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Potentials for Energy Saving in
Cane Sugar Industry

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Abstract

This work investigates the possibilities of saving energy (heat and electrical) in the cane sugar factories in order to save bagasse for co-generation. Co-generation is a recent trend in the cane sugar industry, which can generate huge income for the sugar factory.

Many possibilities for energy saving have been considered in this work. Emphasis was made on two fields, namely the application of pressure evaporation, the configuration of heat circuit with intensive vapor bleeding, stage-wise heating of the juice and utilization of condensate.

Three cases were considered with the current heat circuit of Alguneid Sugar Factory as a base case. The configuration implies the use of steam at 124°C and pressure 2.2 bar with vapor bleeding from all effects, the condensate is utilized for juice heaters, with the 2\textsuperscript{nd} vapor bleeding for the vacuum pans that has resulted in a lowest steam consumption of 23.988 kg/100kg cane with a difference of 6.863 kg/100kg cane compared to the base case. This configuration requires no addition of new equipment so it can be applied without incurring any extra costs.

According to the investigation carried out in this work, it is possible to save an amount of bagasse in the range of 2.6 kg/100kg cane. Applying these calculations to the data obtained from Alguneid Sugar Factory, it is possible to generate an extra amount of electricity of 0.55 kWh/100kg cane which if it is sold will produce annual revenue of 106,260,000 SD, without incurring any extra costs for new equipment.
ملخص الدراسة

يتناول هذا البحث إمكانيات ترشيد الطاقة (الحرارية والكهربائية) في مصنع لإنتاج السكر من القصب لكي يتولى الباقاس من أجل توليد الكهرباء. بعد توليد الكهرباء اتجاهًا حديثًا في صناعة السكر القصب حيث يمكن أن يدر دخلا كبيرا لمصنع السكر.

هناك عدد من الخيارات الممكنة لتوفير الطاقة قد أخذت بعين الاعتبار في هذا العمل، وقد تم التركيز بصورة أساسية على مجالين هما: استخدام بخار عالي الضغط في التبخير وتشكيل منظومة التوزيع الحراري مع الاستفادة من البخار المستنزف في تxin العصير على مراحل المصنع باستخدام المكتاف المتزامن المتبخرّ.

تمت دراسة ثلاث حالات حيث أخذ التوزيع الحراري الحالي لمصنع سكر الجديد كحالة مرجعية.

من بين تلك الحالات فإن الحالة التي استخدم فيها البخار عند حرارة 124°ق وضغط 2.2 بار مع استنزاف البخار من كل المبخرات و أيضا استخدام الماء المتكافف لسخانات العصير و البخار المستنزف من المبخر الثاني للمثلورات تحت الفراغ. قد نتج عن هذه التشكيلة استهلاك أقل للبخار ما يعادل 23.988 كغ/100 كغ قصب مع فارق 6.863 كغ/100 كغ قصب مقارنة مع الحالة المرجعية. هذه التشكيلة لا تتطلب أية إضافية لمعدات جديدة لذلك يمكن تطبيقها دون تحمل أية تكاليف إضافية. وفقًا للدراسة التي تم أجراؤها في إطار هذا العمل فأن الممكن توفر كمية من الباقاس تصل إلى 2.6 كغ/100 كغ قصب. يطبق هذه الحسابات على البيانات الخاصة بمصنع سكر الجنيد فإنه يمكن إنتاج كمية من الكهرباء قدرها 0.55 كيلو واط ساعة /100 كغ قصب و التي إذا ما بيعت سوف تدر دخلا سنويا يعادل 106,260,000 دينار سوداني دون تحمل أية تكاليف إضافية لتجهيزات جديدة.
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Abbreviations

FAO : Food and Agriculture Organization
HP : High pressure
LP : Low pressure
EU : European Union
SADC : Southern African Development Community
CEST : High-pressure condensing-extraction steam turbine system
GSTIG : Steam-injected gas turbines run on gasified bagasse
FFE : Falling film evaporator
DC : Direct current; direct coupled
Chapter I
Introduction

1.1 General

Considering the “green energy” potential, just a few countries have identified bagasse as a source of electric power production and created the economical framework to make excess power production for cane sugar factories. Worldwide more than 1 Billion tons of cane are harvested annually, which produce more than 300 Million tons of bagasse, mostly used as fuel in cane sugar factories. FAO data show that about 1248 Million tons of cane were produced in 1997, which offered 312 Million tons of bagasse. The energy content per one ton bagasse (50% moisture content) is 2.85 GJ. This excludes top, leaves and trash - representing the largest energy fraction of the sugar cane (55%) - which is currently mostly burned off or left to rot in the fields. This large potential is thus currently almost entirely wasted [1]. Table (1.1) shows the global market potential of bagasse available in the country in 2004 [2].

Sugar producers have been using bagasse to raise steam for on-site processes for centuries, but very inefficiently. However, recently economic pressures have forced many sugar cane factories to look for alternatives and to achieve self-sufficiency in energy. Interest in co-generation has increased considerably in many sugar cane producing countries, of which Brazil, India, Thailand and Mauritius are good examples [1]. The average power production of cane factories lies in the range of 10 – 20 kWh/t cane and often covers all self-consumption. The improvement of the process steam economy, from present values of more than 50 % steam demand on cane (o.c) to the range of 20 - 30 % (o.c), offers the prospect to increase the power production significantly. Only some of the cane sugar
factories have achieved today a process steam demand between 30 – 35 % (o.c) and there is still great potential in reduction of energy consumption figures. There is no doubt, that there will be a great challenge of adoption of certain measures to achieve this goal; it is only a matter of time when this will happen [3].

Table (1.1): Global market potential of bagasse.

<table>
<thead>
<tr>
<th>Country</th>
<th>Sugar cane production ton/yr</th>
<th>Potential for electricity production GWh/yr</th>
<th>Bagasse potential as percentage of electricity demand</th>
</tr>
</thead>
<tbody>
<tr>
<td>Brazil</td>
<td>386,232,000</td>
<td>38,623</td>
<td>11.50</td>
</tr>
<tr>
<td>India</td>
<td>290,000,000</td>
<td>29,000</td>
<td>5.83</td>
</tr>
<tr>
<td>China</td>
<td>93,900,000</td>
<td>9,390</td>
<td>0.72</td>
</tr>
<tr>
<td>Thailand</td>
<td>74,071,952</td>
<td>7,407</td>
<td>6.15</td>
</tr>
<tr>
<td>Pakistan</td>
<td>52,055,800</td>
<td>5,206</td>
<td>8.36</td>
</tr>
<tr>
<td>Mexico</td>
<td>45,126,500</td>
<td>4,513</td>
<td>2.42</td>
</tr>
<tr>
<td>Colombia</td>
<td>36,600,000</td>
<td>3,660</td>
<td>9.19</td>
</tr>
<tr>
<td>Australia</td>
<td>36,012,000</td>
<td>3,601</td>
<td>1.95</td>
</tr>
<tr>
<td>Cuba</td>
<td>34,700,000</td>
<td>3,470</td>
<td>25.93</td>
</tr>
<tr>
<td>USA</td>
<td>31,178,130</td>
<td>3,118</td>
<td>0.09</td>
</tr>
<tr>
<td>Philippines</td>
<td>25,835,000</td>
<td>2,584</td>
<td>6.16</td>
</tr>
<tr>
<td>Other</td>
<td>244,581,738</td>
<td>24,458</td>
<td>0.32</td>
</tr>
<tr>
<td>Total</td>
<td>1,350,293,120</td>
<td>135,029</td>
<td>0.97</td>
</tr>
</tbody>
</table>

Two key goals in energy optimization are to maximize high-pressure (HP) steam production and to minimize the requirements for low-pressure (LP) steam. Reducing energy requirements may involve a variety of steps such as:
1. The complete electrification of the sugar factory (i.e. replacement of mill turbines with motors driven by electricity).

2. Optimum heat circuit within the process area.

3. Reducing energy demands of the final evaporator stage by increasing the number of evaporator stages, or by using vapor recompression systems, or by increasing the temperature of the vapor coming from the final evaporator so it can be used elsewhere in the factory (e.g. heaters and pans).

4. Using continuous operations instead of batch ones (e.g. crystallization) [4].

One possibility for saving energy is the adoption of pressure evaporation, which is the standard practice in the beet sugar industry; the conventional pressure-vacuum evaporation prevails in the cane sugar industry. Pressure evaporation leads to energy saving because it offers steam of high quality for the various heating proposes in the factory instead of exhaust steam. Pressure evaporation brings about certain adverse effects, such as sucrose destruction (hydrolysis), intensifying color formation, and increasing rate of encrustation (scale formation). This is because of the nature of the technical sugar solution in the cane sugar industry, which is chemically unstable due to the presence of invert sugars in it.

It was planned to investigate the ill effects of pressure evaporation on sugar production experimentally, but due to the difficulties encountered in conducting the experimental work, this part of the research was replaced with computer simulation program. The program was written in FORTRAN language for calculation of the amount of steam consumed for juice heating and concentration. First a mathematical model for the calculation of the evaporator station was developed. This model is used as a basis for writing a simulation program for the calculation of the amount of steam consumed in the evaporators. Influence of temperature variation
1.2 Objectives

General objective:
The aim of this work is to investigate the possibilities of saving energy in the cane sugar industry by adopting certain measures in energy generation, distribution and utilization in the factory.

Specific objectives:

1. To investigate the means of saving energy in the cane sugar factory.
2. To investigate the suitability of pressure evaporation for the cane sugar industry and its effects on steam required in evaporator station.
3. To determine the optimum operation conditions in the evaporator station.
4. To develop an energy distribution circuit, which will lead to energy saving by reducing process steam usage.
5. To evaluate an energy consumption of Alguneid Sugar Factory.
6. To recommend measures regarding energy generation, distribution, and utilization.
Chapter II

Literature Review

2.1 Cane sugar production

2.1.1 International production

The sugar industry processes sugar cane and sugar beet to produce commercial sugar. Sucrose ($C_{12}H_{22}O_{11}$) which is a disaccharide is made up of two molecules, glucose and fructose. More than 100 countries around the world produce sugar, approximately 75% of which is produced from sugar cane grown primarily in the tropical and sub-tropical zones of the southern hemisphere. The balance is produced from sugar beet which, is grown in the temperate zones of the northern hemisphere. Prior to 1990, about 40% of sugar was made from beet, but this has decreased to current levels as cane sugar producers have made considerable gains in expanding their sugar markets due to the lower costs of cane sugar production. About 71% of production is consumed in the country of origin whilst the balance is traded on world markets [5].

Sugar manufacturing is a highly seasonal industry, with season lengths of about 6 to 18 weeks for beet and 20 to 32 weeks for cane. A comparison between sugar cane and sugar beet is shown in table (2.1)[6].

Table (2.1): Comparison between sugar cane and sugar beet properties.

<table>
<thead>
<tr>
<th>Crop</th>
<th>Water</th>
<th>Sucrose</th>
<th>Fiber or pulp</th>
<th>Soluble impurities</th>
<th>Sugar yield ton/ha</th>
<th>Season length weeks</th>
<th>Crop yield ton/ha</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sugar cane</td>
<td>70%</td>
<td>10-15%</td>
<td>14%</td>
<td>2.7%</td>
<td>5-13 t/ha</td>
<td>20 - 32</td>
<td>50-150 t/ha</td>
</tr>
<tr>
<td>Sugar beet</td>
<td>75%</td>
<td>15-19%</td>
<td>7%</td>
<td>2.5%</td>
<td>7-10 t/ha</td>
<td>6 - 18</td>
<td>45-70 t/ha</td>
</tr>
</tbody>
</table>
Total world sugar production in the 2003/04 season was about 141 Million tons, compared to 128 Million tons in 2000/01, with consumption of 143 Million tons (130 Million: 2000/01). The five largest exporters, being Brazil, the EU, Thailand, Australia and Cuba, supplying approximately 71% of all world free market exports (table 2.2) [5].

Table (2.2): Top ten sugar producers.

<table>
<thead>
<tr>
<th>2002/03</th>
<th>Production million tons</th>
<th>Export million tons</th>
<th>Population million</th>
<th>Per capita Consumption kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>Brazil</td>
<td>22,703</td>
<td>12,445</td>
<td>170</td>
<td>55</td>
</tr>
<tr>
<td>India</td>
<td>19,457</td>
<td>1,630</td>
<td>1,043</td>
<td>18</td>
</tr>
<tr>
<td>EU</td>
<td>18,341</td>
<td>5,786</td>
<td>380</td>
<td>37</td>
</tr>
<tr>
<td>China</td>
<td>9,783</td>
<td>109</td>
<td>1,294</td>
<td>7</td>
</tr>
<tr>
<td>USA</td>
<td>7,425</td>
<td>113</td>
<td>280</td>
<td>31</td>
</tr>
<tr>
<td>Thailand</td>
<td>6,895</td>
<td>4,616</td>
<td>64</td>
<td>36</td>
</tr>
<tr>
<td>Australia</td>
<td>5,569</td>
<td>4,537</td>
<td>20</td>
<td>46</td>
</tr>
<tr>
<td>SADC</td>
<td>5,467</td>
<td>1,682</td>
<td>155</td>
<td>21</td>
</tr>
<tr>
<td>Mexico</td>
<td>5,062</td>
<td>272</td>
<td>104</td>
<td>46</td>
</tr>
<tr>
<td>Cuba</td>
<td>2,400</td>
<td>1,804</td>
<td>11</td>
<td>51</td>
</tr>
</tbody>
</table>

2.1.2 Local production

There are five sugar factories in Sudan producing sugar from sugar cane. Their production capacity is 700,000 tons annually. The Sudanese Sugar Co. LTD administrates the four public sugar factories:

The fifth factory is Kenana Sugar Company, a share-holding company between the Government of Sudan and some Arab countries. The bulk of the investment was $ 670,000,000. The factory includes irrigated sugar cane farms, an industrial complex, and a refinery. It processes 17,000 tons of sugar cane per day. The design capacity of white sugar is 300,000 tons, and in 1999 the production reached 370,000 tons. The factory exports bring 7 Million dollars annually to Sudan. It produces, in addition to the white sugar, sugar cubes, cane honey, molasses, and fodder. There is a plan for the production of char-coal from bagasse [7]. Table (2.3) presents the production of sugar in the five factories during the production seasons 1994/1995-1997/1998 [7].


<table>
<thead>
<tr>
<th>Year</th>
<th>Alguneid</th>
<th>Halfa</th>
<th>Sinnar</th>
<th>Asalaya</th>
<th>Kenana</th>
<th>Total</th>
</tr>
</thead>
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<tr>
<td>94-95</td>
<td>44124</td>
<td>57227</td>
<td>42244</td>
<td>31555</td>
<td>253000</td>
<td>428150</td>
</tr>
<tr>
<td>95-96</td>
<td>43524</td>
<td>55269</td>
<td>44650</td>
<td>35658</td>
<td>280000</td>
<td>438329</td>
</tr>
<tr>
<td>96-97</td>
<td>46000</td>
<td>52100</td>
<td>36000</td>
<td>33400</td>
<td>333000</td>
<td>500500</td>
</tr>
<tr>
<td>97-98</td>
<td>58567</td>
<td>61615</td>
<td>43005</td>
<td>43023</td>
<td>356000</td>
<td>557224</td>
</tr>
</tbody>
</table>
2.2 Cane sugar manufacturing

2.2.1 International method for the production of sugar from sugar cane

1- harvesting

Sugar cane is harvested by chopping down the stems, but leaving the roots, so that it re-grow with time for the next crop (rationing). Harvest times tend to be during the dry season and the length of the harvest period ranges from as little as 2 ½ months up to 11 months. Harvesting is done either by hand or by machine. Typically, cane is processed within 24 hours after cutting. The cane is burned in the field after thrust for about 2-3 weeks to release the leaves and fibers. Then it is cut for specific length (about 30 cm); this helps preparation later in the factory. The cutting is achieved either manually or mechanically.

2- Cane cleaning

Cane is usually washed with water in order to remove the soil particles and attached ash, and other impurities that may damage the mills and hindered the sugar gaining process. Dry cleaning by air is another way which can be used to remove the soil accompanying the cane from the field, some small fibers, and the green trash from the plant. Also there is a big magnet to attract metallic materials and protect the mills.

3- Preparation

The washed cane is then carried by a belt to be prepared for grinding by revolving set of knifes (driven by steam turbine using live steam or electricity) that cut the stalks into chips. In the modern sugar factories in addition to knifes, shredders are used to improve cane preparation. The shredders have heavy hummers, and are generally driven by steam turbine
and seldom by electric motor because of the high energy consumption. Shredders are used to open the cells and make the juice discharge from them easily but extract no juice [8].

4- Extraction of juice

The conventional method of extraction of juice from sugar cane is either through a train of three-roller mill or a diffuser as is in beet sugar manufacturing. Both systems are working in the cane sugar industry. Juice extraction by crushing the cane between massive rollers is the first step in raw sugar processing. Mills consist of multiple units of three-roller combinations through which the crushed cane or bagasse passes successively. To aid in the extraction of the juice, sprays of water or thin juice are directed on the blanket of bagasse as it emerges from each mill unit help to leach out the sugar. The process termed imbibition and has many modifications. In best milling practice, more than 95% of the sugar in the cane goes into the juice [8]. A typical mixed juice from extraction will contain approximately 15% sugar and the residual fiber after juice extraction is called bagasse. A typical cane might contain 12 to 14% fiber that, gives about 25 to 30 tons of bagasse of 50% moisture content per 100 tons of cane [8]. This material usually goes to the boilers as fuel for energy generation, or other commercial by-product utilization.

5- Clarification of juice

The dark-green juice from the mills is acidic (pH 4.8 - 6) and turbid. The clarification process is designed to remove both soluble and insoluble impurities and universally employs lime and heat as the clarifying agents. Here lime and phosphoric acid are added to the juice, and surface active materials (polymer flocculants, coagulants) are also added to aid the
purification process. Milk of lime, about 1lb (0.5kg) CaO per ton of cane, neutralizes the natural acidity of the juice, forming insoluble lime salts, mostly calcium phosphate. Heating the limed juice to boiling or slightly above coagulates the albumin and some of the fats, waxes, and gums. The muds are separated from the clear juice by sedimentation. The muds are filtered on rotary drum vacuum filters. The filtered juice returns to the process or goes directly to clarified juice, and the press cake is discarded or used in the fields as fertilizer [8].

6- Evaporation

The clarified juice, having much the same composition as the raw extracted juice except for the precipitated impurities removed by the lime treatment, contains about 85% water. Two-thirds of this water is removed in the multiple effects evaporators consisting of a succession of set of “bodies” arranged in series so that each succeeding body has a lower pressure, and therefore heats at a lower temperature. The vapor from one body is used to heat the juice in the next one. By this system the steam introduced into the first body does multiple-effect evaporation. The vapor from the final body goes into a condenser. The syrup leaves the last body continuously with about 65% solids and 35% water [8].

7- Crystallization

Crystallization takes place in single-effect vacuum pans, where the syrup is evaporated until saturated with sugar. At this point seed grain is added to serve as nuclei for the sugar crystals, and more syrup is added as the water evaporates. The growth of the crystals continues until the pan is full. Given a skilled sugar boiler, the original crystals can be grown without the formation of additional crystals, so that when the pan is just full, the
crystals are all of the desired size, and the crystals and syrup form a dense mass known as massecuite. The strike is then discharged through a foot valve into a mixer or cooling crystallizer. There are many crystallization systems, one product, two products, and three products…etc. In the three crystallization system, which is shown in the flow diagram, the first boiling of raw syrup yields raw sugar and A molasses, which is returned to the vacuum pan to be reboiled on a footing of first-grade massecuite to a second massecuite B that in turn yields a second crop of crystals. The B sugar is mixed with the A sugar to form the commercial output of the factory. The second or B molasses is of much lower purity and in turn is reboiled on a footing of syrup crystals to form low-grade or C massecuite. These low-grade massecuies remain in cooling crystallizers for several days, where they cool while being kept in motion with stirring arms. The C sugar is mingled with syrup and used for A and B massecuite seed [8].

8-Centrifugation
The massecuites from mixer or crystallizer is drawing into revolving machines called centrifuges. The cylindrical basket suspended on a spindle has perforated sides lined with wire cloth, inside of which are metal sheets containing 400-600 perforations per square inch. The basket revolves at speeds from 1000 to 1800 rpm. The perforated lining retains the sugar crystals, which may be washed with water if desired. The mother liquor, molasses, passes through the lining, and after the sugar is purged it is cut down, leaving the centrifuge ready for another charge of massecuite. The final molasses, or blackstrap, a heavy, viscous material containing approximately 1-2% sucrose, 1.5 reducing sugars, and the remainder ash, organic non-sugars, and water, serves as a base for cattle feed, in the manufacture of industrial alcohol, in yeast production, and so on[8]. Fig (2.1) is a flow diagram of a cane sugar factory [9]. Fig (2.2) is a flow sheet for the production of sugar from sugar cane.
2.2.2 Production of refined sugar
There are three basic steps in the refining of raw sugar. These are (1) washing of sugar crystals, dissolving, (2) decolorization of the sugar, and (3) recrystallization and finishing.

1- Affination or sugar washing
The purpose of the affination process is to remove the film of molasses from the surface of the raw sugar. The sugar crystals are essentially pure sucrose. This is done by mixing high density syrup with the raw sugar, then purging the mixture in centrifuges, after the removal of the syrup, the crystals are washed with hot water to effect further purification. The washed crystals are dissolved in about half their mass of water in dissolvers. Hot water is used in order to accelerate dissolving of the sugar. The next step taken with the sugar solution from the dissolvers is to remove insoluble material, which has accumulated during the previous steps. The next step is the treatment of the sugar solution with certain substances and then heating to prepare it for filtration or clarification. The most common materials used are phosphoric acid and lime or carbon dioxide and lime followed by filtration. Another method is the flotation process [10].

