the benefits of replacing the conventional evaporation system with membrane distillation plus evaporation process both from energy and exergy point of view. A sugar manufacturing facility’s statistics is used from beyond literature. The analyses display fine consequences for substitution co-green in phrases of strength and for sustainability index in terms of exergy. If the evaporation station within the present sugar factories is changed via the proposed membrane distillation plus evaporation system, it would result in widespread advantages in terms of efficiency and environment.

Keywords: sugarcane, evaporation, energy conservation, membrane distillation, substitution co-efficient, energy, sustainability, exergy, sugar.

INTRODUCTION
The Indian economy is agriculture driven and sugarcane production and its usage takes a prominent place. Concentration of thin sugarcane juice by evaporation is one of the most important and energy intensive operation in sugar factory. Clarified juice enters into multiple effect evaporators (MEE) with 14-16 Brix (dry substance in the solution) and is concentrated to thick syrup with 60-65 Brix. This process is primarily based on Rillieux principle (Hugot E., 1986). Since the process of evaporation involves a change in phase and is energy intensive, it has attracted the attention of lot of scientists and energy investigators. Literature available reveals regarding retrofitting, simulation, analysis, mathematical modeling of MEE system in sugar factory (Urbaniec K. et al., 2000; A.E. Lewis, et al., 2010; L.M.M. George et al., 2010; M. Higa, et al., 2009; Durmus Kaya, and H. Ibrahim Sarac, 2007). Available literature (Adriano V. Ensinas et al., 2007) also shows that the process steam demand for sugar and ethanol plants can be reduced significantly with process thermal integration.

Literature discussed till now deals with energy aspects of MEE system. But exergy analysis is far beneficial in identifying the locations and magnitudes of losses as compared to energy analysis, which is well documented (Adrian Bejan, 2002; M.A. Rosen, 1999; Dincer I. and Rosen M. A. 2007). Some researchers carried out pinch and exergy analysis of MEE system of a sugar factory (J. Raghu Ram and Rangan Banerjee, 2003). It is shown that quintuple effect evaporation system is more beneficial than quadruple effect in terms of energy and exergy. Evaluation and comparison of quintuple effect evaporation systems is also found in the past literature (S. M. Bapat et al., 2012). It is found that
employing heat recovery devices for MEE system could reduce energy consumption but is not attractive in terms of exergy. The literature discussed till now show that there is not much scope for energy consumption minimization and optimization in the present MEE system. Finding low energy consumption devices has always been a challenge for energy investigators. In this context membrane distillation processes offer energy effective methods for dehydration of sugarcane juice due to the absence of phase change. Past literature shows that a lot of work is done in employing membrane processes, such as reverse osmosis (RO), nano-filtration (NF), ultra-filtration (UF) etc. in sugar syrup concentration (R.F.Madsen, 1973; Kshihara, S. et al., 1981; Balakrishnan M. et al., 2000; Nene S. et al., 2002; Ghosh A.M. and Balakrishnan M. 2003). A comparison of evaporation and reverse osmosis system is also carried out in sugar industry in terms of energy consumption (S S Madaeni and S Zereshki, 2010). It is shown that a 33% reduction in terms of energy consumption is possible by employing reverse osmosis in combination with conventional evaporation process. Some researchers used multistage pressure-driven membrane concentration process by using reverse osmosis and nano-filtration membranes with different rejections and recycle streams (Saeed Gul and Micheal Hasasek, 2012). It showed that the energy savings relative to the conventional MEE system is to the extent of 84%.

But however it is found that very little information is available regarding exergy analysis of membrane distillation plus evaporation processes employed in sugar industries. Also unlike past studies, the present work compares conventional MEE system (with five effects) with energy effective membrane processes in terms of exergy. The actual operating data from a sugar factory using conventional MEE system is utilized. A combination of evaporation and membrane driven process in sugar syrup concentration is proposed. It is evaluated and compared with past literature in terms of energy and exergy. The following sections discuss initially regarding conventional sugar manufacturing and the mechanism of multiple effect evaporation process. The proposed membrane distillation plus evaporation process is explained in detail, followed by its analysis, comparison, results and discussion.

Conventional sugar manufacturing process

A simplified block diagram of sugar production process is shown in figure 1. Sugarcane is initially washed and cleaned. The cane is brought to the required size and shape using cutter, knives and shredders. The juice from the cane is extracted using milling arrangement. The cane is squeezed under high pressure between successive pairs of mills. The juice extracted, generally has a temperature of 30 to 35°C. This juice is preheated to a temperature of 108-110°C using tubular heaters. Sulphitation and lime addition takes place during preheating. This is in order to maintain pH value, removal of colored matter, defecation/ clarification.

The clarified juice which is very near to boiling point is fed to the evaporators where the juice is evaporated using exhaust steam. This is done using multiple effect evaporators (MEE) for maximum energy economy. Part of the vapor generated during evaporation is used for juice preheating, milling and crystallizing activities. The syrup from the evaporators is sent to the boiling house and then to pan crystallizers. Pan crystallizers involve a process of converting evaporator syrup into crystals of sugar under vacuum conditions. During centrifugation the sugar crystals are separated from the massecuite.
by centrifugal action. The sugar is then dried and bagged as an end product.

**Multiple effect evaporation**
This principle was developed by Norbert Rillieux (Hugot E., 1986). The evaporators are usually connected in series (refer figure 2). Steam usually in the range of 120-125°C is used to heat juice in the first evaporator. The juice in turn is sent to the second evaporator for which the heating media used is the vapor generated from the juice in the first evaporator. Thus instead of steam, vapor generated in the previous evaporator is used as a heating media for subsequent evaporator. This helps to improve steam economy. But since a temperature difference is necessary between the heating fluid and the fluid to be heated, Rillieux resolved this difficulty by putting subsequent evaporators under vacuum conditions. Operation under vacuum also proved to be advantageous from the point of view of inversion and of coloration of the juice.

According to Rillieux in a multiple effect evaporation of N effects, one kg of steam will evaporate N kg of water (theoretically). Nearly all cane sugar mills have an evaporator system comprising of 3 to 5 effects. In Indian sugar industry scenario lot of sugar mills are using quintuple effect evaporation station (5 effects) with a view to increase steam economy. Also it is already proved that quintuple effect evaporation system is advantageous both in terms of energy and exergy as compared to quadruple effect evaporation system (J. Raghu Ram and
Rangan Banerjee, 2003). It is a general practice, that in order to attain energy/steam economy, vapor from initial effects is bled for heating duties in juice preheating and pan boiling stations.

**Conventional evaporation process**

In the present context a case examines is taken into consideration in which evaporation is performed in 5 outcomes (S. M. Bapat et al., 2012). A detailed pictorial description regarding evaporation, vapor bleeding, juice preheating and pan station is given in determine three. Juice preheating activity is executed the usage of vapor from first, 2nd, 1/3 and 5th results.

The pan boiling station utilizes vapor from first and 2nd impact. Exhaust steam is used as a heating media and is supplied to the primary effect of evaporation station at 1250C and 2.5 bar (abs). The steam consumption is found to be 26.58 kg/s for an average cane crushing rate of 250 tons/hr (i.e. 69.44 kg/s). Hence the exhaust steam demand (ESD) for conventional process is found to be 38.56% on cane. The present study analyzes the possibility of partly replacing the energy intensive evaporation process with membrane + evaporation process in order to remove the water content in the juice.

**Fig 3:** Conventional Multiple effect evaporation (MEE) process with juice pre-heaters and pan station.

**Proposed membrane plus evaporation concentration process**

Consider figure 4. The juice preheating consists of 3 raw juice heaters (RJH), 2 sulphited juice heaters (SJH) and a preheater. The heating step and the heating media used for these heaters are shown in table 1. The juice gets heated to a temperature of 102°C before entering the clarifier. After clarification the juice temperature lowers to 97°C due to juice settlement or detainment. After juice clarification the clear juice enters the plate heat exchangers (PHE; liquid-liquid heater) where a drop in
temperature is noticed from 97°C to 80°C because it is used to heat the raw juice in RJH-2. The drop in temperature in PHE has 2 benefits: Firstly a drop in temperature to 80°C is conducive for the process of membrane distillation, and secondly this heat would be utilized to heat raw juice in RJH2 which would facilitate energy economy. Use of liquid-liquid type plate heat exchangers (PHE) is more advantageous than using a shell and tube kind of heat exchanger because they have high heat transfer coefficients, more compact and lower liquid holdup. The mixed or raw juice acting as a cold fluid in PHE, consists of bagasse fiber which would tend to block the gap between adjacent plates. Hence ‘wide gap’ plates are used for PHE (Peter Rein, 2007). In the present context it is assumed that 60% of heat from the hot fluid is transferred to the cold fluid. Based on this the heating steps for RJH-2 and PHE are decided. Moreover utilization of liquid-liquid plate heat exchangers is one of the ways to attain energy economy (Peter Rein, 2007).

Clarified juice at 80°C and 15 brix (amount of solid content) is fed to the multistage pressure driven membrane concentration system. In the membrane, juice gets concentrated because of the applied pressure to concentrate the solution by forcing water through semi permeable membrane. It allows water to pass through but prevents the passage of solutes. The multistage membrane used here is a combination of RO (reverse osmosis) and NF (nano-filtration) membranes with different rejection and recycles streams (Saeed Gul and Micheal Hasasek, 2012), with an operating pressure of 32 bars. The cane juice gets concentrated to 50 brix at 80°C and is fed to the juice pre-heater where the temperature rises to 110°C. The heating media used for juice pre-heater is the exhaust steam. Also preheating juice close

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**Table 1: Description of juice pre-heating heaters network.**

<table>
<thead>
<tr>
<th>stream</th>
<th>heating step</th>
<th>T&lt;sub&gt;in&lt;/sub&gt; (°C)</th>
<th>T&lt;sub&gt;out&lt;/sub&gt; (°C)</th>
<th>Heating media</th>
</tr>
</thead>
<tbody>
<tr>
<td>Raw juice heating (RJH)</td>
<td>1</td>
<td>30</td>
<td>42.10</td>
<td>III vapor</td>
</tr>
<tr>
<td></td>
<td>2</td>
<td>42.10</td>
<td>52.30</td>
<td>Clear juice from PHE</td>
</tr>
<tr>
<td></td>
<td>3</td>
<td>52.30</td>
<td>72</td>
<td>Exhaust steam</td>
</tr>
<tr>
<td>Sulphited juice heating (SJH)</td>
<td>1</td>
<td>72</td>
<td>92</td>
<td>Exhaust steam</td>
</tr>
<tr>
<td></td>
<td>2</td>
<td>92</td>
<td>102</td>
<td>Exhaust steam</td>
</tr>
<tr>
<td>Juice pre-heater</td>
<td>1</td>
<td>80</td>
<td>110</td>
<td>Exhaust steam</td>
</tr>
</tbody>
</table>
to boiling point before entering the evaporator, improves the energy economy in the evaporation process. This is an efficient use of the heating surface and improves the rate of evaporation (Peter Rein, 2007). Compared to quintuple effect evaporation system in the conventional process, in the present context a triple effect evaporator is used. First vapor is used as heating media for second evaporator and in turn the vapor generated in the second evaporator is used as a heating media for the third evaporator. Vapor from third effect is used for heating raw juice in RJH-1. Exhaust steam is used as heating media for evaporation, which is fed to the first evaporator. The cane juice gets concentrated in the form of syrup in the triple effect evaporator from 50 brix to 67.79 brix. The concentrated syrup from third effect is then sent to crystallization which has a 3-scheme pan station (A,B,C). The heating media for pan station is exhaust steam.

Thus the conventional quintuple effect evaporation system is replaced by a plate heat exchanger (PHE), reverse osmosis (RO) and nano-filtration (NF) membrane, a juice pre-heater and a triple effect evaporation station. A detailed pictorial presentation of the proposed process with its essential components and salient state points is shown in figure 5. The juice preheating and pan station would be retained as before. The details of exhaust steam requirements for the proposed membrane + evaporation process are shown in table 2.