2- Decolorization
The next step in sugar purification is decolorization of the solution. This is an important step because the presence of any colorant in the sucrose is highly objectionable for the use of the product in food manufacturing. Colorants may originate from plant pigments, which have remained in the sugar solution. Or colorants are due to the reaction of amino acids and reducing sugars (Millard reaction), or from materials, which have been derived from thermal decomposition of sucrose or of reducing sugars. Carbon products are the most commonly used adsorbents for the removal
of colorants in sugar solutions within the refinery. These carbons are regenerated at intervals in order to maintain their effectiveness. In addition to the use of carbon, ion-exchange resins are becoming more widely applied in sugar refining.

3- Recrystallization and finishing

The third basic step in the manufacture of sugar is recrystallization and finishing. The sugar solution of 65% to 68% solids from the decolorizing process is fed into vacuum pans for further removal of water and crystallizing of the sucrose. The most widely used vacuum pan is the calandria type, which utilizes low-pressure steam. After the crystals are formed, they are washed and then transferred to centrifuges for removal of all free liquor. These moist crystals contain about 1% moisture upon being discharged from the centrifuges, are conveyed to holding bins located above dryers. These dryers are called granulators because they rotate and keep the crystals from adhering during the drying cycle. After drying, the sugar crystals are classified according to size; an experienced operator can hold the size distribution relatively constant. In such limited situations where very close control on crystal size is required, screen classifiers are used. The final handling step is to store the granulated sugar in silos, bins, or bags [10]. Fig (2.3) shows a flow sheet for refined sugar production.
2.2.3 Sudanese cane sugar production

2.2.3.1 Process description of Alguneid Cane Sugar Factory

Alguneid Sugar Factory is located about 135 km south of Khartoum on the east side of the Blue Nile. Production started in 1962/1963 with a design capacity of 4800 ton cane/day. The cane is mechanically harvested; its properties are shown in table (2.4) [11]. Then it is transported to the mill using trailers.

Table (2.4): The properties of sugar cane in Alguneid.

<table>
<thead>
<tr>
<th>Crushing season</th>
<th>Average yield</th>
<th>Pol</th>
<th>Fiber</th>
<th>Trash</th>
<th>Brix</th>
<th>Moisture</th>
<th>Recovery</th>
</tr>
</thead>
<tbody>
<tr>
<td>220 Days</td>
<td>113.83 t/ha</td>
<td>13%</td>
<td>17%</td>
<td>3.29%</td>
<td>16%</td>
<td>67%</td>
<td>10.08%</td>
</tr>
</tbody>
</table>

The cane is received at the mill (203t/h) and prepared for extraction of the juice. At first, the cane is weighed then mechanically unloaded, placed in a large moving table and, is washed with sprayed clean water (400-500 t/h). From there the cane continuously travels through a system of conveyors to the cane preparation unit. This process occurs by breaking the hard structure of the cane into small pieces using two revolving knives driven by high-voltage electricity (6600 kV). For the milling of the crushed cane, multiple sets of three-roller mills, driven by steam turbine using live steam (360ºC, 19.61bar), consisting of six mills tandem, are used. Conveyors transport the crushed cane from one mill to the next. Imbibition is the process in which water is applied to the crushed cane to enhance the extraction of the juice. In imbibition the water, mainly condensate, is introduced into the last two mills (25-29% cane) and transferred from mill to mill towards the first two mills while the crushed cane travels from the first to the last mill. The crushed cane exiting from the last mill is called
bagasse. This bagasse is primarily used as a fuel in the boilers for the generation of process steam. The juice taken from the first two mills, called mixed juice (86% cane), is then clarified. Clarification is done with heating and liming using milk of lime (0.081%cane); small quantities of phosphoric acid are also added (0.010%cane). The lime is added at 30°C, and the temperature of the juice is raised to about 105°C in a series of five heaters. A muddy precipitate forms which is separated from the juice in two clarifiers at 103-105°C [11].

Clear juice is separated from the mud in a rotary vacuum filter; the filter cake is used as fertilizer. The clarified juice is passed through two heat exchangers to raise its temperature to 115°C and then goes to the evaporator stations without additional treatment. Evaporation is performed in an evaporator station to concentrate the juice from 16 to 60% by mass. Evaporator stations consist of four bodies in series, termed quadruple-effect evaporators. Exhaust steam from mills, turbo-alternators and make-up is used to heat the first evaporator, and the vapor formed in the first evaporator is used to heat the second evaporator. This heat transfer process continues through the four evaporators and as the temperature decreases from evaporator to evaporator, the pressure inside each evaporator also decreases. This allows the juice to boil at lower temperatures in the subsequent evaporator. The vapor that formed in the first three evaporators is used for various heating purposes in the factory. The evaporator station in cane sugar manufacture typically produces syrup with about 65 % solids and 35 % water. The syrup goes to the vacuum pans for crystallization. Crystallization of the sugar starts in the vacuum pans, whose function is to produce sugar crystals from the syrup. In the pan boiling process, the syrup is evaporated until it reaches the super-saturation stage. At this point, the crystallization process is initiated by “seeding”. When the volume of the mixture of liquor and crystals, known as massecuite, reaches the desired
capacity of the pan, the contents of the vacuum pans are discharged into cooling crystallizer, whose function is to increase the size of sugar crystal and further exhaust the massecuite of sugar. In the factory 3-scheme Crystallization system is applied; there are (4 pans) for A massecuite. From the crystallizer, the massecuite A massecuite is transferred to high-speed centrifugal machines (3 batch centrifugals for A massecuite), in which the mother liquor A heavy molasses is centrifuged to the outer shell and the crystals A sugar remain in the inner centrifugal basket [11]. The crystals are washed with water and the wash water centrifuged from the crystals A light molasses. A light molasses goes to C pans. The liquor A heavy molasses from the A centrifuge is returned to B vacuum pans (2 pans) and reboiled to yield a second massecuite B massecuite, that in turn yields a second batch of crystals. The B massecuite is transferred to the crystallizer and then to the B centrifuge (5 continues), and the B sugar is separated from the molasses. This B sugar is mixed with water to make the magma, which goes to A pans to serve as a base for this strike. The molasses from the second boiling B molasses is of much lower purity than the first molasses. It is reboiled with A light molasses in C pans (3 pans) to form a low-grade massecuite C massecuite, which goes to a crystallizer and then to a C centrifuge (4 continues). This low-grade cane sugar is mingled with water to make the dissolved C sugar and is used in the A vacuum pans. The final molasses from the third stage (blackstrap molasses) is a heavy, viscous material used primarily as a supplement in cattle feed. The cane sugar from the A massecuite is dried in rotary driers and cooled. After cooling, the cane sugar is transferred to packing bins and then sent to storage [11]. Specifications of final product and by-products are shown in table (2.5) [11].
Table (2.5): Specifications of final product and by-products in Alguneid Sugar Factory.

<table>
<thead>
<tr>
<th>Analysis</th>
<th>% Cane</th>
<th>Brix %</th>
<th>Pol %</th>
<th>Purity %</th>
<th>Moisture%</th>
<th>Fiber %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sugar</td>
<td>10.08</td>
<td>99.88</td>
<td>97.80</td>
<td>97.91</td>
<td>0.115</td>
<td></td>
</tr>
<tr>
<td>Final molasse</td>
<td>3.49</td>
<td>90.40</td>
<td>34.76</td>
<td>38.45</td>
<td>9.60</td>
<td></td>
</tr>
<tr>
<td>Bagasse</td>
<td>43.19</td>
<td>4.64</td>
<td>3.63</td>
<td>78.23</td>
<td>50.80</td>
<td>44.56</td>
</tr>
<tr>
<td>Filter cake</td>
<td>3.00</td>
<td>1.25</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

2.2.3.2 Energy in the factory

1- Boiler station
There are four boilers producing live steam with a total design capacity of 100 t/h, the actual capacity is 90 t/h at 360ºC and 19.61 bar. Bagasse is used as fuel. Feed water consists of condensates of the process steam and treated water make up (total feed water flow 94 t/h at 100ºC).

2- Power house
There are two turbo-alternators; with a total design power production of 6 MWh, the actual power production is 4.5-5 MWh, using live steam (total amount 44 t/h), and produce exhaust steam at 124ºC and 1.18 bar. A flow sheet of Alguneid Cane Sugar Factory is shown in Fig (2.4).
3.1 Steam generation
3.1.1 Co-generation
Today, co-generation is considered to be the simultaneous production of power and heat, usually in the form of electricity and steam. Co-generation requires considerable additional investment. This additional investment may be rewarding, and co-generation will be attractive, when:

- Fuel is available at low cost; this arises in situation such as sugar factories, where bagasse is used as fuels. High-pressure steam is usually raised in a boiler and passed through a turbine to generate power. The low pressure turbine, that exhausts steam, is then available for process heating.

- Co-generation schemes need to be tailored to the particular user demands and considerable professional expertise is needed to understand and meet all requirements in a satisfactory way [12].

3.1.2 Bagasse properties
Bagasse is the fibrous residue of sugar cane remaining after the extraction of juice. It is well suited for use as a fuel, having adequate calorific value, is low in ash and virtually free of environmental contaminants such as sulfur. The fiber is composed of the cellulose-ligninpentosan complex common to the structural matter of plants. The fiber carries with it the absorbed juice and water not removed in the de-watering step together with a mixed product of disintegrated cane trash along with a small quantity of inorganic material termed ash [12]. Typical properties of bagasse are outlined in table (3.1) [8].
Table (3.1): Typical properties of bagasse.

<table>
<thead>
<tr>
<th>Water content</th>
<th>46 – 52 %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fiber content</td>
<td>43 – 52 %</td>
</tr>
<tr>
<td>Soluble solids</td>
<td>2 – 6 %</td>
</tr>
<tr>
<td>Average density</td>
<td>150 kg/m³</td>
</tr>
</tbody>
</table>

3.1.3 Use of bagasse as a fuel

Bagasse is burnt in suitable boilers to produce steam which in turn is transformed via prime movers, for example turbines, to mechanical or electrical energy. Exhaust steam from the turbines is used as a source of thermal energy, for example, in the heating cycle of the cane juice within a sugar factory. The ‘as fired’ analyses of typical bagasse, a medium volatile coal, fuel oil and pine wood are presented in table (3.2) [13]

Table (3.2): Analyses of typical fuels as fired.

<table>
<thead>
<tr>
<th>Constituents</th>
<th>Bagasse %</th>
<th>Coal %</th>
<th>Oil %</th>
<th>Wood %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon</td>
<td>22.9</td>
<td>70.3</td>
<td>85.8</td>
<td>26.7</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>2.8</td>
<td>4.0</td>
<td>10.8</td>
<td>2.8</td>
</tr>
<tr>
<td>Sulphur</td>
<td>0.0</td>
<td>0.5</td>
<td>2.8</td>
<td>0.0</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.0</td>
<td>1.8</td>
<td>0.2</td>
<td>0.1</td>
</tr>
<tr>
<td>Oxygen</td>
<td>21.3</td>
<td>6.0</td>
<td>0.0</td>
<td>18.9</td>
</tr>
<tr>
<td>Moisture</td>
<td>50.0</td>
<td>6.0</td>
<td>0.4</td>
<td>50.0</td>
</tr>
<tr>
<td>Ash</td>
<td>3.0</td>
<td>11.4</td>
<td>0.0</td>
<td>1.5</td>
</tr>
<tr>
<td>GCV kJ/kg</td>
<td>9,177</td>
<td>28,410</td>
<td>43,000</td>
<td>7,936</td>
</tr>
<tr>
<td>NCV kJ/kg</td>
<td>7,409</td>
<td>27,385</td>
<td>40,887</td>
<td>7,936</td>
</tr>
<tr>
<td>Stoic.CO₂ %</td>
<td>20.7</td>
<td>18.8</td>
<td>16.1</td>
<td>20.6</td>
</tr>
</tbody>
</table>
For bagasse fuel, it is a usual practice to burn the fuel in a boiler to produce high pressure steam, which is fed into the turbine to generate mechanical and electrical power, and the exhaust steam leaving the turbines is used for heating in the sugar factory.

This method allows the power turbo machinery to work in a clean steam environment, keeping the fuel and ash problems confined to the boiler stack systems. Modern boilers are highly developed and well understood. Usually it is possible to capture more than 85% of the lower heating value of the fuel as high pressure steam energy.

Depending on the scale of operations, steam conditions, and turbine details, some 15-25% of the HP steam energy can be extracted as power. The balance is available in the LP turbine exhaust steam. The use of bagasse as an energy source has been ongoing for a long period of time. The quantity of excess bagasse was insufficient to warrant consideration of co-generation of electricity for export to the grid. For a factory grinding 300tc/h, when producing 95-100 tons bagasse per hour, gives a net surplus of 25-30 tons per hour or about 650 - 700 tc/day [12]. At this time of high production, the excess is resulting in a disposal problem; a problem that could readily be converted into energy and save the country a lot of foreign exchange. Bagasse is a by-product of cane sugar manufacture; hence the acquisition cost is low. However, there is the cost of conveying, storing and retrieving the material as required.

3.2 Co-generation systems
Most cane sugar factories have been designed to be energy self-sufficient, with sugar as the primary product. A bagasse-fired co-generation system, made up of “medium” pressure boilers (15-20 bar) plus small steam driven turbo-alternators, provides all the steam and electricity needed to run the
cane mills and factory, leaving little surplus bagasse as a “free” fuel. There are two types of bagasse-fired co-generation systems, which potentially offer much higher electricity production than those found in most sugar factories today:

(1) High-pressure condensing-extraction steam turbine system (CEST), and (2) steam-injected gas turbines run on gasified bagasse (GSTIG). “High pressure” (40-80 bar) refers to boiler pressures typical of condensing extraction steam turbines. “Medium pressure” (15-20 bar) refers to steam used for cane mills, which equals the boiler pressure in most sugar factories today. “Low pressure” (2-3 bar) refers to mill and turbo-alternator exhaust steam used in the process. High pressure or (CEST) co-generation systems are now used in a few cane sugar factories and are being considered for several others. When small medium pressure turbo-alternators are replaced with a high-pressure (CEST) system, the total electricity production can be increased from about 20 kWh/t (just enough to run the factory) to perhaps 70-120 kWh/t. Thus, in addition to sugar making, about 50-100 kWh/t becomes available for export to the utility grid. Electricity and steam production in co-generation system is that the electricity (in kWh/t), and steam production (in kg of medium pressure steam produced per ton cane) [14]. For both (CEST) and (GSTIG) co-generation systems, the steam and electricity production can be varied over a range of operating conditions, so that more electricity can be produced when the steam demand is lower.

For a high-pressure steam turbine co-generation system, the potential exists to boost in season export electricity production significantly by factory steam economy. For example, if the process steam demand were reduced from 500 to 400 kg/tc, an extra10.5 kWh/tc of electricity could be exported to the utility grid. For a factory crushing 175 tons of cane per hour (tc/h), this would mean an extra 1.84 MW of exportable electric
power in the season, more than a 10% increase. If the season is 210 days long, and the factory runs 23 hours per day, revenue over one season is about US$ 500,000, assuming that the electricity is worth $0.06/kWh. Moreover, decreasing the factory low pressure steam demand below 270-300 kg/tc means that the more electrically efficient gasifier/gas turbine co-generation system could potentially be used, and still meet factory process steam demands [14].

3.3 Steam usage

The sugar industry is a major consumer of steam, with the evaporators and vacuum pans consuming substantial quantities for concentration of juice and manufacture of sugar. Apart from these, the juice heaters, centrifuges, sugar dryers and sugar melting also consume some steam. The washing of pans and other equipment need some marginal steam. The steam consumption is discussed in detail below. In the case of old sugar factories the source of low-pressure steam and high-pressure steam are independent. Modern sugar factories use the combined unit type or the control unit type. Each boiler will distribute the high-pressure steam through the main or the common steam header, thereafter steam will be separated at several points for being directly consumed or conditioned before being used as per the requirement here-under.

1- High-pressure steam used directly at (20-30 bar) as superheated steams with a temperature of 350°C; for the steam turbine, the exhaust steam flow combined with low-pressure steam in the main line.

2- Low-pressure steam at (1-2 bar) as saturated steam originated from the superheated steam combined with the exhaust steam with a temperature of 120-125°C, for juice heater, evaporators, vacuum pans and heating elements in the process.
3- Make-up steam and desuperheated steam as low-pressure steam originated from the main line high-pressure steam, which has already passed through a pressure reducing valve and a desuperheated before being used as make up for the insufficient low-pressure steam [15].

Total energy in raw sugar factory with / without refinery: The modern cane sugar factories are designed to generate steam at around 30 bar or higher and 400°C. After the live steam is used in the mills and turbo-generators, the exhaust steam is used for all process requirements. The steam distributions based on a sugar factory with the operating conditions of table (3.3) is shown in Fig (3.1) [8].

Table (3.3): The process parameters.

<table>
<thead>
<tr>
<th>Cane grinding rate</th>
<th>4000   tc/d</th>
</tr>
</thead>
<tbody>
<tr>
<td>Actual grinding over 23h</td>
<td>174   tc/h</td>
</tr>
<tr>
<td>Cane Pol</td>
<td>12.5   %</td>
</tr>
<tr>
<td>Fiber</td>
<td>14.0   % cane</td>
</tr>
<tr>
<td>Imbibition water(on fiber)</td>
<td>200   %</td>
</tr>
<tr>
<td>Filtrate return</td>
<td>15   %</td>
</tr>
<tr>
<td>Mixed juice purity</td>
<td>82   %</td>
</tr>
</tbody>
</table>
3.4 Electricity systems:
Generally all turbo generators used in the sugar mills are of the non-condensing type. The inlet steam pressure of about (20-30 bar) as superheated steam with a temperature of 350°C gives the exhaust steam at pressure of about (1-2 bar) and can distribute electricity between 1.000-1.5000 kW at 380-11.000 volts, 3 phase and 50 Hz.

The main electricity lines are divided as follows:

- High voltage for 3300/6000 volts, 3-phase and 50 Hz distributing to electric motors, which require a big load such as the cane knife motor.
- Low voltage for 380 volts 3-phase and 50Hz distributing for lighting and general usage [15].
Chapter IV
Optimization of Energy in Cane Sugar Factories

4.1 Steam and energy economy in sugar factory

Potential for steam savings, and resulting increases in potential electric power output, exist in some factories. The available data did not allow a site by site engineering study to quantify present steam usage for each of the major unit operations in the sugar factories. Factors such as steam, juice flows, as well as temperatures need to be monitored. Due to the present lack of economic value of bagasse and steam, the factories are not equipped with adequate instrumentation to monitor parameters for material and energy balances for each unit operation. For this reason, the implementation of energy savings will require investments in engineering time and capital, which will vary from one case to another. The first step in such an undertaking will be to install instrumentation to monitor, measure and control various process parameters. Without such data as references, potential improvements cannot be accurately determined. Furthermore, the implementation of steam savings will require management to make, for example, a lower imbibition rate on cane will result in less steam being used in the juice evaporator, but may also result in a lower mill extraction. Depending on the economics, the management may decide to achieve more steam savings even at the expense of a slight reduction in sugar production [16]. Although steam consumption data were not available for the major unit operations, the factories did compile data on total steam consumption in kg per 100kg of cane. On average, steam consumption is approximately 55 kg per 100kg of cane in many of the factories surveyed. This number can be compared with approximately 40 kg per 100kg of cane achieved in some Hawaiian sugar factories. Based on this comparison, it is possible to
make general and qualitative suggestions on some areas for potential steam savings, although improvements cannot be quantified for each unit operation without performing detailed energy audits at such location. The potential areas of steam and bagasse savings are classified as short-term, medium-term and long-term based on ease of implementation [16].

4.1.1 Short-term steam economy

An obvious contributor to high steam usage is the generally high bagasse moisture in many factories. In Hawaii, bagasse moistures are around 47%, compared to 51% at the Indian sugar factories, and 50% in Alguneid. Reduction in bagasse moisture can be achieved operationally through tighter mill settings and lower imbibition rates. Improved cane preparation with knives and shredders will also help. High rates of imbibition cause dilution of mixed juice which in turn increases the juice evaporation load, requiring more exhaust steam into the first effect of the evaporator. High bagasse moisture also results in less steam being produced per ton of bagasse and hence in lowering boiler efficiency. These facts are well known to all sugar technologists, but without any economic value being attached to bagasse and steam savings, plant management has no incentive to risk reducing mill extraction or crushing rate by trying to save steam or bagasse. There is also potential for further steam economy in sugar factories by increasing the number of evaporator effects from four to five. The use of vapor for heating the juice heaters and the vacuum pans can be optimized by maximizing the use of steam bled from the lower pressure evaporator effects, and minimizing the use of exhaust steam. The use of quintuple effect evaporators has become more common in sugar factories like Hawaii as a result of the growing importance of co-generation. Presently, the use of pressure reduction valves is widespread, resulting in no electricity production when the steam expands. With double extraction condensing turbine the use of pressure reduction valves will be minimized,
resulting in more electricity generation from pressure reduction of steam. The use of automatic electronic control to optimize excess air in boiler operations should be considered as a means of increasing boiler efficiency.

4.1.2 Medium-term steam economy
There is a strong potential for reducing steam consumption through vacuum pan automation, which will reduce the amount of sugar dissolving and evaporation in the pans. Also the use of continuous vacuum pans should be considered. Savings in electrical power will result from the use of automatic continuous centrifuges, and steam savings in the mill turbines can be achieved by the introduction of electronic control of cane feed at the milling tandem in order to reduce fluctuations in steam flows. The implementation of these measures will require considerable investments and would not be economically justifiable without a guarantee of reasonable income from selling of electricity.

4.1.3 Long-term steam economy
Use of bagasse dryers to reduce bagasse moisture by the flue gases is a possibility of improving the calorific value of bagasse. The recovery of useful heat from the flue gases may be affected in a well-designed boiler system through the use of air preheaters, economizers and superheaters. As for reduction of bagasse moisture, it can be effectively accomplished by adjustments of mill settings and other mill management practices. While bagasse dryers may have some merit, there are many lower cost and simpler measures that can be undertaken before installing them. Besides using flue gas, solar drying may be attractive in this context. In addition to bagasse drying [16].