![Diagram](image)

**Fig 5:** A detailed view of the proposed membrane plus evaporation process with salient state points.

<table>
<thead>
<tr>
<th>Stream</th>
<th>Exhaust steam demand (ESD) (kg/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Raw juice heater-3 (RJH-3)</td>
<td>2.44</td>
</tr>
<tr>
<td>Sulphited juice heater-1 (SJH-1)</td>
<td>2.48</td>
</tr>
<tr>
<td>Sulphited juice heater-2 (SJH-2)</td>
<td>1.24</td>
</tr>
<tr>
<td>Juice evaporation</td>
<td>2.48</td>
</tr>
<tr>
<td>Pan boiling</td>
<td></td>
</tr>
<tr>
<td>Pan A</td>
<td>5.48</td>
</tr>
<tr>
<td>Pan B</td>
<td>2.00</td>
</tr>
<tr>
<td>Pan C</td>
<td>1.98</td>
</tr>
<tr>
<td>Juice pre-heater</td>
<td>0.86</td>
</tr>
<tr>
<td>Total</td>
<td>18.96</td>
</tr>
</tbody>
</table>

**Table 2:** Details of exhaust steam demand/consumption.

**Analysis**

The mass balance for steady state steady flow process is as given below,

\[ \sum m_{in} = \sum m_{out} \quad \ldots \ldots \ldots \ldots (1) \]

suffixes ‘in’ and ‘out’ indicate inlet and outlet conditions respectively.
The solid mass balance is given by,

\[ m_{in} B_{in} = m_{out} B_{out} \]  

(2)

\( m \) is mass flow rate of the juice and \( B \) is brix (amount of solid content) of the cane juice. Equation 2 is applied to all the three evaporators in order to estimate the amount of vapor generated. The % of cane juice evaporated in the evaporator station is evaluated using, (E. Hugot, 1986),

\[ \% \text{evaporation} = \frac{B_{out} - B_{in}}{B_{out}} \times 100 \]  

(3)

The % evaporation taking place in the proposed process is calculated using equation 3 and is shown in table 6. Exhaust steam is used as heating media for sensible heating of cane juice during juice preheating. It is estimated based on energy balance equation as below (Peter Rein, 2007),

\[ m_s h_{fg} = m_j C_{pm} \Delta T \]  

(4)

\( h_{fg} \) (2188.05 kJ/kg) is the latent heat at 125\(^\circ\)C. \( C_{pm} \) is the mean specific heat of the juice. This equation is applied assuming that the juice specific heat capacity remains constant. The condensed steam leaves at steam vapor saturation temperature. Heat losses are negligible.

Exhaust steam is used as a heating media for RJH-3, SJH-1, SJH-2 and juice preheater. The estimated exhaust steam demand/consumption for these components is calculated using equation 4 and is shown in table 2. Juice preheating for RJH-2 uses clear juice in plate heat exchanger as a heating media. Since it is liquid-liquid plate heater an energy balance between the hot and cold fluid can be established assuming that the heat lost by the hot fluid is equal to the heat gained by the cold fluid. But under practical operating conditions of the nature of raw juice, use of wide gap plates in PHE and the juice hold up time, it is assumed that 60% of the heat from the hot fluid is transferred to the cold fluid. Now the temperature rise of the cold fluid in RJH-2 is estimated using the equation,

\[ 0.60 \left( m_h \Delta C T_h \right) = m_c \Delta C T_c \]  

(5)

Suffix ‘\( h \)’ indicates hot fluid and ‘\( c \)’ indicates cold fluid. The estimated value of temperature rise of the cold fluid in RJH-2 is shown in table 1. Exhaust steam used as a heating media for juice evaporation is estimated using equation (L.M.M. George et al., 2010),

\[ S_i = \frac{F_i \left( C_{pm} \Delta T_j \right) + V_i h_{fg}^i}{h_{fg}^i} \]  

(6)

Where mean specific heat of juice is given by \( C_{pm} = 4.186 - 2.512 B_m (C_{pm} \) is measured in kJ/kg K). The estimated value of exhaust steam consumption for evaporation is shown in table 2. The details of vapor as heating media used for the subsequent evaporators is also calculated using equation 6 and the values are shown in table 3. Juice preheating in RJH-1 is done using III vapor at 70\(^\circ\)C as the heating media. Based on the vapor generation rate in III effect the temperature rise of raw juice is calculated using equation 4. However the raw juice from mills is fixed at an average temperature of 30\(^\circ\)C. Exhaust steam is also used as a heating media for pan boiling i.e. pan A, pan B and pan C. It is estimated using the equation (Hugot E., 1986),

\[ \text{Steam consumption} = Q km \]  

(7)

Where \( Q \) is flow rate of massecuite in the pan section measured in terms of kg /hr tch. The values of constant \( k \) for pan A massecuite is 0.275, for pan B massecuite it is 0.3 and for pan C massecuite it is 0.360. The values of constant \( m \) for batch pan operation is 1.5 and for continuous pan operation it is 1.2. In the existing plant condition Pan A and C are of batch type
whereas Pan B is horizontal continuous type (S. M. Bapat et.al. 2012). A demonstration of the method of the massecuite flow rate for pan station with three massecuite system (i.e. CBA) is given (Hugot E., 1986). The assumed values of factory conditions for juice syrup and massecuite prevailing under factory conditions are shown in table 4. Depending upon the amount of syrup flow rate to pan A from the third effect evaporator and an average cane crushing rate of 250 tons per hour the exhaust steam consumption for pan boiling is estimated using equation 7 and is shown in table 2. The process of evaporation+ membrane distillation needs electrical energy also. The process of membrane distillation needs high pressure pump to remove water from the solution. The theoretical amount of energy required to separate water from a solution by distillation is estimated using the equation (Saeed Gul, and Micheal Harasek, 2012),

$$E_p = \frac{q_j \Delta p}{\eta_{\text{pump}}} \quad \text{(8)}$$

The efficiency of the high pressure piston pump ($\eta_{\text{pump}}$) is generally between 50% and 90%. Here it is assumed to have an average value of 70%. The present membrane process operates at a trans-membrane pressure of 32 bars with juice temperature of 80°C. Juice flow rate $q_j$ is volumetric i.e. m$^3$/sec, whereas the available flow rate of clarified juice is gravimetric i.e. kg/sec. The mass flow rate of juice at the inlet of the membrane is 71.52 kg/s at 15 brix and 80°C. The mass density of clarified juice can be found by (Pacheco et.al. 1999),

$$\rho_{\text{CLARIF}} = 1000 \left[1 + \frac{B(B + 200)}{54000} \right] \left[1 - \frac{0.036(T - 20)}{(160 - T)} \right] \quad \text{(9)}$$

Where T is the temperature of juice in °C, and B is the brix content in %. The clarified juice volumetric flow rate would be,

$$q_j = \frac{m_j}{\rho_{\text{CLARIF}}} \quad \text{(10)}$$

Juice evaporators need condensate extraction pumps (CEP) of centrifugal type. The power required for these three pumps for the three evaporators is evaluated by using the equation (Peter Rein, 2007),

$$P = \frac{\rho_2QH}{\eta_{\text{pump}}1000} \quad \text{kw} \quad \text{(11)}$$

Centrifugal pumps have an efficiency ranging from 50% to 70%. Here an average efficiency of 60% is assumed. The head against which the pump needs to work is assumed as 25 meters. The capacity of syrup extraction pump (SEP) in the existing plant capacity is 37 kW, which is kept unchanged. The electrical pumping requirements are estimated using equation 8 and 11 and are shown in table 5.

The conventional evaporation process and the proposed evaporation + membrane distillation consume electrical and thermal energy. In this context it is very important to estimate the advantage of using the proposed process in terms of energy. Past literature reveals that the method of substitution co-efficient is commonly used for membrane processes. Substitution co-efficient is defined by the ratio of primary energy (thermal) saved in the proposed process with respect to the conventional process and the amount of electrical energy consumed relative to the conventional process. Mathematically it is defined as (R. Molinari et al., 1995),

$$CS = \frac{C_1 - C_2}{E_2 - E_1} \quad \text{(12)}$$

Another non-dimensional parameter based on energy analysis particularly for Multiple effect evaporation (MEE) process
is the Steam Economy (SE), which is defined as (Balkan F. et al., 2005),

\[ SE = \frac{m_{s} \cdot \sum_{j} m_{v,j}}{m_{s}} \]  \hspace{1cm} (13)

Where \( m_{v,j} \) is the mass flow rate of the vapor produced in \( i^{th} \) effect and \( m_{s} \) is the mass flow rate of steam supplied to the MEE system. In the present context the equation for SE would be (refer fig.5 and table 6),

\[ SE = \frac{m_{1} + m_{2} + m_{7}}{m_{7}} \]  \hspace{1cm} (14)

Analysis undertaken so far is related to the energy aspects only. But rather than energy it is exergy or available energy which is important. Also rather than energy based methods, exergy based methods are quite useful because they represent qualitatively the ‘useful’ energy or the ability to do or receive work (Adrian Bejan, 2002). The present work on exergy analysis comprises of exergy balance, exergy destruction, exergy efficiency, improvement potential (IP) and exergy ratios. Lastly the assumptions considered are mentioned. The general exergy balance equation is given by,

\[ \sum \dot{E}_{in} - \sum \dot{E}_{out} = \sum \dot{E}_{dest} \]  \hspace{1cm} (15)

The specific flow exergy is given as,

\[ ex = (h - h_{o}) - T_{o} (s - s_{o}) \]  \hspace{1cm} (16)

Where \( T_{o}, s_{o} \) and \( h_{o} \) are temperature, entropy and enthalpy at the reference state.

For incompressible flows, the specific flow exergy is given as (Bejan A., 1988).

\[ ex_{in} = C \left[ (T - T_{o}) - T_{o} \ln \left( \frac{T}{T_{o}} \right) \right] \]  \hspace{1cm} (17)

The total exergy is given as,

\[ \dot{E}_{x} = ex \cdot m \]  \hspace{1cm} (18)

The total exergy destruction taking place in the proposed membrane + evaporation process could be evaluated by knowing the exergy destruction rates of individual components. Applying exergy balance equation to membrane and pre-heater we have,

\[ \dot{E}_{x1} + \dot{E}_{x2} - \dot{E}_{x3} - \dot{E}_{x4} = \dot{E}_{xdest,mem} \]  \hspace{1cm} (19)

\[ \dot{E}_{x3} + \dot{E}_{x5} - \dot{E}_{x6} - \dot{E}_{x8} = \dot{E}_{xdest,ph} \]  \hspace{1cm} (20)

The general exergy balance for the evaporators would be,

\[ \dot{E}_{xsteam} + \dot{E}_{xpump} + \dot{E}_{xproductin} = \dot{E}_{xdest} + \dot{E}_{xproductout} + \dot{E}_{xcond} + \dot{E}_{xvap} \]  \hspace{1cm} (21)

Applying same principles, the exergy balance equation for three evaporators would be,

\[ \dot{E}_{x_{dest,1}} = \dot{E}_{x7} + \dot{E}_{x8} + \dot{E}_{x_{pump}} - \dot{E}_{x9} - \dot{E}_{x10} - \dot{E}_{x11} \]  \hspace{1cm} (22)

\[ \dot{E}_{x_{dest,II}} = \dot{E}_{x9} + \dot{E}_{x11} + \dot{E}_{x_{pump}} - \dot{E}_{x12} - \dot{E}_{x13} - \dot{E}_{x14} \]  \hspace{1cm} (23)

\[ \dot{E}_{x_{dest,III}} = \dot{E}_{x12} + \dot{E}_{x13} + \dot{E}_{x_{pump}} - \dot{E}_{x15} - \dot{E}_{x16} - \dot{E}_{x17} \]  \hspace{1cm} (24)

The exergy efficiency of membrane and pre-heater are evaluated as,

\[ \eta_{ex} = 1 - \frac{\dot{E}_{x_{out}}}{\dot{E}_{x_{in}}} = \frac{\dot{E}_{x_{dest}}}{\dot{E}_{x_{in}}} \]  \hspace{1cm} (25)
The evaporators are nothing but heat exchangers. The exergy efficiency for a heat exchanger is defined as (Yilmaz M., 2001),

\[
\eta_{ex} = \frac{E_{x_{cold,\text{out}}} - E_{x_{cold,\text{in}}}}{E_{x_{hot,\text{in}}} - E_{x_{hot,\text{out}}}} = \frac{\Delta E_{x_{cold}}}{\Delta E_{x_{hot}}} \quad \ldots \ldots \ldots \ldots \ldots (26)
\]

For example, application of the equation 26 to the first evaporator would lead to the equation,

\[
\eta_{ex} = \frac{E_{x_{11}} + E_{x_{9}} - E_{x_{8}}}{E_{x_{7}} - E_{x_{10}}} \quad \ldots \ldots \ldots \ldots \ldots (27)
\]

Another important parameter is the concept of exergetic improvement potential (IP) which is given by (Van Gool W., 1997),

\[
IP = (1 - \eta_{ex}) \left( \dot{E}_{x_{\text{in}}} - \dot{E}_{x_{\text{out}}} \right) = (1 - \eta_{ex}) \dot{E}_{x_{\text{dest}}} \quad \ldots \ldots \ldots (28)
\]