4.2 Means to save energy
This generally entails investments in higher-pressure boilers, condensing turbines, and efficiency improvements in the rest of the plant. The operator can choose to generate power only during the crushing season or to take
advantage of what would otherwise be idle generation capacity by keeping the power plant in operation during the remainder of the year. If co-
generation is profitable, conditions exist for adopting the most efficient technical options. The following techniques are recommended:

4.2.1 Extraction

Limitation of milling train to say a five-mill tandem. Extraction process is requiring less power than milling. Juice extraction by diffusion and two dewatering mills, guarantees the highest extraction, lowest energy consumption and mixed juice drafting at 111 % on cane. The concept is quite different from the conventional milling extraction and has also positive influence on the power demand for the extraction and subsequent processes [17].

4.2.2 Boilers

1. The use of high-pressure boilers of sufficient capacity to burn practically all the factory’s bagasse during crop time, with provision for an off- crop alternative fuel to guarantee continuous energy supply to the grid is also recommend. The choice of the pressure has to be viewed in the light of costs, efficiency and operating techniques, which should not exceed the technical capability of the cane sugar industry. High-pressure boilers combined with extraction and condensing turbines allow achieving specific power generation ratios of up to 120 kWh /ton cane [17].

2. Bagasse drying technologies and the use of automatic electronic control to optimize excess air in boiler operations should be considered as a means of increasing boiler efficiency. Because high bagasse moisture also results in a lowering of boiler efficiency and less steam being produced per ton of bagasse.

3. As for reduction of bagasse moisture, it can be effectively accomplished by adjustments of mill settings and other mill
management practices. In addition to bagasse drying, the use of mechanical vapor recompression to reduce steam consumption in the evaporator may be applicable in certain cases.

4.2.3 Co-generation

1. The use of large, high-efficiency passes out of the condensing turbines for generator drive. The implementation of the basified technologies is an important step to improve the power yield up to 300 kWh/t cane when applied in conjunction with combined gas and steam turbine cycles.

2. Presently, the use of pressure reduction valves is a widespread technique, resulting in no electricity production when the steam expands. With the use of a double extraction-condensing turbine, the use of pressure reduction valves will be minimized, resulting in more electricity being generated from pressure reduction of steam.

3. A third technique is the complete electrification of all prime movers including mill drives. Methods of reducing the high-pressure steam consumption of the prime movers to a level consistent with an improved low-pressure steam cycle include: improved efficiency of prime movers and drives, and the use of same prime movers not requiring steam as source of energy [17].

4.2.4 Evaporation

1. Technical achievable of multiple effect pressure evaporator, with upper conditions of pressure and temperature being limited to avoid adverse effects on juice quality and steam turbines efficiency, but high enough to allow maximum bleeding on the last effect. Such conditions are met by working between limits of 25 bar /139°C,
obtainable from a pass out/condensing turbo-generator and 2 bar /105°C [17].

2. There is also potential for further steam economy at many factories by increasing the number of evaporator effects from four to five. The use of vapor for heating the juice heaters and the vacuum pans can be optimized by maximizing the use of steam bled from the lower pressure evaporator effects, and minimizing the use of exhaust steam. The use of quintuple effect evaporators has become more common in some factories as a result of the growing importance of co-generation.

3. Introduction of technologies that offer the potential to decrease the process steam demand of cane sugar factories like plate heater and falling film evaporator should be considered also others like full electrification are major steps to operate cane sugar factories with a steam demand of below 30% on cane and have been already achieved in some sugar industries [3].

4. The pressure and temperature of these effects are raised appropriately for more effective use of vapors. The juice in evaporators is kept at low level, which decreases the hydrostatic loss, increases heat transfer efficiency, shortens the retention time of juice, and reduces the sucrose inversion and color formation in evaporators.

4.3 Parameters influencing steam consumption
In sugar factories, the evaporation station not only works as a concentrator to make the juice thicken into syrup, but also plays a role of reducing the pressure of steam. It works as a center of supplying vapors from different effects to juice heaters and vacuum pans. This makes the best use of steam and reduces its consumption. If quintuple-effect evaporators are applied,
the vapors bled from the preceding four effects are used for different heaters and pans. The pressure and temperature of these effects are raised appropriately for more effective use of vapors.

The syrup concentration is usually kept as high as 62-65%, and even 68-70% in advanced factories, resulting in lower total steam consumption.

4.4 Operation conditions
As already mentioned, the sugar industry is a major consumer of steam, with the evaporators and vacuum pans consuming substantial quantities for concentration of juice and manufacture of sugar. The steam consumers are discussed in detail below [18].

4.4.1 Evaporator
The evaporator is the major steam consumer in a sugar plant. The evaporator concentrates the juice from a level of 14 – 16° Brix to a level of 60 – 65° Brix. The exhaust steam is used for this purpose. Further to the concentration to a higher level, the concentrated syrup is transferred to the vacuum pan section, for evapo-crystallization to produce sugar. Several arrangements of evaporators are used in the sugar industry. The commonly used are the quadruple and quintuple-effect short-tube evaporators. Typically, the steam enters the first effect at a pressure of 1.8 bar, at a temperature of 105°C and the vacuum in the last effect is around 650 mm Hg. Another very important aspect, which needs to be considered in an evaporator, is the effective utilization of the area available in the evaporator. It is recommended to raise the clear juice temperature very close to the boiling point at the pressure prevailing in the first effect. The heating can be done by use of exhaust steam in a plate-type heat exchanger. The multiple effect evaporators have higher steam economics
of 3 to 5, depending on the number of effects. A good level of instrumentation is necessary in the evaporator system to maintain the quality of syrup and also operate the system efficiently. Presently, control systems are available for controlling steam flow, for maintaining the Brix of the syrup, based on the juice flow and evaporator temperature. The recovery of the condensate from the evaporator section is also very important for higher levels of thermal efficiency of the system. Generally, the exhaust condensate is directly taken to the boiler, while the later effect condensate (vapor-condensate) is utilized as hot water, for other purposes (imbibition, cleaning, centrifuge, washing……etc.)[18].

Apart from the standard type of evaporators, Kestner evaporator and falling film evaporators are gaining popularity usage in the sugar plants. Kestner type evaporator is very advantageous to have short retention time, high heat transfer coefficient and occupies less space. Also, as the liquid flows at higher velocities, the utilization of higher temperature steam can be done without any color / inversion effects.

Falling film evaporator (FFE) is another popular evaporator which is being considered by many sugar industries. The falling film evaporators have many advantages over the conventional evaporators as below.

1. The FFE's have better heat transfer, as there is no elevation in boiling point due to hydrostatic pressure.
2. The average contact time between juice and steam in a falling film evaporator is about 30 seconds as against 3 minutes in the Kestner evaporator and 6-8 minutes in the conventional short tube evaporator. This offers an excellent potential for using high temperature and pressure steam, as the risk of caramelization is
reduced. Hence, 2.8 bar steam at temperatures of 131.2°C can be used in the first effect of the evaporator station [19].

3. The design of the evaporators is such that, the juice is in contact with the heating surface in a thin layer over the length of the heating surface. So the vapor is entrained with the juice filling the interior of the tube. This avoids the vapor bubble obstruction, which is present in the conventional evaporator.

In view of the above advantages, it is recommended to install FFE's for the first three effects of the evaporator system. FFE cannot be used for the IV and V-effects, as the possibility of thick & sticky scale formation is high in the thickened juice. In such a configuration, there will be possibilities of extensive vapor bleeding, with only marginal vapor going to the condenser. The vapor bleeding can be as below:
1) III-effect vapor bleeding to raw juice heater.
2) II-effect vapor bleeding to pans. With continuous pans, is possibility to use third effect vapor also.
3) I-effect vapor bleeding to juice heater.
With such a system, it is possible to reduce the steam consumption in evaporators to less than 36% [18].

4.4.2 Vacuum pans

The vacuum pans are used for further concentrating the syrup produced in the evaporators, to massecuite and finally separated into sugar and molasses. Conventionally, the sugar industry has been using the batch pan. With the recent introduction of the continuous pans, there has been a reduction in the steam consumption to the extent of 15 – 20 % [17]. The continuous vacuum pan is an integrated system, with multiple compartments of gradually increasing volume, through which the
massecuite follows a methodical route. The heating is carried out in banks of horizontal / vertical tubes. Apart from the steam reduction, the utilization of continuous vacuum pans also result in

1. improved grain size,
2. reduced sugar losses,
3. better control and system,
4. reduced power consumption for injection water pumps.

4.4.3 Centrifuge wash water super-heater

The centrifuges are washed with hot water (at around 125°C). The hot water is generated by heating with M.P steam, in an indirect heat exchanger, since the LP steam cannot be used for generating 125°C temperature. Hence, it is recommended to install two-stage heating system for centrifuge wash water i.e. the hot water will be initially heated to 105°C by LP steam (exhaust steam) and further heating to the required temperature with medium pressure (MP) steam.

4.4.4 Sugar dryer

Conventionally, MP steam (7.85 bar) or HP steam (at generation pressure) is reduced and used for this purpose. The maximum temperature requirements for this purpose are only 80 – 90°C [18].

4.4.6 Conclusion

The sugar industry is a high energy intensive industry with good growth potential. The industry also has a potential for generating excess power, commonly referred as commercial co-generation, which can be sold to the grid. This can result in generating additional revenue for the sugar plant
and also contribute to the national revenue. The following aspects need to be strongly considered at the design stage to ensure energy efficiency.

1. Good cane preparation system to have more than 85% preparation.
2. Installation of LP extraction systems can be considered.
3. Utilization of hydraulic drives / DC motors for mill drives.
4. Installation of variable frequency drives for pumps.
5. Installation of falling film evaporators or Kestner type evaporators for initial evaporation.
6. Installation of continuous vacuum pans.
7. Installation of efficient multi-jet condensers, with hot water for jet and cold water for spray.
8. Installation of bigger size batch centrifugals with plough discharge.
9. Utilization of LP steam for sugar dryer, sugar dissolver and centrifuge water heating.
10. Installation of a high-pressure boiler with commercial co-generation system, for selling excess power to the grid [18].

These important aspects should be taken into consideration at the design stage, which can help the sugar industry to be energy efficient and competitive.
Chapter V
Materials and Heat Balances for the Case Study
“Alguneid Cane Sugar Factory”

5.1 Material balance
4600 t/d of sugar cane are to be processed. The factory operates for 23 hours per day. Thus 200 tons of sugar cane has to be processed per hour.

Note: 1. All units unless otherwise mentioned are in ton per hour.

Sugar cane properties of Alguneid Sugar Factory as below:
- Water: 67.0 %
- Sucrose: 13.0 %
- Fiber: 17.0 %
- Impurities: 3.0 %

Thus 200 tons cane feed per hour will contain:
- Water: 134.00
- Sucrose: 26.00
- Fiber: 34.00
- Impurities: 6.00

5.1.1 Milling plant
Water used in milling operation is 25 – 30% of sugar cane. Practical value is 25%.

Therefore imbibition water used = 0.25 × 200 = 50.0

Milling efficiency 93% i.e. 93% of sucrose goes into the juice.
Thus sucrose amount in juice = 0.93 × 26 = 24.18

Unextracted sucrose = 0.07 × 26 = 1.82

Assuming 70% of impurities stay in the juice and 30% goes with bagasse,

Impurities in the juice = 0.70 × 6 = 4.2
Impurities in bagasse = 0.30 × 6 = 1.8
The final bagasse from last mill contains the unextracted sucrose and impurities, fiber and 50% water.

Thus dry matter amount in bagasse = 34.0 + 1.82 + 1.8 = 37.62

Amount of bagasse = 50% dry matter + 50% water

= 37.62 dry matter + 37.62 water = 75.24

Overall output of mill or juice entering the clarifier will have the following composition:

Water = 134 + 50.0 – 37.62 = 146.38
Sucrose = 24.18
Impurities = 4.20

Thus total juice = 174.76

% of solids in this juice = {(24.18 + 4.2) × 100}/174.76= 16.24%

Purity of mixed juice = 24.18/ (24.18+4.20) = 0.85 i.e. 85%

5.1.2 Clarifier

Reagent used: lime 0.5 kg lime/ton of sugar cane

Thus lime to be added = 0.5 × 200 = 100 kg = 0.100 ton

Assume 6% efficiency of clarification process (impurities remove).

Therefore impurities to be removed = 0.06 × 4.2 = 0.252

Total sludge from clarifier = 0.100 + 0.252 = 0.352

Amount of clarified juice entering in the evaporator =174.76–0.252 =174.508

This clarified juice is fed to the first effect of the quadruple effect evaporator and analyses as follows:

Water 146.38
Sucrose 24.18
Impurities 3.95
Thus total juice 174.51

% of solids in this juice = {(24.18 + 3.95) × 100}/174.51= 16.12%

Purity of clear juice = 24.18/ (24.18+3.95) = 0.8596 i.e. 85.96%
5.1.3 Evaporator

Typical evaporator load = 75 – 80% of clarified juice

Input to the evaporator = 146.38 water + 24.18 sucrose + 3.95 impurities = 174.51

Quantity of water to be removed = \( m_f(1 - \frac{X_j}{X_s}) \)

\( m_f = \text{mass of juice input to the evaporator.} \)

\( X_j = \text{Brix of juice.} \quad X_s = \text{Brix of syrup.} \)

Water removed by evaporation = 174.51 \( (1 - \frac{16.12}{60}) = 127.62 \)

Water remaining in the juice = 146.38 – 127.62 = 18.76

Hence output of evaporator analyses as follows:

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>18.76</td>
</tr>
<tr>
<td>Sucrose</td>
<td>24.18</td>
</tr>
<tr>
<td>Impurities</td>
<td>3.95</td>
</tr>
<tr>
<td><strong>Thus total solution</strong></td>
<td><strong>46.89</strong></td>
</tr>
</tbody>
</table>

\% solids in this concentrated solution = \( \frac{(24.18 + 3.95)\times 100}{46.89} = 60\% \)

5.1.4 Crystallizer

Crystallization of concentrated juice is done in a vacuum pan crystallizer.

Crystallization is done at vacuum not exceeding 635 mmHg.

Hence assume 580 mm Hg vacuum in crystallizer.

Therefore absolute pressure = 760 – 580 = 180 mmHg

\[ = \frac{180 \times 1.013}{760} = 0.2399 \text{ bar} \]

Thus boiling point (from sugar tables) at this pressure = 64°C

Boiling point rise = 4°C

According to the equation by Hugot [20], solubility of sucrose in water is given by:

\[ S = 64.18 + 0.1348 \times t + 0.000531 \times t^2 \]

where \( S \) is \% sucrose at saturation, and \( t \) is the temperature in °C

At 64°C, \( S \) becomes = 74.98 %

Therefore sucrose per kg of water = \( \frac{S}{(100 - S)} = \frac{74.98}{(100 - 74.98)} \)

\[ = 2.997 \text{ kg Sucrose/ kg of water} \]
The purity can be calculated as

\[
Purity = \frac{\text{mass of sucrose}}{\text{total mass of solids}}
\]

\[
= \frac{24.18}{24.18 + 3.95} = 0.8596 \text{ i.e. } 85.96\%
\]

Hence the solubility is reduced by a factor called solubility coefficient.

For purity of 85.96 %, Solubility coefficient = 0.95

Thus effective solubility = \(2.997 \times 0.95 = 2.847\) kg of sucrose/kg of water

Input to crystallizer = (18.76 water + 24.18 sucrose + 3.95 impurities)

Assume 88.5% of sucrose is recovered as crystal with respect to initial sucrose content in feed.

Mass of sucrose crystal formed = \(0.885 \times 24.18 = 21.40\)

Moisture associated with crystals = 1% of crystal mass

\[
= 0.01 \times 21.4 = 0.214
\]

Sucrose in molasses = 24.18 - 21.4 = 2.78

Impurities left in molasses = 3.95 - 0.395 = 3.555

Total solid in molasses = 2.78 + 3.555 = 6.335

Water left in molasses = \(\frac{6.335}{2.847} = 2.225\)

Hence water to be evaporated = initial water – associated water – water in molasses

\[
= 18.76 - 0.214 - 2.225 = 16.321
\]

Output of crystallizer:

(1) Solids: 22.009

(21.4 sucrose + 0.214 water) crystals + 0.395 impurities

(2) Molasses: 8.56

(2.225 water + 2.78 sucrose + 3.555 impurities)

(1) + (2) = Input to centrifuge = 22.009 + 8.56 = 30.569

5.1.5 Centrifuge

Assume 10% molasses adheres to the crystal.

Output of centrifuge
= (21.4sucrose+0.214water) +0.395 impurities+0.856molasses =22.865

Wet crystals from centrifuge analyses as follows:

Sucrose = 21.4 + 0.278 = 21.678
Impurities = 0.395+0.3555 = 0.7505
Dry crystals = 21.678+0.7505 = 22.4285
Water = 0.214+ 0.2225 = 0.4365
Total = 22.865

5.1.6 Dryer
With a dryer, the moisture content may be reduced to between 0.2 and 0.5%.
Drying by contact with hot air involves heating the air, to increase its
capacity for absorbing water and bringing it into intimate contact with the
sugar from which it evaporates the moisture.
Assume that the final moisture content is 0.2%.
Mass of dry crystal is 22.4285 (as calculated before).
Therefore final sugar produced = 22.4285÷ (1 – 0.002) = 22.47 ton/h
Thus overall yield of raw sugar based on cane crushed
= (22.47÷ 200) × 100 = 11.235 %
Thus moisture to be removed, m = 22.865– 22.47
= 0.395 ton/h = 395 kg/h

There are two possible methods of circulation of air and sugar, i.e. parallel
flow and countercurrent flow.
For safety, the calculation is based on most unfavorable condition, that is the
ambient air fed to the heater before drying is assumed to be saturated. On the
other hand, the air leaving a dryer is generally saturated; it is assumed that in
case of countercurrent flow, it has absorbed only two- thirds of the quantity
of water that it could have absorbed if it had left in a saturated condition.
Then in the case of countercurrent condition,
A = (100 × m) ÷ {(2 ÷ 3) × (H₁ − H₀)} = (1500 × m) ÷ (H₁ − H₀)

Where, A = Mass of air to be passed through dryer (kg/h).

m = Moisture to be removed (kg/h).

H₀ = Mass of water vapor contained in saturated air at a temperature t₀ of entering air to the heater (ambient temperature) in kg/1000 kg. Taking ambient temperature as 30°C. From fig (5.1)[20].

Fig (5.1): Weight of water vapor in saturated air.

H₀ = 26 kg water/1000 kg saturated air.

H₁ = Mass of water vapor contained in saturated air at the temperature t₁ of exit from the dryer in kg/1000 kg.
The temperature of air leaving the dryer is between 45 – 52°C. Assuming it to be 50°C. From fig (5.1).

\[ H_1 = 85 \text{ kg water}/1000 \text{ kg saturated air}. \]

Therefore, \[ A = \frac{1500 \times 421.3}{85 - 26} = 10711.02 \text{ kg/h} \]

Thus mass of air to be passed through the dryer = \( A = 10711.02 \text{ kg/h} \)

Hence volume of air required= \( V = \frac{A}{a_0 + e_0} \). From fig(5.1)and fig(5.2) [20].