Total Exergy Destruction ratio (TExDR) is defined as the ratio of total exergy destruction of the system to the total exergy input to the system. Mathematically (Coskun C. et al., 2011),

\[
TExDR = \frac{E_{x_{\text{Tot,dest}}}}{E_{x_{\text{Tot,in}}}} \quad \ldots \ldots \ldots \ldots \ldots (29)
\]

In the present analysis the sum of exergy destruction rates in membrane, pre-heater, and three evaporators is taken as the numerator and the sum of total exergy content in the exhaust steam, cane juice and pumping power needed is taken as the denominator. In the present context the equation would be,

\[
TExDR = \frac{E_{x_{\text{dest,mem}}} + \dot{E}_{x_{\text{dest,ph}}} + \dot{E}_{x_{\text{dest,II}}} + \dot{E}_{x_{\text{dest,III}}} + E_{x_{\text{1}}} + \sum \dot{E}_{x_{\text{in,ex}}} + \sum E_{x_{\text{pump}}}}{E_{x_{\text{Tot,in}}}} \quad \ldots \ldots \ldots \ldots \ldots (30)
\]

Another useful exergy ratio named Exergetic Renewability ratio (ERR) is defined as the ratio of useful renewable exergy obtained from the system to the total exergy input to the system. Mathematically it is defined as (Coskun C. et al., 2011),

\[
ERR = \frac{E_{x_{\text{usf}}}}{E_{x_{\text{in}}} + \sum \dot{E}_{x_{\text{pump}}} + E_{x_{\text{1}}} + \sum E_{x_{\text{in,ex}}}} = \frac{\sum E_{x_{\text{usf}}} + \sum E_{x_{\text{productus}}} = E_{x_{4}} + E_{x_{11}} + E_{x_{12}} + E_{x_{17}} + E_{x_{3}} + E_{x_{9}} + E_{x_{13}} + E_{x_{15}} \ldots}{\sum \dot{E}_{x_{\text{pump}}} + E_{x_{1}} + \sum E_{x_{\text{in,ex}}}}
\]

\[
(31)
\]

The following assumptions are made during exergy analysis:
1. Only physical/flow exergy is taken into account.
2. The changes in kinetic and potential energies are not taken into account during exergy calculations.
3. All the evaporators are considered steady-state open thermodynamic systems.
4. The vapor in the evaporator is considered to be in saturated condition.
5. The changes in the ambient temperature conditions are neglected.
6. The reference conditions adopted are 25°C and 1.01325 bars.

In the present analysis a natural-environment-subsystem model is adopted for the reference condition defined in table 2 (Dincer, I. and Rosen, M. A., 2007), since it reflects the mean ambient temperature of the location of the system.
RESULTS
The calculated results of energy analysis are shown in tables 2, 3, 4, 5 and 6. Table 2 shows the values of Exhaust Steam Demand (ESD) for individual heaters as well as the total steam demand. The exhaust steam condition is taken as 125°C and 2.5 bar (abs). The highest value for ESD is 5.48 kg/s for Pan A and the lowest value of ESD is 1.24 kg/s for SJH-2. The total ESD for the proposed membrane + evaporation process is 18.96 kg/s. The electrical energy required for pumps in the proposed process is calculated using equations 8 and 11. The results are shown in table 5. The total pumping power requirement is 356.97 kW. Highest power is 317.25 kW which is consumed by the high pressure pump used in membrane process. The second highest power demand is for syrup extraction pump (SEP), which is due to high viscosity of the syrup.

**Table 3: Details of heating media (exhaust steam/ vapor) for each effect**

<table>
<thead>
<tr>
<th>Evaporator</th>
<th>I</th>
<th>II</th>
<th>III</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heating media</td>
<td>Exhaust steam</td>
<td>I vapor</td>
<td>II vapor</td>
</tr>
<tr>
<td>Heating media temp. (°C)</td>
<td>125</td>
<td>112</td>
<td>90</td>
</tr>
<tr>
<td>Latent heat of heating media $h_{fg}$ (kJ/kg)</td>
<td>2188.05</td>
<td>2224.5</td>
<td>2283.2</td>
</tr>
<tr>
<td>Vapor temp. (°C)</td>
<td>112</td>
<td>90</td>
<td>70</td>
</tr>
<tr>
<td>Vapor flow rate $V$ (kg/s)</td>
<td>2.39</td>
<td>1.83</td>
<td>1.41</td>
</tr>
<tr>
<td>Latent heat of vapor $h_{fg}$ (kJ/kg)</td>
<td>2224.5</td>
<td>2283.2</td>
<td>2333.9</td>
</tr>
<tr>
<td>Heating media flow rate $S$ (kg/s)</td>
<td>2.48</td>
<td>2.39</td>
<td>1.83</td>
</tr>
<tr>
<td>$B_{in}$ (%)</td>
<td>50.00</td>
<td>56.27</td>
<td>62.25</td>
</tr>
<tr>
<td>$B_{out}$ (%)</td>
<td>56.27</td>
<td>62.25</td>
<td>67.79</td>
</tr>
<tr>
<td>$B_{mm}$ (%)</td>
<td>53.13</td>
<td>59.25</td>
<td>65.01</td>
</tr>
<tr>
<td>$C_{pm}$ (kJ/kgK)</td>
<td>2.85</td>
<td>2.69</td>
<td>2.55</td>
</tr>
<tr>
<td>$\Delta T$ (°C)</td>
<td>2</td>
<td>22</td>
<td>20</td>
</tr>
</tbody>
</table>

**Table 4: Normal factory conditions for juice, syrup and massecuite (Hugot E., 1986)**

<table>
<thead>
<tr>
<th>Substance</th>
<th>Brix</th>
<th>Normal value of massecuite dry substance</th>
</tr>
</thead>
<tbody>
<tr>
<td>Clarified juice</td>
<td>15</td>
<td>----</td>
</tr>
<tr>
<td>Syrup</td>
<td>68</td>
<td>----</td>
</tr>
<tr>
<td>Massecuite A</td>
<td>93</td>
<td>191.4 kg/tch</td>
</tr>
<tr>
<td>Massecuite B</td>
<td>94</td>
<td>80.1 kg/tch</td>
</tr>
<tr>
<td>Massecuite C</td>
<td>97</td>
<td>52.8 kg/tch</td>
</tr>
</tbody>
</table>