Fig (5.2): Weight of dry air per unit volume of saturated air.

\[ a_0 = \text{Density of air at } t_0 = 1.12 \text{ kg/m}^3 \]

\[ e_0 = \text{Mass of vapor contained in saturated air at } t_0 = 0.03 \text{ kg/m}^3 \]

Therefore, \[ V = 10711.02 \div (1.12 + 0.03) = 9313.9 \text{ m}^3/\text{h} \]
5.2 Heat Balance

5.2.1 Energy consumption
Live steam conditions 19.61 bar, 360°C
Exhaust steam conditions 2.16 bar, 124°C.
Two sets of knives: 930 kW ×10 = 9,300 kg/h
Six mills: live steam: 34 × 6× 15=3060 kW ×11.76 kg/kWh=36,000 kg/h
Power load: 3501 kW ×10 = 35,010 kg/h

5.2.2 Heater
Assuming the mixed juice from milling tandem is available at 30°C.
Before adding lime to juice, the juice has to be heated. Thus the final
temperatures to which juices are heated in clarification varies from
extremes of 90 to 115°C, although it is common practice to heat slightly
above the boiling point. But it is generally believed today that superheating
is not advantageous and temperatures just above the boiling point say 103-
105°C are the maximum for good practice.
Mixed juice is heated to 105°C in five heaters before clarifier, and it is
heated to 115°C in two heaters after clarifier.
Heat required to raise the temperature of mixed juice from $t_0$ to $t_1$ is given
by:
\[ Q = m \times \Delta t = m_s \times \lambda \]  
\[ \text{Cp} = \text{mean heat capacity of mixed juice} = 4.19 - 2.35 \times X \text{kJ/kg K} \]
Where: $X$ is mass fraction of dissolved solids. [21].
Feed to the heater is 174.51tons/h and % of solids content is 16. 24
(from material balance).
Therefore, \[ \text{Cp} = 4.19 - 2.35 \times 0.1624 = 3.81 \text{kJ/kg K} \]
and \[ m = 174.51 \times 1000 = 174510 \text{ kg/h} \]
1- Steam requirement in I.Heater
This heater is heated by third vapor coming from III-effect at temperature 88°C, so \( \lambda = 2288 \text{ kJ/kg} \)
\( \Delta t = 45 - 30 = 15\text{C°} \)
Thus from equation (1),
\[ Q_1 = 174510 \times 3.81 \times 15 = 9973246.5 \text{ kW} \]
Steam requirement, \( m_s = \frac{Q_1}{\lambda} = \frac{9973246.5}{2288} = 4358.9 \text{ kg/h} \)

2- Steam requirement in II.Heater
This heater is heated by second vapor coming from II-effect at temperature 100°C, so \( \lambda = 2256 \text{ kJ/kg} \)
\( \Delta t = 60 - 45 = 15\text{C°} \)
Thus from equation (1),
\[ Q_2 = 174510 \times 3.81 \times 15 = 9973246.5 \text{ kW} \]
Steam requirement, \( m_s = \frac{Q_2}{\lambda} = \frac{9973246.5}{2256} = 4420.77 \text{ kg/h} \)

3- Steam requirement in III.Heater
This heater is heated by first vapor coming from I-effect at temperature 115°C, so \( \lambda = 2216 \text{ kJ/kg} \)
\( \Delta t = 80 - 60 = 20\text{C°} \)
Thus from equation (1),
\[ Q_3 = 174510 \times 3.81 \times 20 = 13297662 \text{ kW} \]
Steam requirement, \( m_s = \frac{Q_3}{\lambda} = \frac{13297662}{2216} = 6000.75 \text{ kg/h} \)

4- Steam requirement in IV.Heater
This heater is heated by exhaust steam at temperature 124°C, so \( \lambda = 2192 \text{ kJ/kg} \)
\( \Delta t = 95 - 80 = 15\text{C°} \)
Thus from equation (1),
\[ Q_4 = 174510 \times 3.81 \times 15 = 9973246.5 \text{ kW} \]
Steam requirement, \( m_s = \frac{Q_4}{\lambda} = \frac{9973246.5}{2192} = 4549.84 \text{ kg/h} \)

5- Steam requirement in V.Heater
This heater is heated by exhaust steam at temperature 124°C, so \( \lambda = 2192 \) kJ/kg
\[ \Delta t = 105 - 95 = 10°C \]
Thus from equation (1),

\[ Q_5 = 174510 \times 3.81 \times 10 = 6648831 \text{ kW} \]

Steam requirement, \( m_s = \frac{Q}{\lambda} = \frac{6648831}{2192} = 3033.23 \text{ kg/h} \)

Amount of clarified juice heated in the second stage = 174.510 kg/h

% of solids in this juice = \( \frac{(24.18 + 3.95) \times 100}{174.51} = 16.12\% \)

Therefore \( C_p = 4.19 - 2.35 \times 0.1612 = 3.81 \text{ kJ/kg K} \)

6- Steam requirement in VI.Heater

This heater is heated by exhaust steam at temperature 124°C, so \( \lambda = 2192 \) kJ/kg
\[ \Delta t = 110 - 105 = 5°C \]
Thus from equation (1),

\[ Q_6 = 174510 \times 3.81 \times 5 = 3324415.5 \text{ kW} \]

Steam requirement, \( m_s = \frac{Q}{\lambda} = \frac{3324415.5}{2192} = 1516.61 \text{ kg/h} \)

7- Steam requirement in VII.Heater

This heater is heated by exhaust steam at temperature 124°C, so \( \lambda = 2192 \) kJ/kg
\[ \Delta t = 115 - 110 = 5°C \]
Thus from equation (1),

\[ Q_7 = 174510 \times 3.81 \times 5 = 3324415.5 \text{ kW} \]

Steam requirement, \( m_s = \frac{Q}{\lambda} = \frac{3324415.5}{2192} = 1516.61 \text{ kg/h} \)

5.2.3 Quadruple effect evaporator:

The evaporator station consists of four bodies.

From material balance

Feed to the first effect = \( F = 174.51 \text{ ton/h} = 174510 \text{ kg/h} \)

Fraction of solids in this feed = \( X_f = 0.1612 \)
Solution from the last effect = L₄ = 46.89 ton/h

= 46.89×1000=46890 kg/h

Fraction of solids in this output = X₄ = 0.6093

Therefore evaporators load = 174510 - 46890= 127620 kg/h

Assume equal amount of water is vaporized in each effect plus vapor bleeding goes to other processes.

V₁= V+ 6000.75 + 16053.56

V₂= V+ 4420.77

V₃= V+ 4358.9

V₄=V

Therefore, V₁ + V₂ + V₃ + V₄ = 4V+ 30833.98 =127620 kg/h

V= 24196.505

Material balance in each effect is as follows:

I-Effect

Liquid outlet from first effect = L₁ = F – V₁

= 174510 – 46250.815= 128259.185 kg/h

Now the solid balance is, F × Xᵢ = L₁ × X₁

Therefore, X₁ = (174510 ×0.1612/128259.185) = 0.219

In this liquid outlet, the solids per 100 parts of water = (21.9×100)/ (100–21.9) = 28.04

Boiling point rise C° according to Hugot [20]:

BPR = 2B / (100-B)

B = Brix of sugar solution

Therefore, boiling point rise l = BPR₁ = 0.56C°

II-Effect

Liquid outlet from second effect = L₂ = L₁ – V₂

=128259.185–28617.275= 99641.91 kg/h

Now the solid balance is, L₁ ×X₁ = L₂ ×X₂

Therefore, X₂ = (128259.185 ×0.219)/99641.91= 0.282
In this liquid outlet, the solids per 100 parts of water = \( \frac{28.2 \times 100}{100 - 28.2} \) = 39.28

Therefore, boiling point rise 2 = BPR\(_2\) = 0.77°C

III-Effect

Liquid outlet from third effect = \( L_3 = L_2 - V_3 \)

\[ = 99641.91 - 28555.405 = 71086.505 \text{ kg/h} \]

Now the solid balance is, \( L_2 \times X_2 = L_3 \times X_3 \)

Therefore, \( X_3 = \frac{99641.91 \times 0.282}{71086.505} = 0.395 \)

In this liquid outlet, the solids per 100 parts of water = \( \frac{39.5 \times 100}{100 - 39.5} \) = 65.29

Therefore, boiling point rise 3 = BPR\(_3\) = 1.31°C

IV-Effect

Liquid outlet from fourth effect = \( L_4 = 46890 \text{ kg/h} \)

Now the solid balance is, \( L_3 \times X_3 = L_4 \times X_4 \)

\[ X_4 = \frac{71086.505 \times 0.395}{46890} = 0.5988 \]

Therefore, \( X_4 = 0.60 \)

In this liquid outlet, the solids per 100 parts of water = \( \frac{60 \times 100}{100 - 60} \) = 150

Therefore, boiling point rise 4 = BPR\(_4\) = 3°C

To calculate \( \Delta t \)

Steam is available to first effect at \( P_s = 2.2 \text{ bar} \)

Therefore, \( t_{1s} = 124°C \)

The pressure in the vapor space of the 4\(^{th}\) effect is at a vacuum of 600mmHg.

i.e. \( P_4 = 760 - 600 = 160 \text{ Hg absolute} = 0.21 \text{ bar} \)

Thus, \( t_{5s} = 60°C \)

Therefore, the overall temperature drop \( \Delta t = t_{1s} - t_{5s} = 124 - 60 = 64 °C \)

Effective \( \Delta t = \Delta t - (BPR_1 + BPR_2 + BPR_3 + BPR_4) \)

\[ = 64 - (0.56 + 0.77 + 1.31 + 3) = 58.36°C \]
To calculate $\Delta t$ in each effect, neglecting the sensible heat necessary to heat the feed to the boiling point. Approximately all the latent heat of condensing steam appears as latent heat in the vapor.

Hence, $q_1 = q_2 = q_3 = q_4$

$U_1A_1\Delta t_1 = U_2A_2\Delta t_2 = U_3A_3\Delta t_3 = U_4A_4\Delta t_4$

Usually, the areas in all effects are equal.

Therefore, $U_1\Delta t_1 = U_2\Delta t_2 = U_3\Delta t_3 = U_4\Delta t_4$

According to Hugot [20], the overall heat transfer coefficients in each effect are given in table (5.1) [20]. The assumed overall heat transfer coefficient in each effect is in table (5.2), which depends on the physical properties of the sugar solution such as temperature.

Table (5.1): The overall heat transfer coefficients in each effect.

<table>
<thead>
<tr>
<th>Effect</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
</tr>
</thead>
<tbody>
<tr>
<td>$U$ (W/m$^2$ K)</td>
<td>2271-2839</td>
<td>1561.5-2129</td>
<td>1135.6-1531.5</td>
<td>709.78-851.7</td>
</tr>
</tbody>
</table>

Table (5.2): Assuming overall heat transfer coefficients in each effect.

<table>
<thead>
<tr>
<th>Effect</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
</tr>
</thead>
<tbody>
<tr>
<td>$U$ (W/m$^2$ K)</td>
<td>2555</td>
<td>1845</td>
<td>1420</td>
<td>795</td>
</tr>
</tbody>
</table>

Therefore, $(\Delta t_2/\Delta t_1) = U_1/U_2 = 2555/1845 = 1.385$

$\Delta t_2 = 1.385 \times \Delta t_1$

$\Delta t_3/\Delta t_2 = U_2/U_3 = 1845/1420 = 1.299$

$\Delta t_3 = 1.299 \times \Delta t_2 = 1.299 \times 1.385 \times \Delta t_1$

$\Delta t_4/\Delta t_3 = U_3/U_4 = 1420/795 = 1.7862$

$\Delta t_4 = 1.7862 \times \Delta t_3 = 1.7862 \times 1.299 \times 1.385 \times \Delta t_1$

But $\Delta t_1 + \Delta t_2 + \Delta t_3 + \Delta t_4 = 58.36$

$\Delta t_1 (1 + 1.385 + 1.299 \times 1.385 + 1.299 \times 1.385 \times 1.7862) = 58.36$
Therefore, $\Delta t_1 = 7.9$

$\Delta t_2 = 10.91$

$\Delta t_3 = 14.2$

$\Delta t_4 = 25.35$

58.36

The actual boiling point of the solution in each effect is calculated as follows:

I-Effect

Boiling point of solution = $t_1 = t_{1s} - \Delta t_1$

Where, $t_{1s}$ = saturated temperature of steam to first effect = 115°C

Therefore, $t_1 = 124 - 7.90 = 116.1°C$

II-Effect

$t_{2s} = $ saturated temperature of steam to second effect = $t_1 - BPR_1$

$= 116.1 - 0.56 = 115.54°C$

Boiling point of solution = $t_2 = t_{2s} - \Delta t_2$

$= 115.54 - 10.91 = 104.63°C$

III-Effect

$t_{3s} = $ saturated temperature of steam to third effect = $t_2 - BPR_2$

$= 104.63 - 0.77 = 103.86°C$

Boiling point of solution = $t_3 = t_{3s} - \Delta t_3$

$= 103.86 - 14.2 = 89.66°C$

IV-Effect

$t_{4s} = $ saturated temperature of steam to fourth effect = $t_3 - BPR_3$

$= 89.66 - 1.31 = 88.35°C$

Boiling point of solution = $t_4 = t_{4s} - \Delta t_4$

$= 88.35 - 25.35 = 63 °$

$t_{5s} = $ saturated temperature of steam to condenser = $t_4 - BPR_4$

$= 63 - 3 = 60°C$
Table (5.3): Temperature in each effect.

<table>
<thead>
<tr>
<th>I-Effect</th>
<th>II-Effect</th>
<th>III-Effect</th>
<th>IV-Effect</th>
<th>Condenser</th>
</tr>
</thead>
<tbody>
<tr>
<td>$t_{1s}$ = 124°C</td>
<td>$t_{2s}$ = 115.54°C</td>
<td>$t_{3s}$ = 103.86°C</td>
<td>$t_{4s}$ = 88.35°C</td>
<td>$t_{5s}$ = 60°C</td>
</tr>
<tr>
<td>$t_1$ = 116.1°C</td>
<td>$t_2$ = 104.63°C</td>
<td>$t_3$ = 89.66°C</td>
<td>$t_4$ = 63°C</td>
<td></td>
</tr>
</tbody>
</table>

Heat Balance

Heat capacity of sugar solution

The heat capacity of sugar solution in each effect is calculated from equation:

$$C_p = 4.19 - 2.35 \times X \text{ (kJ/kg K)}$$

where:

- $X$: Mass fraction of dissolved solids.
- $C_p$: Mean heat capacity (kJ/kg K). [21].

Table (5.4): Heat capacities of juice in each effect.

| $X_f = 0.1612$ | $C_{pf} = 3.81$ |
| $X_1 = 0.219$ | $C_{p1} = 3.68$ |
| $X_2 = 0.282$ | $C_{p2} = 3.53$ |
| $X_3 = 0.395$ | $C_{p3} = 3.26$ |
| $X_4 = 0.60$ | $C_{p4} = 2.78$ |
Table (5.5): Enthalpy data from steam tables.

<table>
<thead>
<tr>
<th>Temperature (°C)</th>
<th>Enthalpy of saturated water vapor (h_g :kJ/kg)</th>
<th>Enthalpy of saturated liquid water (h_f :kJ/kg)</th>
<th>Enthalpy ( \lambda ) (hfg :kJ/kg)</th>
</tr>
</thead>
<tbody>
<tr>
<td>( t_{1s} = 124 )</td>
<td>h_g = 2712</td>
<td>h_f = 521</td>
<td>( \lambda_{1s} = 2191 )</td>
</tr>
<tr>
<td>( t_{2s} = 115.54 )</td>
<td>h_g = 2700</td>
<td>h_f = 485</td>
<td>( \lambda_{2s} = 2215 )</td>
</tr>
<tr>
<td>( t_{3s} = 103.86 )</td>
<td>h_g = 2682</td>
<td>h_f = 437</td>
<td>( \lambda_{3s} = 2245 )</td>
</tr>
<tr>
<td>( t_{4s} = 88.35 )</td>
<td>h_g = 2658</td>
<td>h_f = 372</td>
<td>( \lambda_{4s} = 2286 )</td>
</tr>
<tr>
<td>( t_{5s} = 60 )</td>
<td>h_g = 2609</td>
<td>h_f = 251</td>
<td>( \lambda_{5s} = 2358 )</td>
</tr>
</tbody>
</table>

Table (5.6): Heat capacity of steam in each effect (from steam tables).

<table>
<thead>
<tr>
<th>Temperature in each effect (°C)</th>
<th>Specific heat of steam (kJ/kg K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>( t_1 = 116.1 )</td>
<td>2.071</td>
</tr>
<tr>
<td>( t_2 = 104.63 )</td>
<td>2.027</td>
</tr>
<tr>
<td>( t_3 = 89.66 )</td>
<td>1.963</td>
</tr>
<tr>
<td>( t_4 = 63 )</td>
<td>1.917</td>
</tr>
</tbody>
</table>

I-Effect

The condensate is assumed to leave at the same temperature of the heating steam so, the heat balance for first effect is as follows:

\[ m_S \lambda_{1s} + m_F h_f = m_{V1} h_{V1} + m_{L1} h_{L1} \]

\( \lambda_{1s} = h_g - h_f = 2712 - 521 = 2191 \text{ kJ/kg} \)

\( m_F = 174510 \text{ kg/h} \)

Taking reference temperature as 0°C and feed to the first effect is at 115°C.

Thus \( t_f = 115^\circ \text{C} \)

\( h_f = C_p f \times (t_f - 0) = 3.81 \times 115 = 438.15 \text{ kJ/kg} \)

\( h_{V1} = h_{g2} + C_p \times \text{BPR}_1 \)

Where: \( C_{ps} \) is heat capacity of heating steam 124°C.

\( h_{V1} = 2700 + 2.113 \times 0.56 = 2701.2 \text{ kJ/kg} \)
\[ h_{L1} = C_p \times t_1 = 3.68 \times 116.1 = 427.25 \text{ kJ/kg} \]

\[ m_{L1} = 174510 - m_{V1} \]

Therefore

\[ m_s \times 2191 + 174510 \times 438.15 = m_{V1} \times 2701.2 + (174510 - m_{V1}) \times 427.25 \]

\[ 2191 \times m_s - 2273.95 \times m_{V1} = -1902508.02 \]

\[ m_s - 1.038 \times m_{V1} = -868.33 \]  

\[ \text{---(1)} \]

II-Effect

The heat balance for second effect is as follows:

\[ m_{V1} \lambda_{2s} + m_{L1} h_{L1} = m_{V2} h_{V2} + m_{L2} h_{L2} \]

\[ \lambda_{2s} = h_{V1} - h_{f2} = 2701.2 - 485 = 2216.2 \text{ kJ/kg} \]

\[ h_{V2} = h_{g3} + C_p t_1 \times \text{BPR}_2 \]

Where: \( C_p \) is heat capacity of steam.

\[ h_{V2} = 2682 + 2.071 \times 0.77 = 2683.6 \text{ kJ/kg} \]

\[ h_{L2} = C_p \times t_2 = 3.53 \times 104.63 = 369.34 \text{ kJ/kg} \]

\[ m_{L2} = 174510 - m_{V1} - m_{V2} \]

Therefore

\[ m_{V1} \times 2216.2 + (174510 - m_{V1}) \times 427.25 = m_{V2} \times 2683.6 + (174510 - m_{V1} - m_{V2}) \times 369.34 \]

\[ m_{V1} - 1.018 \times m_{V2} = -4443.9 \]  

\[ \text{---(2)} \]

III-Effect

The heat balance for third effect is as follows:

\[ m_{V2} \lambda_{3s} + m_{L2} h_{L2} = m_{V3} h_{V3} + m_{L3} h_{L3} \]

\[ \lambda_{3s} = h_{V2} - h_{f3} = 2683.6 - 437 = 2246.6 \text{ kJ/kg} \]

\[ h_{V3} = h_{g4} + C_p t_1 \times \text{BPR}_3 \]

Where: \( C_p \) is heat capacity of steam.

\[ h_{V3} = 2658 + 2.027 \times 1.31 = 2660.7 \text{ kJ/kg} \]

\[ h_{L3} = C_p \times t_3 = 3.26 \times 89.66 = 292.3 \text{ kJ/kg} \]

\[ m_{L3} = 174510 - m_{V1} - m_{V2} - m_{V3} \]

Therefore
\[ m_{V2} \times 2246.6 \, + \, (174510 \, - \, m_{V1} \, - \, m_{V2}) \times 369.34 \, = \, m_{V3} \times 2660.7 \, + \, (174510 \, - \, m_{V1} \, - \, m_{V2} \, - \, m_{V3}) \times 292.3 \]

\[ m_{V3} \, = \, 5676.5 \, - \, 0.033 \times m_{V1} \, + \, 0.92 \times m_{V2} \] (3)

**IV-Effect**

The heat balance for third effect is as follows:

\[ m_{V3} \lambda_{4s} \, + \, m_{L3} h_{L3} \, = \, m_{V4} h_{V4} \, + \, m_{L4} h_{L4} \]

\[ \lambda_{4s} \, = \, h_{V3} \, - \, h_{f4} \, = \, 2660.7 \, - \, 372 \, = \, 2288.7 \, \text{kJ/kg} \]

\[ h_{V4} \, = \, h_{g5} \, + \, C_{P3} \times BPR_{4} \]

Where: \( C_{p} \) is heat capacity of steam.

\[ h_{V4} \, = \, 2609 \, + \, 1.963 \times 3 \, = \, 2614.9 \, \text{kJ/kg} \]

\[ h_{L4} \, = \, C_{p4} \times t_{4} \, = \, 2.78 \times 63 \, = \, 175.14 \, \text{kJ/kg} \]

\[ m_{L4} \, = \, 174510 \, - \, m_{V1} \, - \, m_{V2} \, - \, m_{V3} \, - \, m_{V4} \]

Therefore

\[ m_{V3} \times 2288.7 \, + \, (174510 \, - \, m_{V1} \, - \, m_{V2} \, - \, m_{V3}) \times 292.3 \, = \, 2614.9 \times m_{V4} \, + \, (174510 \, - \, m_{V1} \, - \, m_{V2} \, - \, m_{V3} \, - \, m_{V4}) \times 175.14 \]

\[ m_{V4} \, = \, 8380.2 \, - \, 0.048 \times m_{V1} \, - \, 0.048 \times m_{V2} \, + \, 0.89 \times m_{V3} \] (4)

But \( m_{V1} \, + \, m_{V2} \, + \, m_{V3} \, + \, m_{V4} \, = \, 127620 \) (5)

The solution of equations (1) to (5), will give,

\[ m_{V1} \, = \, 28508.2 \, \text{kg/h} \]

\[ m_{V2} \, = \, 32361.3 \, \text{kg/h} \]

\[ m_{V3} \, = \, 34490.9 \, \text{kg/h} \]

\[ m_{V4} \, = \, 36035.75 \, \text{kg/h} \]

\[ m_{S} \, = \, 28723.2 \, \text{kg/h} \]

Saturated steam required in evaporators, as it is calculated before is \( m_{S} \, = \, 28723.2 \, \text{kg/h} \)

Saturation temperature of steam = 124°C at which \( \lambda \, = \, 2192 \, \text{kJ/kg} \)

Therefore heat supplied by this steam = \( m_{S} \times \lambda \)

\[ = \, 28723.2 \times 2192 \, = \, 62961254.4 \, \text{kW} \]

Steam economy = amount of water evaporated (evaporator load)/steam used
5.2.4 Crystallizer

As it is known from the material balance calculations, the crystallizer operates at 580 mmHg vacuum.

At this pressure, boiling point of water is 61.42°C.

But boiling point rise (BPR) is 4°C.

Temperature in crystallizer = 65.42°C

Heat balance in crystallizer can be written as

\[ m_F h_F + m_s \lambda = m_L h_L + m_V h_v \] \[ \text{-----------------------------(1)} \]

F = feed rate to crystallizer

\[ = (18.76 \text{water} + 24.18 \text{sucrose} + 3.95 \text{impurities}) \times 1000 = 46890 \text{ kg/h} \]

\[ C_{pf} = 4.19 - 2.35 \times X \]

Here \( X = 0.60 \)

Therefore \( C_{pf} = 2.78 \text{ kJ/kg K} \)

Feed temperature = \( t_f = 55.45°C \)

Taking base temperature as 0°C.

But \( h_F = C_{pf} \times (t_f - 0) = 2.78 \times 55.45 = 154.15 \text{ kJ/kg} \)

\( m_s = \) steam rate required in kg/s

This pan is heated by first vapor coming from 1\textsuperscript{st} effect at temperature 115°C, so \( \lambda = 2216 \text{ kJ/kg} \)

\( m_L = \) output of crystallizer

\[ = (22.009 + 8.823) \times 1000 = 30.832 \times 1000 = 30832 \text{ kg/h} \]

\[ C_{pl} = 4.19 - 2.35 \times X \]

Where \( X = \) Mass fraction of solids in crystallizer output

\[ = 22.009/30.832 = 0.7138 \]

Thus \( C_{pl} = 4.19 - 2.35 \times 0.7138 = 2.51 \text{ kJ/kg K} \)

Temperature of output of crystallizer = \( t_l = 65.42°C \)

But \( h_L = C_{pl} \times (t_l - 0) = 2.51 \times 65.42 = 164.37 \text{ kJ/kg} \)
\[ m_V = \text{amount of water evaporated in the crystallizer} \]
\[ = 16.058 \times 1000 = 16058 \text{ kg/h} \]
\[ h_v = \lambda \text{ at } 65.42^\circ C + (\text{BPR} \times \text{Cp steam}) \]
\[ \lambda \text{ at } 65.42^\circ C = 2342.2 \text{ kJ/kg} \]
\[ \text{Cp steam at } 65.42^\circ C = 1.928 \text{ kJ/kg K} \]
Therefore, \[ h_v = 2342.2 + (4 \times 1.928) = 2349.91 \text{ kJ/kg K} \]
Now equation (1) becomes,
\[ (46890 \times 154.15) + (m_s \times 2216) = (30832 \times 164.37) + (16058 \times 2349.91) \]
Hence \[ m_s = \frac{35574688.94}{2216} = 16053.56 \text{ kg/h} \]
Steam to be supplied to the crystallizer is \[ m_s = 16053.56 \text{ kg/h} \]

5.2.5 Centrifugal separator
Feed to the separator = 30.569 ton/h
For this capacity and for speed of 1500 rpm, the power required for an operation as suggested by Hugot = 45 kW
Table (5.7) shows the power consumption of centrifugals at speed [20].
Table (5.7): Power consumption of centrifugals at speed

<table>
<thead>
<tr>
<th>Dimension</th>
<th>H.P. required at 1,000 r.p.m.</th>
<th>H.P. required at 1,500 r.p.m.</th>
<th>H.P. required at 2,000 r.p.m.</th>
</tr>
</thead>
<tbody>
<tr>
<td>mm</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1.015 x 510</td>
<td>3.8</td>
<td>11.9</td>
<td>27.2</td>
</tr>
<tr>
<td>1.015 x 610</td>
<td>4.5</td>
<td>14.2</td>
<td>32.6</td>
</tr>
<tr>
<td>1.015 x 760</td>
<td>5.7</td>
<td>17.9</td>
<td>40.9</td>
</tr>
<tr>
<td>1.065 x 510</td>
<td>4.6</td>
<td>14.4</td>
<td>----</td>
</tr>
<tr>
<td>1.065 x 610</td>
<td>5.5</td>
<td>17.3</td>
<td>----</td>
</tr>
<tr>
<td>1.065 x 760</td>
<td>6.9</td>
<td>21.8</td>
<td>----</td>
</tr>
<tr>
<td>1.220 x 760</td>
<td>11.8</td>
<td>37.1</td>
<td>----</td>
</tr>
<tr>
<td>1.220 x 915</td>
<td>14.1</td>
<td>44.5</td>
<td>----</td>
</tr>
<tr>
<td>1.370 x 1.065</td>
<td>26.4</td>
<td>83.3</td>
<td>----</td>
</tr>
</tbody>
</table>

5.2.6 Dryer

The heating surface of air heater is calculated as follows:

The air heater generally consists of tubes, which are supplied with steam at about 3 bar. The condensate leaves at bottom.

The heating surface of this small heat exchanger will be given by

\[
A = Q \div \left[ U \times \left( t - \left( t_1 + t_0 \right)/2 \right) \right]
\]

Where, \( A \) = heating surface of air heater (m\(^2\)).

\( Q \) = Quantity of heat to be transmitted in kJ/h.

\( U \) = Heat transfer coefficient in kJ/(m\(^2\)h°C).

\( t \) = Temperature of saturated steam employed in °C.

\( t_0 \) = Temperature of cold air entering = 30°C.

\( t_1 \) = Temperature of hot air leaving = 50°C.

Since steam is available at 3 bar, therefore \( t = 133.5°C \) (from steam table).

As suggested by Hugot [20], \( U = 41.9 \) kJ/(m\(^2\)h°C).
The quantity of heat $Q$ to be transferred is calculated from the fact that it consists of three principle terms.

(i) The heat necessary to heat the weight of air is given by:

$$q_1 = M \times C \times (t_1 - t_0)$$

$M$ = Amount of air entering = 7451.69 kg/h
$C$ = Specific heat of air = 1 kJ/kg K

Therefore: $q_1 = 7451.69 \times 1 \times (50 - 30) = 149754$ kJ/h

(ii) The heat necessary to evaporate the water contained in the sugar is given by:

$$q_2 = m \times [607 + 0.3 \times t_1 - t_0]$$

$m$ = Amount of water evaporated = 421.3 kg/h

Thus: $q_2 = 395 \times [607 + 0.3 \times 50 - 30]$

$$= 244505\text{kcal/h} = 1023693.50 \text{kJ/h}$$

The quantity $q_2$ of heat correspond to that given up by the hot air between $T$ reached at the outlet from the air heater and $t_1$, which has not been included in $q_1$.

(iii) The heat necessary to heat the vapor contained in the mass $M$ of air, assumed saturated is given by

$$q_3 = M \times H_0 \times C \times (t_1 - t_0)$$

$C$ = specific heat of vapor = 1.99 kJ/kg K
$H_0$ = it is expressed as kg of water per kg of air = 0.026

Therefore $q_3 = 7451.69 \times 0.026 \times 0.475 \times (50 - 30)$

$$= 7706 \text{kJ/h}$$

The heat lost to ambient air $Q$, is taken into account by the following expression:

$$Q = 1.25 \times (q_1 + q_2 + q_3)$$

$$= 1.25 \times (149754 + 1023693.50 + 7706)$$

$$= 1476441.875 \text{kJ/h}$$

Therefore heating surface of heat exchanger will be,
\[ A = 1476441.875 \div [41.9 \times \{133.5 - (30 + 50) / 2\}] \]
\[ = 376.87 \text{ m}^2 \]

The steam consumption will be:

\[ ms = Q \div \lambda \]

\( ms \) = Steam consumption for air heater (kg/h)
\( \lambda \): Latent heat of steam used = 2164.41 kJ/kg

Therefore, \( ms = 1476441.875 \div 2164.41 = 682.14 \text{ kg/h}. \)
Chapter VI

Energy Consumption of an Ideal Sugar Factory

6.1 Sugar factory specifications

A sugar factory model will be used for the calculations with the following specifications:

Capacity of sugar factory is 6000 t/d of sugar cane are to be processed. Assuming the factory operates 24 hours per day, thus 250 tons of sugar cane will be processed per hour.

Sugar cane properties:

Water 70.0 %
Sucrose 13.5 %
Fiber 14.0 %
Impurities 2.5%

Thus 250 tons cane feed per hour contain:

Water 175.00
Sucrose 33.75
Fiber 35.00
Impurities 6.25

6.2 Sugar manufacturing

6.2.1 Preparation of cane for milling

Preparation of cane is accomplished by combination of two sets of revolving cane knives that cut the cane into chips, and by sherdders that tear the cane into small pieces but extract no juice.

6.2.2 Milling

Assume 5 mill tandems and each mill consists of three rollers. Water used in milling operation is 28% of sugar cane, and the mill extraction is 95%. The final bagasse has moisture content of 50%.
6.2.3 Clarification
Reagent used is lime (0.5kg/ton cane). The efficiency of clarification process (impurities remove) is 6%.

6.2.4 Heating
There are 4 heaters before clarifier that raise the temperature of the juice from 35°C to 103°C as follows:
First heater heats the juice from 35°C to 55°C.
Second heater heats the juice from 55°C to 80°C.
Third heater heats the juice from 80°C to 90°C.
Fourth heater heats the juice from 90°C to 103°C.
There are also 3 heaters after clarifier heating the juice to the temperature of the first body of evaporators as follows:
Fifth heater heats the juice from 103°C to 113°C.
Sixth heater heats the juice from 113°C to 120°C.
Seventh heater heats the juice from 120°C to 125°C.
Heating is assumed to be in a heating circuit as shown in table (6.1).

Table (6.1): Heating circuit.

<table>
<thead>
<tr>
<th>Heater no</th>
<th>Heating range</th>
<th>Heating agent</th>
<th>Temperature of heat agent</th>
</tr>
</thead>
<tbody>
<tr>
<td>1st</td>
<td>35 – 55 °C</td>
<td>5th vapor</td>
<td>87 °C</td>
</tr>
<tr>
<td>2nd</td>
<td>55 – 80 °C</td>
<td>Condensates</td>
<td>99.5 °C</td>
</tr>
<tr>
<td>3rd</td>
<td>80 – 90 °C</td>
<td>4th vapor</td>
<td>99.5 °C</td>
</tr>
<tr>
<td>4th</td>
<td>90 – 103 °C</td>
<td>3rd vapor</td>
<td>109.5 °C</td>
</tr>
<tr>
<td>5th</td>
<td>103 – 113 °C</td>
<td>2nd vapor</td>
<td>118.5 °C</td>
</tr>
<tr>
<td>6th</td>
<td>113 – 120 °C</td>
<td>1st vapor</td>
<td>124.5 °C</td>
</tr>
<tr>
<td>7th</td>
<td>120 – 125 °C</td>
<td>Exhaust steam</td>
<td>131.2 °C</td>
</tr>
</tbody>
</table>
6.2.5 Evaporation
In the evaporators the juice is concentrated from 14 to 65° Brix in five bodies of evaporators. Exhaust steam is used for heating the first body, the temperatures in each body is as in table (6.2).

Table (6.2): Temperature of evaporator bodies.

<table>
<thead>
<tr>
<th>No of body (effect)</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
</tr>
</thead>
<tbody>
<tr>
<td>t of heating agent (°C)</td>
<td>131.2</td>
<td>124.5</td>
<td>118.5</td>
<td>109.5</td>
<td>99.5</td>
</tr>
<tr>
<td>t in body (°C)</td>
<td>125</td>
<td>119</td>
<td>110</td>
<td>100</td>
<td>87</td>
</tr>
</tbody>
</table>

6.2.6 Crystallization
Boiling all A, B and C sugars is in continuous pans. Seed for the continuous pans is produced in two stirred batch pans, one for A seed and the other for B or C seed. A sugar is the only product marketed. B sugar is used as magma for an A seed; C sugar is dissolved into raw syrup, as is any surplus B magma.

6.2.7 Centrifugation
Batch centrifuges are used for A machines and continuous centrifuges on B and C sugars.

6.2.8 Drying
A rotary drier is used for A sugar only.

6.3 Material balance
6.3.1 Milling plant
Water used in milling operation is assumed to be 28% of sugar cane mass. Therefore imbibition water used = 0.28 × 250 = 70.0 t/h
Assuming milling efficiency to be 95% i.e. 95% of sucrose goes into the juice, thus sucrose amount in juice = 0.95 × 33.75 = 32.0625
Unextracted sucrose = 0.05 × 33.75 = 1.6875
Impurities in the juice = 0.95 × 6.25 = 5.9375
Impurities in bagasse = 0.05 × 6.25 = 0.3125
The final bagasse from last mill contains the unextracted sucrose, impurities, fiber and 50% water.
Thus dry matter content in bagasse = 35.0 + 0.3125 + 1.6875 = 37.0
Amount of bagasse = 50% dry matter + 50% water
= 37.0 dry matter + 37.0 water = 74.0
Overall output of mill or juice entering the clarifier will have the following composition:
Water = 175 + 70.0 − 37.0 = 208
Sucrose = 32.0625
Impurities = 5.9375
Total juice = 246
% of solids in this juice = \{(32.0625 + 5.9375) × 100\}/246 = 15.45%
Purity of mixed juice = 32.0625/ (32.0625 + 5.9375) = 0.84375 i.e. 84.4%

6.3.2 Clarifier
Reagent used: lime (0.5 kg lime / ton of sugar cane)
Thus lime to be added = 0.5 × 250 = 125kg = 0.125 ton
Assume 6% is the efficiency of clarification process (impurities remove).
Therefore impurities to be removed = 0.06 × 5.9375 = 0.35625
Total sludge removed from clarifier = 0.125 + 0.35625 = 0.48125
Amount of clarified juice entering in the evaporator = 246− 0.35625 = 245.64375 t/h
This clarified juice is fed to the first effect of the quintuple effect evaporator and its contents are as follows:
Water                                    208  
Sucrose                                 32.0625  
Impurities                              5.58125  
Total juice                             245.64375  

% of solids in this juice = \{(32.0625 + 5.58125) \times 100\}/245.64375 = 15.32%  
Purity of clear juice = 32.0625/ (32.0625 + 5.58125) = 85.17 %  

6.3.3 Evaporator station  
Typical evaporator load = 75 – 80% of clarified juice  
Input to the evaporator = (208 water + 32.0625sucrose + 5.58125 impurities)  
= 245.64375  
Quantity of water to be removed = m(1 - Xj / Xs)  
Mj=mass of juice input to the evaporator.  
Xj = Brix of juice. Xs =Brix of syrup.  
Water removed by evaporation = 245.64375 (1 – 15.32 /65) =187.75  
Water remaining in the juice = 208 – 187.75= 20.25  
Hence output of evaporator contents is as follows:  
Water                                   20.25  
Sucrose                                32.0625  
Impurities                           5.58125  
Total solution                      57.89375  
Solids in this concentrated solution=(32.0625+5.58125)×100/57.89375=65  

6.3.4 Crystallizer  
Crystallization of concentrated juice is done in a vacuum pan crystallizer, at  
a vacuum not exceeding 635mm.  
Hence assume 580 mm Hg vacuum in crystallizer.  
Therefore absolute pressure = 760 - 580= 180 mmHg  
= (180 \times 1.013)/760 bar= 0.2399bar
Thus boiling point (from steam table) at this pressure = 64°C
Boiling point rise = 4°C
According to equation by (Hugot, 1972)[20]:
Solubility of sucrose in water is given by
\[ S = 64.18 + 0.1348 \times t + 0.000531 \times t^2 \]
Where S is % sucrose at saturation, t is temperature in °C
At 64°C, S becomes = 74.98 %
Therefore sucrose per kg of water = \( S / (100 - S) \) = 74.98 / (100 - 74.98) = 2.997 kg Sucrose/ kg of water
The purity can be calculated as
Purity = mass of sucrose/ total mass of solids
\[ = \frac{32.0625}{32.0625 + 5.58125} \] = 0.8517i.e. 85.17 %
Hence the solubility is reduced by factor called solubility coefficient.
For purity of 85.17 %, Solubility coefficient = 0.95
Thus effective solubility = 2.997 \times 0.95 = 2.847 kg of sucrose/kg of water
Input to crystallizer = (20.25 water + 32.06 25 sucrose + 5.58125 impurities) = 57.89375
Assume 92% of sucrose is recovered as crystal with respect to initial sucrose content in feed.
Mass of sucrose crystal formed = 0.92 \times 32.0625 = 29.5
Moisture associated with crystals = 1% of crystal mass
\[ = 0.01 \times 29.5 = 0.295 \]
Sucrose in molasses \[ = 32.0625 - 29.5 = 2.5625 \]
Impurities left in molasses \[ = 5.58125 - 0.558125 = 5.023125 \]
Total solid in molasses = 2.5625 + 5.023125 = 7.585625
Water left in molasses = 7.585625/2.847 = 2.664
Hence water to be evaporated = initial water – associated water – water in molasses
\[ = 20.25 - 0.295 - 2.664 \] molasses
\[ = 17.291 \]
Out put of crystallizer:

(1) Solids: 30.353
(29.5 sucrose + 0.295 water) crystals + 0.558 impurities

(2) Molasses: 10.2516
(2.664 water + 2.5625 sucrose + 5.023 impurities)

(1) + (2) = Input to centrifuge = 30.353 + 10.2516 = 40.6046

6.3.5 Centrifuge
Assume 10% of molasses adheres to the crystal.
Output of centrifuge
= (29.5 sucrose + 0.295 water) + 0.558 impurities + 1.02516 molasses
= 31.37816

Wet crystals from centrifuge analyses as follows:

Sucrose = 29.5 + 0.256 = 29.756
Impurities = 0.558 + 0.5023 = 1.060
Dry crystals = 29.756 + 1.060 = 30.816
Water = 0.295 + 0.2664 = 0.5614
Total = 31.37816

6.3.6 Dryer
With a dryer, the moisture content may be reduced to between 0.2 and 0.5%.
Drying by contact with hot air involves heating the air to increase the capacity for absorbing water and bringing it into intimate contact with the sugar from which it evaporates the moisture.
Assume that the final moisture content is 0.2%.
Mass of dry crystal is 30.816 as calculated before.
Therefore final sugar produced = 30.816 ÷ (1 – 0.002) = 30.88
Thus overall yield of raw sugar based on cane crushed
= (30.88 ÷ 250) × 100 = 12.35 %
Thus moisture to be removed, \( m = 31.37816 - 30.88 \)

\[ = 0.49816 \text{ t/h} = 498.16 \text{ kg/h} \]

There are two possible methods of circulation of air and sugar, i.e. parallel flow and countercurrent flow. For safety, the calculation is based on most unfavorable condition, i.e. it is assumed that the ambient air is saturated. On the other hand, the air leaving a dryer is generally not saturated; it is assumed that in case of countercurrent flow, it has absorbed only two-thirds of the quantity of water that it could have absorbed if it had left in a saturated condition.

Then, in the case of countercurrent condition

\[ A = (100 \times m) ÷ \{(2 ÷ 3) \times (H_1 - H_0)\} = (1500 \times m) ÷ (H_1 - H_0) \]

Where, 

- \( A \) = Mass of air to be passed through dryer (kg/h)
- \( m \) = Moisture to be removed (kg/h)
- \( H_0 \) = Mass of water vapor contained in saturated air at a temperature \( t_0 \) of air entering the heater (ambient temperature) in kg/1000 kg. Taking ambient temperature as 30°C. From fig( 5.1).

\[ H_0 = 26 \text{ kg water/1000 kg saturated air.} \]

- \( H_1 \) = Mass of water vapor contained in saturated air at the exit temperature \( t_1 \) from the dryer in kg/1000 kg.

The temperature of air leaving the dryer is in the range of 45 – 52°C. Assuming it to be 50°C. From fig (5.1).

\[ H_1 = 85 \text{ kg water/1000 kg saturated air.} \]

Therefore, 

\[ A = (1500 \times 529) ÷ (85 - 26) = 13449.15 \text{ kg/h} \]

Thus mass of air to be passed through the dryer \( A = 13449.15 \text{ kg/h} \)

Hence volume of air required = \( V = A ÷ (a_0 + e_0) \)

From fig (5.1)and fig (5.2).

- \( a_0 \) = Density of air at \( t_0 = 1.12 \text{ kg/m}^3 \)
- \( e_0 \) = Mass of vapor contained in saturated air at \( t_0 = 0.03 \text{ kg/m}^3 \)

Therefore \( V = 13449.15 ÷ (1.12 + 0.03) = 11694.9 \text{ m}^3/\text{h} \)
6.4 Energy balance

6.4.1 Assumption for the energy calculations of a cane sugar factory are based on the following assumptions:
Live steam conditions 40 bar, 450°C.
Exhaust steam conditions 3 bar, 133.5°C.

Two sets of knives: 14.5×25 kW ×10.3 kg/kWh = 3,734 kg/h
Shredder: 14.5×30 kW×12.0 kg/kWh = 5,22 kg/h
Five mills: 14.5×5 ×16 kW× 13.1 kg/kWh =15,196 kg/h
Power load: 1,750 × 10.3 kg/kWh = 18,025 kg/h
Auxiliary use: = 3,000 kg/h
Total: = 45175 kg/h

6.4.2 Heater

Assuming the mixed juice from milling tandem is available at 30°C. Before adding lime to juice, the juice has to be heated. Thus the final temperatures to which juices are heated in clarification vary from extremes of 90 to 115°C, although it is common practice to heat slightly above the boiling point. But it is generally believed today that superheating is not advantageous and temperatures just above the boiling point say 103-105°C is the maximum for good practice.

Mixed juice is heated to 103°C in four heaters before clarifier, and it is heated to 125°C in three heaters after clarifier.

Heat required to raise the temperature of mixed juice from \( t_0 \) to \( t_1 \) is given by:
\[
Q = m \times C_p \times \Delta t = m_s \times \lambda
\]

Where: \( X \) is mass fraction of dissolved solids.