**Table 5: Details of pumping power required for the proposed evaporation + membrane distillation process (kW).**

<table>
<thead>
<tr>
<th>Type of pump</th>
<th>Membrane distillation</th>
<th>I evaporator</th>
<th>II evaporator</th>
<th>III evaporator</th>
</tr>
</thead>
<tbody>
<tr>
<td>High pressure pump</td>
<td>317.25</td>
<td>-----</td>
<td>-----</td>
<td>-----</td>
</tr>
<tr>
<td>CEP</td>
<td>-----</td>
<td>1.01</td>
<td>0.97</td>
<td>0.74</td>
</tr>
<tr>
<td>SEP</td>
<td>-----</td>
<td>-----</td>
<td>-----</td>
<td>37.00</td>
</tr>
<tr>
<td>Total</td>
<td>317.25</td>
<td>1.01</td>
<td>0.97</td>
<td>37.74</td>
</tr>
<tr>
<td>Grand total</td>
<td>356.97</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 6 shows a comparison of ESD and % evaporation for the past literature with quintuple evaporation station and the proposed membrane + evaporation process. ESD value reduced from 26.78 kg/s to 18.96 kg/s and in terms of % of cane it reduced from 38.56% to 27.30%. The % of juice evaporation is hardly 26.24% in the proposed process as compared to 75% in conventional evaporation process. The electrical energy consumption increased from 340.85 kW to
356.97 kW, whereas the thermal energy consumption reduced drastically from 72654.14 kW to 51438.48 kW. It is shown that the total energy consumption savings is 29.04%.

Table 6: Comparison of parameters based on Energy analysis.

<table>
<thead>
<tr>
<th>Particulars</th>
<th>Conventional evaporation (S.M. Bapat et.al. 2012)</th>
<th>Proposed membrane+ evaporation</th>
</tr>
</thead>
<tbody>
<tr>
<td>$B_{in}$ (%)</td>
<td>15</td>
<td>50</td>
</tr>
<tr>
<td>$B_{out}$ (%)</td>
<td>60</td>
<td>67.79</td>
</tr>
<tr>
<td>% Evaporation</td>
<td>75</td>
<td>26.24</td>
</tr>
<tr>
<td>Exhaust Steam demand (kg/s)</td>
<td>26.78</td>
<td>18.96</td>
</tr>
<tr>
<td>Exhaust Steam demand (% of cane)</td>
<td>38.56</td>
<td>27.30</td>
</tr>
<tr>
<td>Electrical Energy consumption (kW)</td>
<td>340.85 ($E_1$)</td>
<td>356.97 ($E_2$)</td>
</tr>
<tr>
<td>Thermal Energy consumption (kW)</td>
<td>72654.14 ($C_1$)</td>
<td>51438.48 ($C_2$)</td>
</tr>
<tr>
<td>Total Energy consumption (kW)</td>
<td>72994.99</td>
<td>51795.45</td>
</tr>
<tr>
<td>% Energy savings</td>
<td>--</td>
<td>29.04</td>
</tr>
<tr>
<td>CS (MJ/kWhr)</td>
<td>--</td>
<td>4746.08</td>
</tr>
<tr>
<td>SE</td>
<td>1.99</td>
<td>2.27</td>
</tr>
</tbody>
</table>

Based on thermal and electrical consumption values, substitution coefficient (CS) is calculated using equation 12, and the resulting value is 4746.08 MJ/kWhr. Steam Economy (SE) is a very important parameter for evaporation process. It is calculated using equation 14, and the resulting value is 2.27 as shown in table 6.

Table 7: Thermodynamic properties and exergy rates at different state points.

<table>
<thead>
<tr>
<th>state point</th>
<th>Substance</th>
<th>$m$ kg/s</th>
<th>B (%)</th>
<th>p bar</th>
<th>$T$ °C</th>
<th>C kJ/kg K</th>
<th>ex kJ/kg</th>
<th>Ex kW</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>cane juice</td>
<td>71.52</td>
<td>15.00</td>
<td>----</td>
<td>80</td>
<td>3.80</td>
<td>17.74</td>
<td>1268.76</td>
</tr>
<tr>
<td>2</td>
<td>electricity</td>
<td>----</td>
<td>----</td>
<td>----</td>
<td>----</td>
<td>----</td>
<td>----</td>
<td>317.25</td>
</tr>
<tr>
<td>3</td>
<td>cane juice</td>
<td>21.45</td>
<td>50.00</td>
<td>----</td>
<td>80</td>
<td>2.93</td>
<td>13.67</td>
<td>293.39</td>
</tr>
<tr>
<td>4</td>
<td>water</td>
<td>50.07</td>
<td>----</td>
<td>----</td>
<td>80</td>
<td>4.19</td>
<td>19.56</td>
<td>979.52</td>
</tr>
<tr>
<td>5</td>
<td>exhaust steam</td>
<td>0.86</td>
<td>----</td>
<td>2.50</td>
<td>125</td>
<td>2.20</td>
<td>609.82</td>
<td>524.44</td>
</tr>
<tr>
<td>6</td>
<td>condensate</td>
<td>0.86</td>
<td>----</td>
<td>2.50</td>
<td>125</td>
<td>2.20</td>
<td>609.82</td>
<td>50.79</td>
</tr>
<tr>
<td>7</td>
<td>exhaust steam</td>
<td>2.48</td>
<td>----</td>
<td>2.50</td>
<td>125</td>
<td>2.20</td>
<td>609.82</td>
<td>1512.35</td>
</tr>
<tr>
<td>8</td>
<td>cane juice</td>
<td>21.45</td>
<td>50.00</td>
<td>----</td>
<td>110</td>
<td>2.93</td>
<td>29.94</td>
<td>642.21</td>
</tr>
<tr>
<td>9</td>
<td>cane juice</td>
<td>19.06</td>
<td>56.27</td>
<td>----</td>
<td>112</td>
<td>2.77</td>
<td>29.54</td>
<td>563.03</td>
</tr>
<tr>
<td>10</td>
<td>condensate</td>
<td>2.48</td>
<td>----</td>
<td>----</td>
<td>125</td>
<td>4.25</td>
<td>58.52</td>
<td>145.13</td>
</tr>
<tr>
<td>11</td>
<td>vapor</td>
<td>2.39</td>
<td>----</td>
<td>1.53</td>
<td>112</td>
<td>2.13</td>
<td>547.21</td>
<td>1307.83</td>
</tr>
<tr>
<td>12</td>
<td>vapor</td>
<td>1.83</td>
<td>----</td>
<td>0.70</td>
<td>90</td>
<td>2.04</td>
<td>431.66</td>
<td>789.93</td>
</tr>
<tr>
<td>13</td>
<td>cane juice</td>
<td>17.23</td>
<td>62.24</td>
<td>----</td>
<td>90</td>
<td>2.62</td>
<td>16.18</td>
<td>278.78</td>
</tr>
<tr>
<td>14</td>
<td>condensate</td>
<td>2.39</td>
<td>----</td>
<td>----</td>
<td>112</td>
<td>4.23</td>
<td>45.12</td>
<td>107.83</td>
</tr>
<tr>
<td>15</td>
<td>syrup</td>
<td>15.82</td>
<td>67.79</td>
<td>----</td>
<td>70</td>
<td>2.48</td>
<td>7.87</td>
<td>124.50</td>
</tr>
<tr>
<td>16</td>
<td>condensate</td>
<td>1.83</td>
<td>----</td>
<td>----</td>
<td>90</td>
<td>4.20</td>
<td>26.04</td>
<td>47.65</td>
</tr>
<tr>
<td>17</td>
<td>vapor</td>
<td>1.41</td>
<td>----</td>
<td>0.31</td>
<td>70</td>
<td>1.98</td>
<td>318.00</td>
<td>448.38</td>
</tr>
</tbody>
</table>

The calculated results for exergy analysis are shown in tables 7, 8 and 9. Table 7 shows the thermodynamic properties and exergy rates at different state points (refer fig. 5). The thermal properties pertaining to steam, vapor and condensate are estimated using a steam calculator. The highest exergy rate is found at state point 7 and the lowest exergy rate at state point 16. Table 8 for exergy analysis, shows the highest and lowest exergy destruction rate for II evaporator and pre-heater respectively. First evaporator has highest exergy efficiency of 89.86% and III
evaporator has lowest exergy efficiency of 39.62%. A detailed comparison of exergy analysis is shown in table 9. Total exergy destruction rate for conventional process is 3708.57 kW, whereas for the proposed process it is 1721.72 kW. The total exergy input for conventional and proposed process is 19449.21 kW and 13187.91 kW respectively. The value of TExDR for conventional process is 0.1906 and for proposed process it is 0.1305. The average exergy efficiency for conventional process is 70.53% whereas for proposed process it is 65.10%. The value of SI (sustainability index) for conventional process is 5.246 whereas for proposed process it is 7.662. The value of ERR for the proposed process is 0.3406, whereas for the conventional process it is 1.557.

**DISCUSSION**

A careful observation of table 6 shows that the value of ESD in terms of kg/s and % on cane for proposed process is lower than conventional process. Usually in sugar plants electricity is typically balanced with combined heat and power (CHP) or cogeneration plants. A reduction in ESD will ensure more steam being bled into the turbine which increases power generation. There is a major decrease in % evaporation rate from 75% to 26.24%. This is due to a large portion of water content removed from cane juice by using membrane distillation. This decreases the size and heating surface area required by the evaporators. Table 6 also shows that there is a marginal increase in electrical energy consumption where as the thermal or primary energy consumption decreased drastically from 72654.14 kW to 51438.48 kW. But the benefits of the proposed membrane process are principally evaluated using substitution co-efficient (CS). Results show a value of 4746.08 MJ/kWhr.

**Table 8: Results of Exergy analysis.**

<table>
<thead>
<tr>
<th>Component</th>
<th>$E_x_{\text{dest}}$ kW</th>
<th>$I_P$ kW</th>
<th>$\eta_{\text{ex}}$ %</th>
<th>TExDR</th>
</tr>
</thead>
<tbody>
<tr>
<td>Membrane</td>
<td>313.10</td>
<td>61.83</td>
<td>80.25</td>
<td>0.1305</td>
</tr>
<tr>
<td>Pre-heater</td>
<td>124.83</td>
<td>32.90</td>
<td>73.64</td>
<td></td>
</tr>
<tr>
<td>I Evaporator</td>
<td>139.58</td>
<td>14.14</td>
<td>89.86</td>
<td></td>
</tr>
<tr>
<td>II Evaporator</td>
<td>695.29</td>
<td>402.29</td>
<td>42.14</td>
<td></td>
</tr>
<tr>
<td>III Evaporator</td>
<td>448.92</td>
<td>271.05</td>
<td>39.62</td>
<td></td>
</tr>
<tr>
<td>Total/average</td>
<td>1721.72</td>
<td>782.21</td>
<td>65.10</td>
<td></td>
</tr>
</tbody>
</table>

**Table 9: Comparison of Exergy analysis.**

<table>
<thead>
<tr>
<th>Sl no.</th>
<th>Particulars</th>
<th>Conventional evaporation (S. M. Bapat et.al. 2012)</th>
<th>Proposed membrane evaporation</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>cold fluid</td>
<td>cane juice</td>
<td>cane juice</td>
</tr>
<tr>
<td>2</td>
<td>hot fluid (steam), T°C</td>
<td>125</td>
<td>125</td>
</tr>
<tr>
<td>3</td>
<td>feed flow sequence</td>
<td>feed forward</td>
<td>feed forward</td>
</tr>
<tr>
<td>4</td>
<td>vapor bleeding</td>
<td>Yes</td>
<td>yes</td>
</tr>
<tr>
<td>5</td>
<td>Reference state</td>
<td>25°C, 1.01325 bar</td>
<td>25°C, 1.01325 bar</td>
</tr>
<tr>
<td>6</td>
<td>Thermal Exergy consumption (kW)</td>
<td>16330.97</td>
<td>11562.18</td>
</tr>
<tr>
<td>7</td>
<td>Electrical exergy consumption (kW)</td>
<td>340.85</td>
<td>356.97</td>
</tr>
<tr>
<td>8</td>
<td>Total Exergy consumption (kW)</td>
<td>16671.82</td>
<td>11919.15</td>
</tr>
<tr>
<td>9</td>
<td>% Exergy savings</td>
<td>---</td>
<td>28.50</td>
</tr>
<tr>
<td>10</td>
<td>$\sum E_x_{\text{dest}}$ (kW)</td>
<td>3708.57</td>
<td>1721.72</td>
</tr>
</tbody>
</table>
Taking into account 1 kWhr of electrical energy available at the utilization site requires in a power station to burn 10.5 MJ of primary energy from a combustible (oil, gas, coal etc.), the substitution (or process innovation) results convenient when the CS is greater than 10.5 MJ/kWhr (R. Molinari et al., 1995). In the present context the value of CS is 4746.08 MJ/kWhr, which is quite acceptable. The result of a very high value of CS is due to a very marginal increase in electrical consumption which is in the denominator, and also phenomenal decrease in primary energy consumption which is in the numerator. Steam Economy (SE) is a very important factor for concentration/evaporation processes. The value of SE is 2.27, which is more than 1.99 for conventional process. Thus the benefits of the proposed process are well understood in terms of energy by CS and SE.