Feed to the heater is 246 t /h and % of solids content is 15.45 (from material balance).
Therefore, \(\text{Cp} = 4.19 - 2.35 \times 0.1545 = 3.83 \text{ kJ/kg K}\)

And \(m = 246 \times 1000 = 246000 \text{ kg/h}\)

1- Steam requirement in I.Heater

This heater is heated by 5th vapor coming from V-effect with temperature 87°C, so \(\lambda = 2290.5 \text{ kJ/kg}\)

\(\Delta t = 55 - 35 = 20\degree C\)

Thus from equation (1),

\[Q_1 = 246000 \times 3.83 \times 20 = 18843600 \text{ kW}\]

Steam requirement, \(m_s = \frac{Q_1}{\lambda} = \frac{18843600}{2290.5} = 8226.85 \text{ kg/h}\)

2- Condensate requirement of II.Heater

This heater is heated by condensates coming from all effects with temperature 99.5°C, so \(h_f - h_{f2} = 417 - 377 = 40 \text{ kJ/kg}\)

\(\Delta t = 55 - 80 = 25\degree C\)

Thus from equation (1),

\[Q_2 = 246000 \times 3.83 \times 25 = 23554500 \text{ kW}\]

Condensate requirement, \(m_s = \frac{Q_2}{h_f} = \frac{23554500}{40} = 588862.5 \text{ kg/h}\)

3- Steam requirement in III.Heater

This heater is heated by 4th vapor coming from IV-effect with temperature 99.5°C, so \(\lambda = 2258 \text{ kJ/kg}\)

\(\Delta t = 80 - 90 = 10\degree C\)

Thus from equation (1),

\[Q_3 = 246000 \times 3.83 \times 10 = 9421800 \text{ kW}\]

Steam requirement, \(m_s = \frac{Q_3}{\lambda} = \frac{9421800}{2258} = 4172.63 \text{ kg/h}\)

4- Steam requirement in IV.Heater

This heater is heated by 3rd vapor coming from III-effect with temperature 109.5°C, so \(\lambda = 2232 \text{ kJ/kg}\)

\(\Delta t = 90 - 103 = 13\degree C\)

Thus from equation (1),

\[Q_4 = 246000 \times 3.83 \times 13 = 12248340 \text{ kW}\]
Steam requirement, \( m_s = \frac{Q}{\lambda} = \frac{12248340}{2232} = 5487.61 \text{ kg/h} \).

Amount of clarified juice heated in the second stage = 245.64375 kg/h

% Of solids in this juice = \( \frac{(32.0625 + 5.58125) \times 100}{245.64375} = 15.32 \)

Therefore \( C_p = 4.19 - 2.35 \times 0.1532 = 3.83 \text{ kJ/kg K} \)

5- Steam requirement in V.Heater

This heater is heated by 2\(^{nd}\) vapor coming from II-effect with temperature 118.5\(^\circ\)C, so \( \lambda = 2206 \text{ kJ/kg} \)

\( \Delta t = 103 - 113 = 10^\circ \text{C} \)

Thus from equation (1),

\[ Q_5 = 245643.75 \times 3.83 \times 10 = 9408155.625 \text{ kW} \]

Steam requirement, \( m_v = \frac{Q}{\lambda} = \frac{9408155.625}{2206} = 4264.80 \text{ kg/h} \)

6- Steam requirement in VI.Heater

This heater is heated by 1\(^{st}\) vapor coming from I-effect with temperature 124.5\(^\circ\)C, so \( \lambda = 2189 \text{ kJ/kg} \)

\( \Delta t = 113 - 120 = 7^\circ \text{C} \)

Thus from equation (1),

\[ Q_6 = 245643.75 \times 3.83 \times 7 = 6585708.9375 \text{ kW} \]

Steam requirement, \( m_v = \frac{Q}{\lambda} = \frac{6585708.9375}{2189} = 3008.55 \text{ kg/h} \)

7- Steam requirement in VII.Heater

This heater is heated by exhaust steam with temperature 131.2\(^\circ\)C, so \( \lambda = 2171 \text{ kJ/kg} \)

\( \Delta t = 120 - 125 = 5^\circ \text{C} \)

Thus from equation (1),

\[ Q_7 = 245643.75 \times 3.83 \times 5 = 4704077.8125 \text{ kW} \]

Steam requirement, \( m_v = \frac{Q}{\lambda} = \frac{4704077.8125}{2171} = 2166.78 \text{ kg/h} \)

6.4.3 Quintuple effect evaporator

The evaporator station consists of five bodies.

From material balance
Feed to the first effect = $F = 245.64375$ tons/h

    = 245643.75 kg/h

Fraction of solids in this feed = $X_f = 0.1532$

Liquid output from the last effect = $L_4 = 57893.75$ kg/h

$X_4 = \frac{245643.75 \times 0.1532}{57893.75} = 0.65$

Fraction of solids in this output = $X_4 = 0.65$

Therefore evaporator load = $245643.75 - 57893.75 = 187750$ kg/h

Assume that equal amount of water is vaporized in each effect plus vapor bleeding goes to other processes.

$V_1 = V + 3008.55 + 17260.77$

$V_2 = V + 4264.8$

$V_3 = V + 5487.61$

$V_4 = V + 8345.26$

$V_5 = V + 8638.19$

Therefore $V_1 + V_2 + V_3 + V_4 + V_5 = 5V + 8638.19 + 8345.26 + 5487.61 + 4264.8 + 3008.55 + 17260.77 = 187750$ kg/h

$V = 28148.964$ kg/h

Material balance in each effect is as follows:

I-Effect

Liquid outlet from first effect = $L_1 = F - V_1$

    = $245643.75 - 48418 = 197225.5$ kg/h

Now the solid balance is $F \times X_f = L_1 \times X_1$

Therefore, $X_1 = \frac{(245643.75 \times 0.1532)/197225.5}{0.1908}$

In this liquid outlet, the solids per 100 parts of water = $19.08 \times 100$ / $(100 - 19.08) = 23.58$

II-Effect

Liquid outlet from second effect = $L_2 = L_1 - V_2$

    = $197225.5 - 32413.764 = 164811.736$ kg/h

Now the solid balance is $L_1 \times X_1 = L_2 \times X_2$
Therefore, \( X_2 = \frac{197225.5 \times 0.1908}{164811.736} = 0.2283 \)

In this liquid outlet, the solids per 100 parts of water = \( \frac{22.83 \times 100}{100 – 22.83} = 29.58 \)

III-Effect

Liquid outlet from third effect = \( L_3 = L_2 – V_3 \)
\[ \quad = 164811.736 - 33636.574 = 131175.162 \text{ kg/s} \]

Now the solid balance is, \( L_2 \times X_2 = L_3 \times X_3 \)
Therefore, \( X_3 = \frac{164811.736 \times 0.2283}{131175.162} = 0.2868 \)

In this liquid outlet, the solids per 100 parts of water = \( \frac{28.68 \times 100}{100 – 28.68} = 40.21 \)

IV-Effect

Liquid outlet from fourth effect = \( L_4 = L_3 – V_4 \)
\[ \quad = 131175.162 - 36494.224 = 94680.938 \text{ kg/s} \]

Now the solid balance is, \( L_3 \times X_3 = L_4 \times X_4 \)
Therefore, \( X_4 = \frac{131175.162 \times 0.2868}{94680.938} = 0.3973 \)

In this liquid outlet, the solids per 100 parts of water = \( \frac{39.73 \times 100}{100 – 39.73} = 65.92 \)

V-Effect

Liquid outlet from fifth effect = \( L_5 = L_4 – V_5 \)
\[ \quad = 94680.938 – 36787.154 = 57893.75 \text{ kg/s} \]

Now the solid balance is, \( L_4 \times X_4 = L_5 \times X_5 \)
Therefore, \( X_5 = \frac{95443.75 \times 0.3942}{57893.75} = 0.6498 \)

In this liquid outlet, the solids per 100 parts of water = \( \frac{65 \times 100}{100 – 65} = 185. \)

Heat Balance

The heat capacity of sugar solution in each effect is calculated from equation
\[ C_p = 4.19 - 2.35 \times X \text{ (kJ/kg K)} \]
X = Weight fraction of dissolved solids %
C_p = Heat capacity (kJ/kg K)
Boiling point rise C°:
BPR = 2B / (100-B)
B = Brix of sugar solution
Table (6.3) encompasses the heat capacity and boiling point rise of steam at the five effects.

Table (6.3): Heat capacity and boiling point rise.

<table>
<thead>
<tr>
<th>X_f = 0.1532</th>
<th>C_{p_f} = 3.83</th>
<th>BPR = 0.36</th>
</tr>
</thead>
<tbody>
<tr>
<td>X_1 = 0.1908</td>
<td>C_{p_1} = 3.74</td>
<td>BPR_1 = 0.47</td>
</tr>
<tr>
<td>X_2 = 0.2283</td>
<td>C_{p_2} = 3.65</td>
<td>BPR_2 = 0.59</td>
</tr>
<tr>
<td>X_3 = 0.2868</td>
<td>C_{p_3} = 3.52</td>
<td>BPR_3 = 0.80</td>
</tr>
<tr>
<td>X_4 = 0.3973</td>
<td>C_{p_4} = 3.26</td>
<td>BPR_4 = 1.32</td>
</tr>
<tr>
<td>X_5 = 0.65</td>
<td>C_{p_5} = 2.78</td>
<td>BPR_5 = 3.7</td>
</tr>
</tbody>
</table>

Table (6.4): Enthalpy data from steam tables.

<table>
<thead>
<tr>
<th>Temperature (°C)</th>
<th>Enthalpy of Saturated water vapor (h_g: kJ/kg)</th>
<th>Enthalpy of saturated liquid water (h_f:kJ/kg)</th>
<th>Enthalpy λ. (h_{fg} :kJ/kg)</th>
</tr>
</thead>
<tbody>
<tr>
<td>t_{1s} = 131.2</td>
<td>h_{g1} = 2722</td>
<td>h_{f1} = 551</td>
<td>λ_{1s} = 2171</td>
</tr>
<tr>
<td>t_{2s} = 124.5</td>
<td>h_{g2} = 2713</td>
<td>h_{f2} = 524</td>
<td>λ_{2s} = 2189</td>
</tr>
<tr>
<td>t_{3s} = 118.5</td>
<td>h_{g3} = 2704</td>
<td>h_{f3} = 498</td>
<td>λ_{3s} = 2206</td>
</tr>
<tr>
<td>t_{4s} = 109.5</td>
<td>h_{g4} = 2690</td>
<td>h_{f4} = 458</td>
<td>λ_{4s} = 2232</td>
</tr>
<tr>
<td>t_{5s} = 99.5</td>
<td>h_{g5} = 2675</td>
<td>h_{f5} = 417</td>
<td>λ_{5s} = 2258</td>
</tr>
<tr>
<td>t_{6s} = 87</td>
<td>h_{g6} = 2655</td>
<td>h_{f6} = 364.5</td>
<td>λ_{6s} = 2290.5</td>
</tr>
</tbody>
</table>
Table (6.5): Heat capacity of steam in each effect (from steam tables).

<table>
<thead>
<tr>
<th>Temperature in each effect (°C)</th>
<th>Specific heat of steam (kJ/kg K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>$t_1 = 125$</td>
<td>2.12</td>
</tr>
<tr>
<td>$t_2 = 119$</td>
<td>2.088</td>
</tr>
<tr>
<td>$t_3 = 110$</td>
<td>2.05</td>
</tr>
<tr>
<td>$t_4 = 100$</td>
<td>2.01</td>
</tr>
<tr>
<td>$t_5 = 87$</td>
<td>1.975</td>
</tr>
</tbody>
</table>

I-Effect

The heat balance for first effect is as follows:

$m_s \lambda_{1s} + m_F h_f = m_{V1} h_{V1} + m_{L1} h_{L1}$

$\lambda_{1s} = 2171$ kJ/kg

$m_F = 245643.75$ kg/h

Taking reference temperature as 0°C and feed to the first effect is at 125°C. Thus $t_f = 125°C$

$h_f = C_p \times (t_f - 0) = 3.83 \times 125 = 478.75$ kJ/kg

$h_{V1} = h_{g2} + C_p \times BPR_1$

$C_p$ is heat capacity of heating steam at 131.2°C = 2.155 kJ/kg.

$h_{V1} = 2713 + 2.155 \times 0.47 = 2714$ kJ/kg

$h_{L1} = C_p \times t_1 = 3.74 \times 125 = 467.5$ kJ/kg

$m_{L1} = 245643.75 - m_{V1}$

Therefore

$m_s \times 2171 + 245643.75 \times 478.75 = m_{V1} \times 2714 + (245643.75 - m_{V1}) \times 467.5$
\[ 2171 \times m_S - 2246.5 \times m_{V1} = -2763492.1875 \]
\[ m_S - 1.035 \times m_{V1} = -1272.9 \]  

II-Effect

The heat balance for second effect is as follows:

\[ m_{V1} \lambda_{2s} + m_{L1} h_{L1} = m_{V2} h_{v2} + m_{L2} h_{L2} \]
\[ \lambda_{2s} = h_{v1} - h_{f2} = 2714 - 524 = 2190 \text{ kJ/kg} \]
\[ h_{v2} = h_{g3} + C_p t_1 \times \text{BPR}_2 \]

Where: \( C_p \) is heat capacity of steam.

\[ h_{v2} = 2704 + 2.12 \times 0.59 = 2705.25 \text{ kJ/kg} \]
\[ h_{L2} = C_p t_2 = 3.65 \times 119 = 434.35 \text{ kJ/kg} \]
\[ m_{L2} = 245643.75 - m_{V1} - m_{V2} \]

Therefore

\[ m_{V1} \times 2190 + (245643.75 - m_{V1}) \times 467.5 = m_{V2} \times 2705.25 + (245643.75 - m_{V1} - m_{V2}) \times 434.35 \]
\[ m_{V1} - 1.053 \times m_{V2} = -3775.5 \]  

III-Effect

The heat balance for third effect is as follows:

\[ m_{V2} \lambda_{3s} + m_{L2} h_{L2} = m_{V3} h_{v3} + m_{L3} h_{L3} \]
\[ \lambda_{3s} = h_{v2} - h_{f3} = 2705.25 - 498 = 2207.25 \text{ kJ/kg} \]
\[ h_{v3} = h_{g4} + C_p t_2 \times \text{BPR}_3 \]

Where: \( C_p \) is heat capacity of steam.

\[ h_{v3} = 2690 + 2.088 \times 0.80 = 2691.67 \text{ kJ/kg} \]
\[ h_{L3} = C_p t_3 = 3.52 \times 110 = 387.2 \text{ kJ/kg} \]
\[ m_{L3} = 245643.75 - m_{V1} - m_{V2} - m_{V3} \]

Therefore

\[ m_{V2} \times 2207.25 + (245643.75 - m_{V1} - m_{V2}) \times 434.35 = m_{V3} \times 2691.67 + (245643.75 - m_{V1} - m_{V2} - m_{V3}) \times 387.2 \]
\[ m_{V3} = 5025.93 - 0.02 \times m_{V1} + 0.937 \times m_{V2} \]  

-------(3)
IV-Effect
The heat balance for fourth effect is as follows:
\[ mV_3 \lambda_{4s} + mL_3hL_3 = mV_4 hV_4 + mL_4hL_4 \]
\[ \lambda_{4s} = hV_3 - h_{i4} = 2691.67 - 458 = 2233.67 \text{ kJ/kg} \]
\[ hV_4 = hg_5 + Cp_3 \times \text{BPR}_4 \]
Where: \( Cp \) is heat capacity of steam.
\[ hV_4 = 2675 + 2.05 \times 1.32 = 2677.7 \text{ kJ/kg} \]
\[ hL_4 = Cp_4 \times t_4 = 3.26 \times 100 = 326 \text{ kJ/kg} \]
\[ mL_4 = 245643.75 - mV_1 - mV_2 - mV_3 - mV_4 \]
Therefore
\[ mV_3 \times 2233.67 + (245643.75 - mV_1 - mV_2 - mV_3) \times 387.2 = 2677.7 \times mV_4 + \\
(245643.75 - mV_1 - mV_2 - mV_3 - mV_4) \times 326 \]
\[ mV_4 = 6392.57 - 0.026 \times mV_1 - 0.026 \times mV_2 + 0.924 \times mV_3 \]
-------------------(4)
V-Effect
The heat balance for fifth effect is as follows:
\[ mV_4 \lambda_{5s} + mL_4hL_4 = mV_5 hV_5 + mL_5hL_5 \]
\[ \lambda_{5s} = hV_4 - h_{i5} = 2677.7 - 417 = 2260.7 \text{ kJ/kg} \]
\[ hV_5 = hg_6 + Cp_4 \times \text{BPR}_5 \]
Where: \( Cp \) is heat capacity of steam.
\[ hV_5 = 2655 + 1.975 \times 3.7 = 2662.31 \text{ kJ/kg} \]
\[ hL_5 = Cp_5 \times t_5 = 2.78 \times 87 = 241.86 \text{ kJ/kg} \]
\[ mL_5 = 245643.75 - mV_1 - mV_2 - mV_3 - mV_4 - mV_5 \]
Therefore
\[ mV_4 \times 2260.7 + (245643.75 - mV_1 - mV_2 - mV_3 - mV_4) \times 326 = 2662.31 \times mV_5 + \\
(245643.75 - mV_1 - mV_2 - mV_3 - mV_4 - mV_5) \times 241.86 \]
\[ mV_5 = 8539 + 0.90mV_4 - 0.035 \times mV_1 - 0.035 \times mV_2 - 0.035 \times mV_3 \]
-------------------(5)
But \( mV_1 + mV_2 + mV_3 + mV_4 + mV_5 = 187750 \text{ kg/h} \)
-------------------(6)
By solving equations (1) to (6), results obtained are,
\[ mV_1 = 33947.7 \]
\[ m_{V2} = 35820.3 \]
\[ m_{V3} = 37923.8 \]
\[ m_{V4} = 39619.1 \]
\[ m_{V5} = 40438.2 \]
\[ m_S = 33863 \text{ kg/h} \]

Saturated steam required, as calculated before, is
\[ m_S = 33863 \text{ kg/h} \]

Saturation temperature of steam = 131.2°C at which \( \lambda = 2171 \text{ kJ/kg} \)

Therefore heat supplied by this steam = \( m_S \times \lambda \)
\[ = 33863 \times 2171 \]
\[ = 73516573 \text{ kW} \]

Steam economy = amount of water evaporated / evaporator load / steam used
\[ = \frac{187750}{33863} = 5.55 \]

6.4.4 Crystallizer

As already obtained from material balance calculation, crystallizer operates at 580-mmHg vacuum.

At this pressure, boiling point of water is 61.42°C.

But boiling point rise (BPR) is 4°C.

Temperature in crystallizer = 65.42°C

Heat balance in crystallizer can be written as
\[ m_F h_F + m_s \lambda_s = m_L h_L + m_V h_v \]  \hspace{1cm} \text{(1)}

\( F = \) feed rate to crystallizer
\[ = (20.25 + 32.0625 + 5.58125) \times 1000 \]
\[ = 57893.75 \text{ kg/h} \]

\%solids in this concentrated solution = \( \frac{(32.0625 + 5.58125) \times 100}{57.89375} = 65\% \)

\[ C_{pf} = 4.19 - 2.35 \times X \]

Here \( X = 0.65 \)
Therefore, $C_{pf} = 2.6625 \text{ kJ/kg K}$

Feed temperature = $t_f = 55.45^\circ\text{C}$

Taking base temperature as $0^\circ\text{C}$.

But $h_F = C_{pf} \times (t_f - 0) = 2.6625 \times 55.45 = 147.64 \text{ kJ/kg}$

$m_s = \text{steam rate required in kg/h}$

This pan is heated by first vapor coming from I effect with temperature $124.5^\circ\text{C}$, so $\lambda = 2189 \text{ kJ/kg}$

$m_L = \text{output of crystallizer}$

$= 30.353125 + 10.531625 = 40.88475$

$= 40.88475 \times 1000 = 40884.75 \text{ kg/h}$

$C_{pl} = 4.19 - 2.35 \times X$

Where $X = \text{Mass fraction of solids in crystallizer output}$

$= 30.353125 / 40.88475$

$= 0.74$

Thus $C_{pl} = 4.19 - 2.35 \times 0.75 = 2.43 \text{ kJ/kg K}$

Temperature of output of crystallizer = $t_l = 65.42^\circ\text{C}$

But $h_L = C_{pl} \times (t_l - 0) = 2.43 \times 65.42 = 158.8 \text{ kJ/kg}$

$m_V = \text{amount of water evaporated in the crystallizer}$

$= 17.009 \times 1000$

$= 17009 \text{ kg/h}$

$h_v = \lambda \text{ at } 65.42^\circ\text{C} + (\text{BPR } \times \text{Cp steam})$

$\lambda \text{ at } 65.42^\circ\text{C} = 2342.2 \text{ kJ/kg}$

$\text{Cp steam at } 65.42^\circ\text{C} = 1.928 \text{ kJ/kg K}$

Therefore, $h_v = 2342.2 + (4 \times 1.928) = 2349.91 \text{ kJ/kg K}$

Now equation (1) becomes,

$(57893.75 \times 147.64) + (m_s \times 2189) = (40884.75 \times 158.8) + (17009 \times 2342.2)$

Hence, $m_s = 37783833.09 / 2189 = 17260.77 \text{ kg/h}$

Steam to be supplied to the crystallizer, $m_s = 17260.77 \text{ kg/h}$
6.4.5 Centrifugal separator

Feed to the separator = 40.6046 t/h

= 40.88475 \times 1000 = 40884.75 kg/h

For this capacity and for speed of 1500 rpm, the power required for an operation as suggested by Hugot (table 5.6) = 45 kW.

6.4.6 Dryer

Calculation of the heating surface of air heater:

The air heater generally consists of tubes, which are supplied with steam at about 3 bar. The heating surface of this small heat exchanger will be given by

\[
A = \frac{Q}{U \times \{t - (t_1 + t_0)/2\}}
\]

Where, 
- \(A\) = heating surface of air heater (m\(^2\))
- \(Q\) = Quantity of heat to be transmitted in kJ/h
- \(U\) = Heat transfer coefficient in kJ/ (m\(^2\)h°C)
- \(t\) = Temperature of saturated steam employed in °C
- \(t_0\) = temperature of cold air entering = 30°C
- \(t_1\) = temperature of hot air leaving = 50°C

Assume steam is available at 3 bar.

Therefore \(t = 133.5°C\) (from steam table)

As suggested by Hugot [20], \(U = 41.9\ \text{kJ/ (m}^2\text{h°C)}\)

The quantity of heat \(Q\) to be transferred is calculated from the fact that it consists of three principle terms.