Table 8 shows that the III evaporator is the least efficient component followed by II evaporator in terms of exergy. The vapor entering III evaporator is at 90°C and the vapor generated is at 70°C. The vapor entering II evaporator is at 112°C and the vapor generated is at 90°C. This is mainly due to large value of ΔT. Table 9 shows that the values of exergy destruction rates, exergy inputs are less for proposed process as compared to conventional process. There is a rise in the value of Improvement potential (IP) for proposed process. The average exergy efficiency is marginally less as compared to conventional process. The value of ERR for the proposed process is very less as compared to conventional process. This is mainly due to drastic reduction in evaporation rates. Also it is to be noted that one of the useful product from the proposed process is the water content from the juice removed with the help of membrane process, which is definitely potable. This would definitely reward in terms of replenishment of water resources. The important thing to be noted is the value of TExDR and SI. SI is the reciprocal of TExDR. There is a fall in the value of TExDR and rise in the value of SI for the proposed process as compared to conventional process. This fact reveals the sustainability aspects of the proposed process in terms of exergy. Moreover the % in energy savings (29.04% as shown in table 6) and the % in exergy savings (28.50% as shown in table 9) are almost equal. Thus estimation of the convenience of using the proposed process in terms of sustainability, environmental impact, exergy and energy savings are well understood by factors like CS, SE, TExDR and SI.

**CONCLUSION**

The following conclusions can be drawn from the present study:

1. The cane juice evaporation rate for proposed membrane is 26.24% which is very less compared to 75% of evaporation in conventional process.
2. Since the evaporation rates are less the values of ESD for proposed process are quite attractive as compared to conventional process. This is highly beneficial in terms of energy consumption.
3. A high value of CS reveals that the proposed process saves lot of primary thermal energy compared to 1 kWhr of electrical energy consumed.
4. The proposed process is also beneficial in terms of SE.
5. The proposed membrane + evaporation technique is not attractive in phrases of common exergy performance, improvement ability and ERR in comparison to standard system.
6. Proposed procedure is useful in terms of TExDR or SI, which indicates its sustainability in terms of exergy.
7. The authors sense that usage of the proposed manner in present cane sugar industries could truly advantage in phrases of economic system, savings in energy and exergy consumption, sustainability, reduction in evaporation station place, substantially smaller carbon footprint.

REFERENCES


Nomenclature

B brix or amount of solid content in cane juice (%)

B_{in} brix content at the inlet of the evaporator (%)

B_{out} brix content at the outlet of the evaporator (%)

B_{m} average value of brix (%) 

C specific heat (kJ/kgK)

C_{pm} mean or average specific heat (kJ/kgK)

ex specific exergy (kJ/kg)

ex_{in} specific exergy of incompressible fluid (kJ/kg)

Ex total exergy (kW)

Ex_{in} total exergy at inlet (kW)

\sum E_{in,ex} total exergy content in the exhaust steam
**Ex**<sub>out</sub> total exergy at outlet (kW)

**Ex**<sub>dest</sub> total exergy destruction rate (kW)

**Ex**<sub>steam</sub> total exergy content in steam (kW)

**Ex**<sub>product.in</sub> total exergy content in the product at inlet (kW)

**Ex**<sub>product.out</sub> total exergy content in the product at outlet (kW)

**Ex**<sub>vap</sub> total exergy content in vapor (kW)

**Ex**<sub>cond</sub> total exergy content in the condensate leaving the evaporator (kW)

**Ex**<sub>pump</sub> total exergy input to the pump (kW)

**Ex**<sub>Tot,dest</sub> sum of total exergy destruction (kW)

**Ex**<sub>Tot,in</sub> sum of total exergy input (kW)

**Ex**<sub>hot,in</sub> total exergy of hot fluid at inlet (kW)

**Ex**<sub>hot,out</sub>total exergy of hot fluid at outlet (kW)

**Ex**<sub>cold,in</sub> total exergy of cold fluid at inlet (kW)

**Ex**<sub>cold,out</sub> total exergy of cold fluid at outlet (kW)

**ΔEx**<sub>cold</sub> rise in total exergy content of cold fluid (kW)

**ΔEx**<sub>hot</sub> fall in total exergy content of hot fluid (kW)

**F**<sub>i</sub> juice flow rate at the inlet of the evaporator i (kg/s)

**g** acceleration due to gravity (m/sec<sup>2</sup>)

**H** head (meters)

**h** specific enthalpy (kJ/kg)

**h**<sub>o</sub> specific enthalpy at reference condition (kJ/kg)

**h**<sub>fg,v</sub> latent heat of vaporization of vapor generated in the evaporator (kJ/kg)

**h**<sub>fg,s</sub> latent heat of vaporization of steam supplied to the evaporator (kJ/kg)

**m** mass flow rate (kg/s)

**m**<sub>s</sub> mass flow rate of steam (kg/s)

**m**<sub>v,i</sub> mass flow rate of vapor in the i<sup>th</sup> effect (kg/s)

**p** pressure (bar)

**Q** volumetric discharge (m<sup>3</sup>/sec)

**s** specific entropy (kJ/kgK)

**s**<sub>0</sub> specific entropy at reference condition (kJ/kgK)

**T** temperature (K)

**T**<sub>o</sub> reference temperature (K)

**ΔT**<sub>j</sub> temperature difference between the juice streams entering and leaving the evaporator i (°C)

**V**<sub>i</sub> mass flow rate of the vapor generated in the evaporator i (kg/s)

**Suffixes:**

CLARIF clarified

in inlet

out outlet

j juice

dest destruction

mem membrane

ph preheater

I first evaporator

II second evaporator

III third evaporator

**Greek symbols:**

**ρ** mass density (kg/m<sup>3</sup>)

**η** efficiency (%)

**η**<sub>ex</sub> exergy efficiency (%)