(i) The heat necessary to heat the mass of air is given by

\[
q_1 = M \times C \times (t_1 - t_0)
\]

\(M = \) amount of air entering = 7451.69 kg/h

\(C = \) specific heat of air = 1 kJ/kg K

Therefore, \(q_1 = 7451.69 \times 1 \times (50 - 30) = 149754\ \text{kJ/h}\)
(ii) The heat necessary to evaporate the water contained in the sugar is given by

\[ q_2 = m \times [607 + 0.3 \times t_1 - t_0] \]

m = amount of water evaporated = 597.2 kg/h

Thus, \[ q_2 = 498.16 \times [607 + 0.3 \times 50 - 30] \]

\[ = 305372.08 \text{ kcal/h} = 1278531.80 \text{ kJ/h} \]

The quantity \( q_2 \) is the heat corresponding to that given up by the hot air between \( T \) reached at the outlet from the air heater and \( t_1 \), which has not been included in \( q_1 \).

(iii) The heat necessary to heat the vapor contained in the mass \( M \) of air, assumed saturated is given by

\[ q_3 = M \times H_0 \times C \times (t_1 - t_0) \]

\( C = \) specific heat of vapor = 1.99 kJ/kg K

\( H_0 = \) it is expressed as kg of water per kg of air = 0.026

Therefore \[ q_3 = 7451.69 \times 0.026 \times 1.99 \times (50 - 30) \]

\[ = 7706 \text{ kJ/h} \]

The heat lost to ambient air \( Q \) is taken into account by the following expression:

\[ Q = 1.25 \times (q_1 + q_2 + q_3) \]

\[ = 1.25 \times (149754 + 1278531.80 + 7706) \]

\[ = 1794989.75 \text{ kJ/h} \]

Therefore heating surface of heat exchanger will be

\[ A = 1794989.75 \div [41.9 \times \{133.5 - (30 + 50)/2\}] \]

\[ = 458.18 \text{ m}^2 \]

The steam consumption will be

\[ ms = Q \div \lambda \]

where, \( ms = \) steam consumption for air heater (kg/h)

\( \lambda = \) latent heat of steam used =2164.41 kJ/kg

Therefore, \[ ms = 1794989.75 \div 2164.41 = 829.32 \text{ kg/h} \].
Chapter VII
Results and Discussion

7.1 Mathematical modeling and simulation of the evaporator station
To investigate the means leading to energy saving in the cane sugar industry, all the possible factors affecting the energy consumption were taken into consideration. Among the possibilities which can lead to energy saving in the sugar factory is the pressure evaporation which was chosen as a potential for energy saving. It is well known that the application of pressure evaporation will reduce the steam consumption in the factory, but it is associated with certain bad effects such as sucrose hydrolysis, color formation, enhancement of scale formation rate. It was plan to investigate experimentally the bad effects accompanying the application of pressure evaporation and to trade them off against the gain in energy that will be achieved. Because of technical difficulties in the laboratory this part was replaced with computer simulation, which uses regression equations relating the various parameters.
A computer program was developed to calculate the steam consumption in the evaporator station using the programming language FORTRAN, which is a function of concentration, retention time and pH value. With this program it is possible to vary the various parameters and investigate their effects on one another, such as temperature, concentration, retention time, pH, …etc.

7.2 Dynamic mathematical modeling of evaporator
The modeling of an evaporator includes the formulation of total mass and component balances together with an energy balance. The balances are
derived with the whole evaporator as the control volume. The resulting balances are listed below:

**Total material balance:**

\[
\frac{d}{dt} (\rho V) = \rho_{in} F_{in} - \rho_{out} F_{out} - W \tag{1}
\]

**Component balance with respect to water:**

\[
\frac{d}{dt} (X_w \rho V) = X_{w,in} \rho_{in} F_{in} - X_{w,out} \rho_{out} F_{out} - W \tag{2}
\]

**Component balance with respect to dry contents in black liquor:**

\[
\frac{d}{dt} (X_s \rho V) = X_{s,in} \rho_{in} F_{in} - X_{s,out} \rho_{out} F_{out} \tag{3}
\]

**Energy balance:**

\[
\frac{d}{dt} (\rho V H) = \rho_{in} F_{in} H_{in} + Q - \rho_{out} F_{out} H_{out} - W H_{out} \tag{4}
\]

The equations (1)-(4) are the basic four balance equations that can be derived under the assumption that the liquid is well stirred. However, it is convenient to do the following rearrangements of the equations:

The left-hand side of equation (2) can be written as:

\[
\frac{d}{dt} (X_w \rho V) = X_w \frac{d}{dt} (\rho V) - (\rho V) \frac{d}{dt} (X_w) \tag{5}
\]

Inserting equation (1) and (2) to (5) gives:

\[
\frac{dX_w}{dt} = \frac{1}{\rho V} [\rho_{in} F_{in} (X_{w,in} - X_{w,out}) - W (1 - X_{w,out})] \tag{6}
\]

The same procedure can be used for black liquor. The left-hand side of equation (2) can be written as:

\[
\frac{d}{dt} (X_s \rho V) = X_s \frac{d}{dt} (\rho V) - (\rho V) \frac{d}{dt} (X_s) \tag{7}
\]

Inserting equation (1) and (3) to (7) gives:

\[
\frac{dX_s}{dt} = \frac{\rho_{in} F_{in}}{\rho V} (X_{s,in} - X_{s,out}) + W/ (\rho V) X_{s,out} \tag{8}
\]

Similarly, the left-hand side of equation (2) can be written as:

\[
\frac{d}{dt} (\rho V H) = H \frac{d}{dt} (\rho V) - (\rho V) \frac{d}{dt} (H) \tag{9}
\]

Inserting equation (1) and (4) to (9) gives:

\[
\frac{dH}{dt} = \frac{1}{\rho V} [\rho_{in} F_{in} (H_{in} - H_{out}) - W (H_{out} - H_{out}) + Q] \tag{10}
\]

The equations (1) and (10) together with one of (6) or (8) are a lumped model that describes the unsteady-state behaviour of an evaporator.
One example of a lumped model is seen below (using equation (8)). The model results in a set of three ordinary differential equations.

\[
\frac{d}{dt} (\rho V) = \rho_{in} F_{in} - \rho_{out} F_{out} - W
\]

\[
\frac{dX_s}{dt} = \frac{\rho_{in} F_{in}}{(\rho V)} (X_{s,in} - X_{s,out}) + \frac{W}{(\rho V)} X_{s,out}
\]

\[
\frac{dH}{dt} = \frac{1}{(\rho V)} \left[ \rho_{in} F_{in} (H_{in} - H_{out}) - W (H^v_{out} - H_{out}) + Q \right]
\]

The steady-state behaviour can be calculated if the derivatives on the left hand side accumulation terms, are set to zero, resulting in a nonlinear equation system.

Nomenclature:
- \( F \) : volume flow (m\(^3\)/s), \( H \) : enthalpy (kJ/kg)
- \( Q \) : energy flux (kJ/s)
- \( X \) : concentration (kg black liquor/kg)
- \( V \) : volume (m\(^3\))
- \( W \) : mass flow steam out of evaporator (kg/s)
- \( \rho \) : density (kg/ m\(^3\)), \( v \) : vapor.

7.3 How the program works
The program uses the trial and error procedure, where a certain value for steam consumption is used to calculate the other parameters, i.e., concentration, surface area, vapor from body to the next one and so on. Finally the calculated concentration of the solution leaving the last body is compared with the desired concentration of the syrup. If the difference between them is within a certain limit specified by the user, then the assumed value of steam consumption is taken as the required amount of steam needed to achieve the required level of concentration. If not, a new value is assumed for steam consumption. The calculations will be repeated until this criteria is satisfied, i.e. the absolute difference between calculated final concentration and the desired concentration is minimum. To run the
program one will need a FORTRAN compiler and the following data should be entered by the user:

1- The number of evaporators in the set (n).
2- The flow rate of the solution entering the first body (kg/100 kg cane), the temperature (°C), the concentration (%), the purity (%), and the concentration that leaving the last body (%).
3- The temperature of the heating steam (°C).
4- The condition in each body; temperature (°C) in the vapor space, pH, retention time (min).
5- vapor bleeded from each body (kg/100 kg cane).
6- Heat transfer coefficients in each body for heating surface calculation (W/m$^2$ K).

The calculation should converge after certain number of iterations, depending on the tolerance chosen for the calculated and desired concentration.

7.4 Program verification
The computer program which was used for energy consumption calculation was first fed with known data, in order to test its reliability. It gave very accurate results compared with the original data. This verification process proves that the program is reliable, accurate and can be used with great confidence to calculate the energy consumption and the associated parameters. The data according to (Wardbana, 1978) were fed and the result obtained from the program is shown below. According to these data, the steam consumption is 31.09 kg/100kg cane, whereas the value obtained from the program is 31.77 kg/100kg cane, which indicates the correspondence between the two values. Fig (7.1) shows the data used in the verification of program [18].
Fig (7.1): Data used in program verification.

7.5 Investigation of the steam consumption via simulation program

The program is used for three different cases; in each one there are four various configurations between application of high pressure evaporation and a modified arrangement for vapor bleeding, which is used for heating the heaters and pans. The study case was Alguneid Sugar Factory with the existing working conditions in the evaporator station. It is assumed that all conditions in the evaporators are constant, except the temperature which will be changed; the pH=7, and the retention time = 5 minutes. All evaporators are considered as rising film type, whereas the maximum retention time is 5 minutes as reported by (Bhagat.J.J)[21]. Although there are other important negative effects as color development and scale formation, and as it is difficult to qualify these effects, emphasis is made on the other parameters. The program is used to calculate the amount of
steam consumption in the evaporator station and the amount of the hydrolyzed sugar, so only the remaining parameters are taken into consideration for the evaluation purpose.

7.5.1 - **Case 1**

**Configuration I**

This is taken as the base case, with the same operating conditions as in Alguneid factory which is as follows:

1- **Heating**

There are 5 heaters before clarifier which raise the temperature of juice to 105°C as following:

First heater heats the juice from 30°C to 45°C.
Second heater heats the juice from 45°C to 60°C.
Third heater heats the juice from 60°C to 80°C.
Fourth heater heats the juice from 80°C to 95°C.
Fifth heater heats the juice from 95°C to 105°C.

Two heaters after clarifier are to raise the temperature of juice to the temperature corresponding to that in the first body as follows:

Sixth heater heats the juice from 105°C to 110°C.
Seventh heater heats the juice from 110°C to 115°C.

Mixed juice is heated to 105°C in five heaters before clarifier, and it is heated to 115°C in two heaters after clarifier.

Heat required to raise the temperature of raw juice from \( t_0 \) to \( t_1 \) is given by the following equation:

\[
Q = m \times C_p \times \Delta t = m_s \times \lambda
\]

Where, \( Q \) = heat required (kW).
\( m \) = mass flow of juice (kg/h).
\( m_s \) = steam required (kg/h).
\( \lambda \) = latent heat of steam (kJ/kg).
Cp = heat capacity of mixed juice = 4.19 – 2.35 × X (kJ/kg K)
Where: X is mass fraction of dissolved solids.
Feed to the first heater is 174.51ton /h and % of solids content is 16.24 % (from material balance).
Therefore Cp = 4.19 – 2.35 × 0.1624 = 3.81 kJ/kg K
And m = 174.51 × 1000 = 174510 kg/h
I.Heater:
This heater is heated by vapor coming from III-effect at temperature 88°C, so λ=2288 kJ/kg
Δt = 45 – 30 = 15°C
Thus from equation (1),
Q₁ = 174510 × 3.81 × 15 = 9973246.5 kW
Steam requirement, mₛ= Q₁/ λ= 9973246.5 /2286 = 4358.9 kg/h
II.Heater:
This heater is heated by vapor coming from II-effect at temperature 100°C, so λ=2256 kJ/kg
ΔT = 60 – 45 = 15°C
Thus from equation (1),
Q₂ = 174510× 3.81 × 15 = 9973246.5 kW
Steam requirement, mₛ= Q₂/ λ= 9973246.5 /2256 = 4420.77 kg/h
III.Heater:
This heater is heated by vapor coming from I-effect at temperature 115°C, so λ=2216 kJ/kg
Δt = 80 – 60 = 20°C
Thus from equation (1),
Q₃ = 174510× 3.81 × 20 = 13297662 kW
Steam requirement, mₛ= Q₃/ λ= 13297662/2216= 6000.75 kg/h
IV.Heater:
This heater is heated by exhaust steam at temperature 124°C, so \( \lambda = 2192 \) kJ/kg
\[ \Delta t = 95 - 80 = 15^\circ C \]
Thus from equation (1),
\[ Q_4 = 174510 \times 3.81 \times 15 = 9973246.5 \text{ kW} \]
Steam requirement, \( m_s = \frac{Q}{\lambda} = \frac{9973246.5}{2192} = 4549.84 \text{ kg/h} \]

V. Heater:
This heater is heated by exhaust steam at temperature 124°C, so \( \lambda = 2192 \) kJ/kg
\[ \Delta t = 105 - 95 = 10^\circ C \]
Thus from equation (1),
\[ Q_5 = 174510 \times 3.81 \times 10 = 6648831 \text{ kW} \]
Steam requirement, \( m_s = \frac{Q}{\lambda} = \frac{6648831}{2192} = 3033.23 \text{ kg/h} \)

Amount of clarified juice heated in the second stage = 174.510 kg/h
% of solids in this juice = \( \frac{(24.18 + 3.95) \times 100}{174.51} = 16.12\% \)
Therefore, \( C_p = 4.19 - 2.35 \times 0.1612 = 3.81 \text{ kJ/kg K} \).

VI. Heater:
This heater is heated by exhaust steam at temperature 124°C, so \( \lambda = 2192 \) kJ/kg
\[ \Delta t = 110 - 105 = 5^\circ C \]
Thus from equation (1),
\[ Q_6 = 174510 \times 3.81 \times 5 = 3324415.5 \text{ kW} \]
Steam requirement, \( m_s = \frac{Q}{\lambda} = \frac{3324415.5}{2192} = 1516.61 \text{ kg/h} \)

VII. Heater:
This heater is heated by exhaust steam at temperature 124°C, so \( \lambda = 2192 \) kJ/kg
\[ \Delta t = 115 - 110 = 5^\circ C \]
Thus from equation (1),
\[ Q_7 = 174510 \times 3.81 \times 5 = 3324415.5 \text{ kW} \]
Steam requirement, \( m_s = \frac{Q}{\lambda} = \frac{3324415.5}{2192} = 1516.61 \text{ kg/h} \)

Table (7.1): Heat circuit with steam required (Case 1-Configuration I).

<table>
<thead>
<tr>
<th>Heater</th>
<th>Heating range</th>
<th>Heating agent</th>
<th>Temperature of heating agent</th>
<th>Q kg/100kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>30 – 45 °C</td>
<td>3rd vapor</td>
<td>88 °C</td>
<td>2.1795</td>
</tr>
<tr>
<td>2</td>
<td>45 – 60 °C</td>
<td>2nd vapor</td>
<td>100 °C</td>
<td>2.2104</td>
</tr>
<tr>
<td>3</td>
<td>60 – 80 °C</td>
<td>1st vapor</td>
<td>115 °C</td>
<td>3.0004</td>
</tr>
<tr>
<td>4</td>
<td>80 – 95 °C</td>
<td>Exhaust steam</td>
<td>124 °C</td>
<td>2.2749</td>
</tr>
<tr>
<td>5</td>
<td>95 – 105 °C</td>
<td>Exhaust steam</td>
<td>124 °C</td>
<td>1.5166</td>
</tr>
<tr>
<td>6</td>
<td>105 – 110 °C</td>
<td>Exhaust steam</td>
<td>124 °C</td>
<td>0.758</td>
</tr>
<tr>
<td>7</td>
<td>110 – 115 °C</td>
<td>Exhaust steam</td>
<td>124 °C</td>
<td>0.758</td>
</tr>
</tbody>
</table>

2- Evaporation
The juice is concentrated from 16 to 60°Bx in four bodies of evaporators. Exhaust steam is used for heating the first body, the temperatures in each body is as shown in table (7.2), (F= 87.255 kg/100kg). Fraction of solids in this feed = \( X_f = 0.1612 \)
Purity of clear juice=85.96 %

Table (7.2): Temperature in each effect (Case 1).

<table>
<thead>
<tr>
<th>effect</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
</tr>
</thead>
<tbody>
<tr>
<td>t of heating agent (°C)</td>
<td>124</td>
<td>115</td>
<td>100</td>
<td>88</td>
</tr>
<tr>
<td>t in body (°C)</td>
<td>115</td>
<td>100</td>
<td>88</td>
<td>60</td>
</tr>
</tbody>
</table>

3- Crystallization
The pan is heated by vapor coming from I-effect at temperature 115°C.
Steam to be supplied to the crystallizer, \( ms = 16053.56 \text{ kg/h} \).
For 100 kg cane, then \( m_s = 8.02678 \text{ kg/100 kg} \).

**Configuration II**

All conditions are as in configuration I except that the pan is heated by vapor coming from II-effect at temperature 100°C instead of vapor from I-effect.

Steam to be supplied to the crystallizer, \( ms = 15768.9 \text{ kg/h} \).
For 100 kg cane then \( m_s = 7.8844 \text{ kg/100 kg} \).

**Configuration III**

Calculations of evaporators with a new heat circuit in the factory are the following:

1- Heating

---

**Table (7.3): New heat circuit with steam required (Case I - Configuration III).**

<table>
<thead>
<tr>
<th>Heater no</th>
<th>Heating range</th>
<th>Heating agent</th>
<th>Temperature of heating agent</th>
<th>Q kg/100kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>30 – 45 °C</td>
<td>4\textsuperscript{th} vapor</td>
<td>60 °C</td>
<td>2.1148</td>
</tr>
<tr>
<td>2</td>
<td>45 – 60 °C</td>
<td>Condensates</td>
<td>88°C</td>
<td>_</td>
</tr>
<tr>
<td>3</td>
<td>60 – 80 °C</td>
<td>3\textsuperscript{rd} vapor</td>
<td>88 °C</td>
<td>2.9059</td>
</tr>
<tr>
<td>4</td>
<td>80 – 95 °C</td>
<td>2\textsuperscript{nd} vapor</td>
<td>100 °C</td>
<td>2.21038</td>
</tr>
<tr>
<td>5</td>
<td>95 – 105 °C</td>
<td>1\textsuperscript{st} vapor</td>
<td>115 °C</td>
<td>1.50018</td>
</tr>
<tr>
<td>6</td>
<td>105 – 110 °C</td>
<td>Exhaust steam</td>
<td>124 °C</td>
<td>0.758</td>
</tr>
<tr>
<td>7</td>
<td>110 – 115 °C</td>
<td>Exhaust steam</td>
<td>124 °C</td>
<td>0.758</td>
</tr>
</tbody>
</table>

---

2- Evaporation:

The juice is concentrated from 16 to 60°Bx in four bodies of evaporators. Exhaust steam is used for heating the first body, the temperatures in each body is as shown in table (7.2), (\( F = 87.255 \text{ kg/100 kg} \)). Fraction of solids in this feed = \( X_f = 0.1612 \).
Purity of clear juice = 85.96%.

3- Crystallization

The pan is heated by vapor coming from I-effect at temperature 115°C. Steam to be supplied to the crystallizer, \( m_s = 16053.56 \text{ kg/h} \)

For 100kg cane then \( m_s = 8.02678 \text{ kg/100kg} \)

**Configuration IV**

All conditions are as in configuration III except that the pan is heated by vapor coming from II-effect at temperature 100°C instead of vapor from I-effect.

Steam to be supplied to the crystallizer, \( m_s = 15768.9 \text{ kg/h} \)

For 100kg cane, then \( m_s = 7.8844 \text{ kg/100kg} \)

Table (7.4): Results of case 1.

<table>
<thead>
<tr>
<th>Configuration</th>
<th>Steam requirement in evaporator station kg/100kg cane</th>
<th>Difference in steam consumption from base case kg/100kg cane</th>
</tr>
</thead>
<tbody>
<tr>
<td>Base case I</td>
<td>30.851</td>
<td>0</td>
</tr>
<tr>
<td>II</td>
<td>28.753</td>
<td>2.098</td>
</tr>
<tr>
<td>III</td>
<td>26.057</td>
<td>4.794</td>
</tr>
<tr>
<td>IV</td>
<td>23.988</td>
<td>6.863</td>
</tr>
</tbody>
</table>
7.5.2 - **Case 2**

**Configuration I**

Calculations of evaporators with a new application of pressure evaporation in the factory as following:

1- Heating

Table (7.5): Heat circuit and steam required (Case 2- Configuration I).

<table>
<thead>
<tr>
<th>Heater no</th>
<th>Heating range</th>
<th>Heating agent</th>
<th>Temperature of heating agent</th>
<th>Q kg/100kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>30 – 45 °C</td>
<td>3rd vapor</td>
<td>95 °C</td>
<td>2.1977</td>
</tr>
<tr>
<td>2</td>
<td>45 – 60 °C</td>
<td>2nd vapor</td>
<td>110 °C</td>
<td>2.2362</td>
</tr>
<tr>
<td>3</td>
<td>60 – 80 °C</td>
<td>1st vapor</td>
<td>120 °C</td>
<td>3.0195</td>
</tr>
<tr>
<td>4</td>
<td>80 – 95 °C</td>
<td>Exhaust steam</td>
<td>127 °C</td>
<td>2.2843</td>
</tr>
<tr>
<td>5</td>
<td>95 – 105 °C</td>
<td>Exhaust steam</td>
<td>127 °C</td>
<td>1.5229</td>
</tr>
<tr>
<td>6</td>
<td>105– 110 °C</td>
<td>Exhaust steam</td>
<td>127 °C</td>
<td>0.7614</td>
</tr>
<tr>
<td>7</td>
<td>110– 120 °C</td>
<td>Exhaust steam</td>
<td>127 °C</td>
<td>1.5229</td>
</tr>
</tbody>
</table>

2- Evaporation

The juice is concentrated from 16 to 60°Bx in four bodies of evaporators. Exhaust steam is used for heating the first body, the temperatures in each body is as shown in table (7.6), (F= 87.255 kg/100kg). Fraction of solids in this feed = \(X_f = 0.1612\)

Purity of clear juice=85.96 %

The temperature in each body is calculated with the assumption that the heat transfer area is equal in all four bodies as follows:
Table (7.6): Temperature in each effect (Case 2).

<table>
<thead>
<tr>
<th>effect</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
</tr>
</thead>
<tbody>
<tr>
<td>t of heating agent (°C)</td>
<td>127</td>
<td>120</td>
<td>110</td>
<td>95</td>
</tr>
<tr>
<td>t in body (°C)</td>
<td>120</td>
<td>110</td>
<td>95</td>
<td>75</td>
</tr>
</tbody>
</table>

3- Crystallization

The pan is heated by vapor coming from I-effect at temperature 120°C.
Steam to be supplied to the crystallizer, \( m_s = 16155.63 \text{ kg/h} \)
For 100kg cane, then \( m_s = 8.0778 \text{ kg/100kg} \)

Configuration II

All conditions are as in configuration I except that the pan is heated by vapor coming from II-effect at temperature 110°C instead of vapor from I-effect.
Steam to be supplied to the crystallizer, \( m_s = 15952 \text{ kg/h} \)
For 100kg cane, then \( m_s = 7.976 \text{ kg/100kg} \)

Configuration III

Calculations of evaporators with the assuming conditions for a new heat circuit in the factory are as following:

1- Heating
Table (7.7): Heat circuit and steam required (Case 2- Configuration III).

<table>
<thead>
<tr>
<th>Heater no</th>
<th>Heating range</th>
<th>Heating agent</th>
<th>Temperature of heating agent</th>
<th>Q kg/100kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>30 – 45 °C</td>
<td>4th vapor</td>
<td>75 °C</td>
<td>2.149</td>
</tr>
<tr>
<td>2</td>
<td>45 – 60 °C</td>
<td>Condensates</td>
<td>95 °C</td>
<td>_</td>
</tr>
<tr>
<td>3</td>
<td>60 – 80 °C</td>
<td>3rd vapor</td>
<td>95 °C</td>
<td>2.93</td>
</tr>
<tr>
<td>4</td>
<td>80 – 95 °C</td>
<td>2nd vapor</td>
<td>110 °C</td>
<td>2.236</td>
</tr>
<tr>
<td>5</td>
<td>95 – 105 °C</td>
<td>1st vapor</td>
<td>120 °C</td>
<td>1.5097</td>
</tr>
<tr>
<td>6</td>
<td>105 – 110 °C</td>
<td>Exhaust steam</td>
<td>127 °C</td>
<td>0.7614</td>
</tr>
<tr>
<td>7</td>
<td>110 – 120 °C</td>
<td>Exhaust steam</td>
<td>127 °C</td>
<td>1.5229</td>
</tr>
</tbody>
</table>

2- Evaporation
The juice is concentrated from 16 to 60°Bx in four bodies of evaporators. Exhaust steam is used for heating the first body, the temperatures in each body is as shown in table (7.6), (F= 87.255 kg/100kg). Fraction of solids in this feed = \( X_f = 0.1612 \)
Purity of clear juice = 85.96 %.

3- Crystallization
The pan is heated by vapor coming from I-effect at temperature 120°C. Steam to be supplied to the crystallizer, \( m_s = 16155.63 \) kg/h
For 100kg cane, then \( m_s = 8.0778 \) kg/100kg

Configuration IV
All conditions are as above in configuration III except that the pan is heated by vapor coming from II-effect at temperature 110°C.
Steam to be supplied to the crystallizer, \( m_s = 15952 \) kg/h
For 100kg cane, then \( m_s = 7.976 \) kg/100kg
Table (7.8): Results of case 2.

<table>
<thead>
<tr>
<th>Configuration</th>
<th>Steam requirement in evaporator station kg/100kg cane</th>
<th>Difference in steam consumption from base case kg/100kg cane</th>
</tr>
</thead>
<tbody>
<tr>
<td>I</td>
<td>32.036</td>
<td>-1.185</td>
</tr>
<tr>
<td>II</td>
<td>30.004</td>
<td>0.847</td>
</tr>
<tr>
<td>III</td>
<td>27.240</td>
<td>3.611</td>
</tr>
<tr>
<td>IV</td>
<td>25.202</td>
<td>5.649</td>
</tr>
</tbody>
</table>

7.5.3 **Case 3**

Configuration I

Calculations of evaporators with a new application of pressure evaporation in the factory as following:

1- Heating:

Table (7.9): Heat circuit and steam required (Case 3- Configuration I).

<table>
<thead>
<tr>
<th>Heater no</th>
<th>Heating range</th>
<th>Heating agent</th>
<th>Temperature of heating agent</th>
<th>Q kg/100kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>30 – 45 °C</td>
<td>3rd vapor</td>
<td>100 °C</td>
<td>2.210385</td>
</tr>
<tr>
<td>2</td>
<td>45 – 60 °C</td>
<td>2nd vapor</td>
<td>113 °C</td>
<td>2.245215</td>
</tr>
<tr>
<td>3</td>
<td>60 – 80 °C</td>
<td>1st vapor</td>
<td>125 °C</td>
<td>3.04016</td>
</tr>
<tr>
<td>4</td>
<td>80 – 95 °C</td>
<td>Exhaust steam</td>
<td>130 °C</td>
<td>2.293755</td>
</tr>
<tr>
<td>5</td>
<td>95 – 105 °C</td>
<td>Exhaust steam</td>
<td>130 °C</td>
<td>1.52917</td>
</tr>
<tr>
<td>6</td>
<td>105 – 115 °C</td>
<td>Exhaust steam</td>
<td>130 °C</td>
<td>1.52917</td>
</tr>
<tr>
<td>7</td>
<td>115 – 125 °C</td>
<td>Exhaust steam</td>
<td>130 °C</td>
<td>1.52917</td>
</tr>
</tbody>
</table>
2- Evaporation

The juice is concentrated from 16 to 60°Bx in four bodies of evaporators. Exhaust steam is used for heating the first body, the temperatures in each body is as shown in table (7.8), (F= 87.255 kg/100kg). Fraction of solids in this feed = X_f = 0.1612

Purity of clear juice=85.96 %

The temperature in each body is calculated with the assumption that the heat transfer area is equal in all four bodies which as follows:

<table>
<thead>
<tr>
<th>No of effect</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
</tr>
</thead>
<tbody>
<tr>
<td>t of heating agent (°C)</td>
<td>130</td>
<td>125</td>
<td>115</td>
<td>100</td>
</tr>
<tr>
<td>t in body (°C)</td>
<td>125</td>
<td>115</td>
<td>100</td>
<td>85</td>
</tr>
</tbody>
</table>

3- Crystallization

The pan is heated by vapor coming from I-effect at temperature 125°C.

Steam to be supplied to the crystallizer, ms = 16266.42kg/h

For 100kg cane, then m_s = 8.13321 kg/100kg

Configuration II

All conditions are as in configuration I except that the pan is heated by vapor coming from II-effect at temperature 115°C instead of vapor from I-effect.

Steam to be supplied to the crystallizer, ms = 16053.56 kg/h

For 100kg cane, then m_s = 8.02678 kg/100kg

Configuration III

Calculations of evaporators with the assuming conditions for a new heat circuit in the factory as following:

1- Heating
Table (7.11): Heat circuit and steam required (Case 3- Configuration III).

<table>
<thead>
<tr>
<th>Heater no</th>
<th>Heating range</th>
<th>Heating agent</th>
<th>Temperature of heating agent</th>
<th>Q kg/100kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>30 – 45 °C</td>
<td>4\textsuperscript{th} vapor</td>
<td>85 °C</td>
<td>2.1728</td>
</tr>
<tr>
<td>2</td>
<td>45 – 60 °C</td>
<td>Condensates</td>
<td>100°C</td>
<td>_</td>
</tr>
<tr>
<td>3</td>
<td>60 – 80 °C</td>
<td>3\textsuperscript{rd} vapor</td>
<td>100°C</td>
<td>2.947</td>
</tr>
<tr>
<td>4</td>
<td>80 – 95 °C</td>
<td>2\textsuperscript{nd} vapor</td>
<td>115 °C</td>
<td>2.25</td>
</tr>
<tr>
<td>5</td>
<td>95 – 105 °C</td>
<td>1\textsuperscript{st} vapor</td>
<td>125 °C</td>
<td>1.52</td>
</tr>
<tr>
<td>6</td>
<td>105– 115 °C</td>
<td>Exhaust steam</td>
<td>130°C</td>
<td>1.52917</td>
</tr>
<tr>
<td>7</td>
<td>115– 125 °C</td>
<td>Exhaust steam</td>
<td>130°C</td>
<td>1.52917</td>
</tr>
</tbody>
</table>

2- Evaporation
The juice is concentrated from 16 to 60°Bx in four bodies of evaporators. Exhaust steam is used for heating the first body, the temperatures in each body is as shown in table (7.8), (F= 87.255 kg/100kg). Fraction of solids in this feed = \( X_f = 0.1612 \)
Purity of clear juice=85.96 %.

3- Crystallization
The pan is heated by vapor coming from I-effect at temperature 125°C.
Steam to be supplied to the crystallizer, \( m_s = 16266.42 \text{kg/h} \)
For 100kg cane, then \( m_s = 8.13321 \text{kg/100kg} \)

Configuration IV
All conditions as above in configuration III except that the pan is heated by vapor coming from II-effect at temperature 115°C instead of first vapor from I-effect.
Steam to be supplied to the crystallizer, \( m_s = 16053.56 \text{kg/h} \)
For 100kg cane, then \( m_s = 8.02678 \text{kg/100kg} \)
Table (7.12): Results of case 3.

<table>
<thead>
<tr>
<th>Configuration</th>
<th>Steam requirement in evaporator station kg/100kg cane</th>
<th>Difference in steam consumption from base case kg/100kg cane</th>
</tr>
</thead>
<tbody>
<tr>
<td>I</td>
<td>32.933</td>
<td>-2.082</td>
</tr>
<tr>
<td>II</td>
<td>30.880</td>
<td>-0.029</td>
</tr>
<tr>
<td>III</td>
<td>28.226</td>
<td>2.625</td>
</tr>
<tr>
<td>IV</td>
<td>26.168</td>
<td>4.683</td>
</tr>
</tbody>
</table>
7.6 Discussion of results

Calculation results are shown in table (7.4) for the four configurations in the quadruple effect of evaporation system which is used in Alguneid Sugar Factory (heating steam 124°C, 2.2bar). In configuration I, which is taken as a base case, vapor from the first body will be supplied to pans as well as to the 2nd body and for heating mixed juice III.heater. Vapor from 2nd body heats the mixed juice II.heater and also supply the 3rd body, and vapor from 3rd body heats the mixed juice I.heater and supply the 4th body. The steam requirement in this configuration is (30.851) kg/100kg cane.

In configuration II, the only difference from configuration I is that the vapor for the pans is taken from the 2nd body instead of the 1st body. The steam requirement is (28.753) kg/100kg cane, with a difference of (2.098) steam kg/100kg cane in comparison to the base case.

In configuration III, vapor from 1st body will be supplied to pans, 2nd body, and heats mixed juice in V.heater. Vapor from 2nd body heats IV.heater and also supply the 3rd body. Vapor from 3rd body heats III.heater and supply the 4th body, and the vapor from last body heats I.heater. In this configuration the II.heater is heated with condensate. The steam requirement in this configuration is (26.057) kg/100kg cane, with a difference of (4.794) steam kg/100kg cane in comparison to the base case. In configuration IV, all conditions as it is in configuration III except that is the 2nd vapor will be used for heating the pans instead of 1st vapor. The steam requirement in this configuration is (23.988) kg/100kg cane, with a difference of (6.863) steam kg/100kg cane in comparison to the base case. The changes of the steam requirement in the evaporator station in the four configurations of case 1 compared to the base case are shown in fig (7.2).
Fig (7.2): Steam requirement in the four configurations of case 1 compared to the base case.
Pressure evaporation i.e., using high temperature, is applied to all the above mentioned configurations, where heating steam of properties 127°C, 2.45 bar in case 2 and 130°C, 2.7 bar in case 3 are used. The result obtained are shown in table (7.8) for case 2 and table (7.12) for case 3 together, with the difference in steam consumption compared the base case of Alguneid Sugar Factory. The changes of the steam requirement in the evaporator station in the four configurations of case 2 and case 3 compared to the base case are shown in fig (7.3) and fig (7.4).

It is noticed from the above results that, in the each case, the steam requirement is continuously decreasing from configuration I to IV, it could also be deduced that the best configuration offers more saving of steam, i.e. is that which has a modified heat circuit with using 2\textsuperscript{nd} vapor for heating pans.

It also appears that the application of high temperature and using of 2\textsuperscript{nd} vapor for heating pans will decrease the steam requirement. This is obviously shown in case 2 when t=127°C, 2.45 bar and in table (7.8) configuration II, where the difference is only (0.847) kg steam/100 kg cane from the base case. In case 3 where t=130°C, 2.7 bar it is noticed that is no saving in steam consumption as it is shown in table (7.12) configuration II. Hence, results show that, when pressure evaporation is applied to the different configurations, the steam consumption decreased systematically showing that applying high pressure or temperature will lead to reduction of steam usage.

The results have also indicated that applying pressure evaporation together with rearrangement of the heat circuit has led to improvement in steam consumption as it is shown in configurations II and IV, both of case 2 and case 3.
Fig (7.3): Steam requirement in the four configurations of case 2 compared to the base case.
Fig (7.4): Steam requirement in the four configurations of case 3 compared to the base case.
It is also indicated that the steam requirement can be minimized if a new heat circuit is used with application the same temperature as in the factory; so the maximum amount of steam can be saved is (6.863) kg steam /100kg cane. This amount of steam can save energy in the form of electricity power, also saves amount of bagasse.

How much this amount of steam generates energy, and how much the saving in bagasse will be calculated here using the following equations[24].

\[ G.C.V= 19.259 – 50S – 193W \] 

Where, 

- \( G.C.V \) = grass calorific value of bagasse kJ/kg
- \( S \) = sucrose% on bagasse = 3.63% on bagasse
- \( W \) = moisture in bagasse% = 50.3%

\[ G.C.V= 19.259 - 50 \times 3.63 -193 \times 50.3 = 9273.1 \text{ kJ/kg} \]

\[ WRT = \frac{3.6 \times 10^9}{(H_1-H_2).\eta_{th} \cdot \eta_r \cdot \eta_g} \] 

Where, 

- \( WRT \) = water rate turbine, kg steam/kWh
- \( H_1 \) = adiabatic enthalpy value (live steam) =3150 kJ/kg
- \( H_2 \) = adiabatic enthalpy value (exhaust steam) =2712 kJ/kg
- \( \eta_{th} \) = thermodynamic efficiency of the turbine % =72%
- \( \eta_r \) = efficiency of reduction gear % = 98 %
- \( \eta_g \) = efficiency of generator % =93 %

\[ WRT = \frac{3.6 \times 10^9}{(3150-2712) 72\times98\times93} = 12.53 \text{ kg steam / kWh} \]

Power generation per ton of bagasse =\( G.C.V.\eta_b .10/( H_1-h_f).WRT \) ..... (3) kWh/ton bagasse.

Where, \( \eta_b \) = boiler overall efficiency = 78%

\[ H_1 = \text{adiabatic enthalpy value (live steam) } =3150 \text{ kJ/kg} \]
\[ h_f = \text{enthalpy of feed water } =417 \text{ kJ/kg} \]

Power generation per ton of bagasse = \( 9273.1 \times 78 \times 10 \div (3150-417) \times 12.53\) = 211.23 kWh/ton bagasse.

As calculated before from equation (2) it is obtained that,
12.53 kg steam is needed to generate 1 kWh

Then: 6.863 kg steam/100kg cane will be generate X kWh

\[
X = \frac{6.863 \times 1}{12.53} = 0.55 \text{ kWh/100kg cane}
\]

From equation (3) it is obtained that,

Every 211.23 kWh is generated from 1000 kg of bagasse

Then: 0.55 kWh is generated from Y kg of bagasse

\[
Y = \frac{0.55 \times 1000}{211.23} = 2.6 \text{ kg of bagasse can be saved/100kg cane.}
\]

It can be saved an amount of electricity of 0.55 kWh /0.1ton, and for crushing capacity of 4600t/d \times 210 days = 966000 t/ year.

Then 5313000 kWh/year, which if sold will produce an annual revenue of 106260000 SD, considering that the price of one kWh is 20 SD.
8.1 Conclusion
The sugar industry is a major consumer of steam, with the evaporators and vacuum pans consuming substantial quantities for concentration of juice and manufacture of sugar. Hence, it is considered as a high energy intensive industry with good growth potential.

The sugar industry has a potential for generating excess power, commonly referred to as commercial co-generation, which can be sold to the national grid.

Bagasse has been identified as a cheap source of simultaneous production of power and heat, usually in the form of electricity and steam.

The objective of this work is to investigate the possibilities of saving energy in the cane sugar factory by adopting certain measures, and determine the optimum conditions in the evaporator station. Among the possibilities leading to energy saving is the application of pressure evaporation, which usually reduces the steam consumption in the factory.

To achieve these objectives a computer program was made to calculate the steam consumption in the evaporator station using the programming language FORTRAN, which is a function of concentration, retention time and the pH value. Various parameters such as temperature, concentration, retention time, pH…etc have been introduced and their effects on one another have been investigated. Alguneid Sugar Factory has been taken as a case study.

Results revealed that the cane sugar industry, worldwide, has at last recognized the huge surplus energy potential of bagasse and its impact on industries and on national economy of the country.
Beside the advantages of an optimized maximization of co-generation there is also the use of clean source of energy offered by cane sugar as the world is becoming more and more pollution conscious.

In most sugar factories the evaporation station works as a center of supplying vapors from different effects to juice heaters and vacuum pans. This makes the best use of the steam and reduces its consumption.

Upon analyzing the various configuration in the quadruple effect of evaporation system, it appears that the application of pressure evaporation, combined with the practice of extensive vapor bleeding, and the simultaneous usage of condensate are beneficial for reducing steam consumption.

8.2 Recommendations

The recommendations which can be withdrawn as results of this work and which can lead to minimum energy consumption are:

1. Complete electrification of the sugar factory so as to increase the conversion efficiency of energy e.g replacement of mills turbine with electrical motor.
2. Adoption of continuous processing instead of batch one especially in the crystallization process.
3. Application of pressure evaporation which will make available vapor of high quality (exergy) and which can be used for all heating purposes in the factory and so reducing the usage of exhaust steam.
4. Extensive vapor bleeding from the evaporators station especially from the latest effects.
5. Utilization of the condensate for heating purposes.
6. Arrangement of the heat circuit in the factory so that all heating processes in the factory are carried out using either bled vapor or condensate.
7. Application of computer simulation programs to investigate the consumption and utilization of steam in the sugar factory, which will give quick and more accurate results than normal calculation methods. It also offer the possibility of varying the various parameters and observing their effect on each other.

8. The sugar industry should pay great attention to the co-generation as a new source of income.

9. Improvement of the efficiencies of energy cycle in the factory which includes bagasse burning, steam production and utilization, electricity generation and distribution.
References


14- Ogden, Joan M. Simone Hochgreb and Michael Hylton, Steam Economy and Cogeneration in Cane Sugar Factories. (1999) INT. SUGAR JNL. VOL 92, NO.1099 (P131-139).
Appendix A

Program Used for Calculation

Program evaporator

c Program for the calculation of the steam consumption in the evaporator station
c The user should give the data from key board

c
parameter (m = 8)
character*3 option
dimension hd(m),hb(m),td(m),tb(m),hle(m),hla(m)
dimension hko(m),wtsle(m),wtsla(m),tko(m),t1wte(m)
dimension t1wta(m),md(m),mb(m),t1wtm(m),tla(m),den(m)
dimension mla(m),mle(m),mbes(m),mben(m),ph(m),tz(m)
dimension cko(m),clwt(m),mdgesch(m),wtsm(m),xr(m)
dimension tle(m),mko(m),mkes(m),k(m),eta(m),deltat(m)
dimension deltats(m),wtslm(m),A(m),Q(m),aq(m),tl(m),mu(m)
dimension xxa(m),xxe(m),dts(m),h1kowt(m),tkowt(m),kr(m),dich(m)
Integer wahl,wert
Real  cw,hz,hk,hl,mu,kw1,kw2,kw3
Real  wtsle, wtsla, mle,mla, tle
Real  td, tb,mlwt,wtslwt,mdgesch
Real  hd,hb, hle,hla,hko,hkowt
Real  md,mb,mben,mbes,mkes,kr,mr

c Equations for the calculation of enthalpy of the steam, of the condensate

c and the technical sugar solution

\[
hz(t) = a_0 + a_1 t + a_2 t^2 + a_3 t^3 + a_4 t^4 + a_5 t^5
\]
\[
hk(t) = b_0 + b_1 t + b_2 t^2 + b_3 t^3 + b_4 t^4 + b_5 t^5
\]
\[
hl(t,wts) = (t - 0.183 \cdot (14.155 + 0.60627 \cdot (t - 30) - 0.0009 \cdot (t^2 - 900)) \cdot wts) \cdot 4186.8
\]

c Equations for the calculation of the heat capacity of water and technical sugar
solution

\[
Cw(t) = a_{10} + a_{11} t + a_{12} t^2 + a_{13} t^3 + a_{14} t^4
\]
\[
Cl(t,wts) = c_0 + c_1 \cdot wts + c_2 t + c_3 \cdot wts^2 + c_4 t^2
\]
\* + c_5 t \cdot wts
Calculation of the boiling point elevation of the technical sugar solution
\[ \text{Delts}(wts, t) = \exp(aa0 + aa1 \cdot wts + aa2 \cdot t + aa3 \cdot wts^2 + aa4 \cdot t^2 + aa5 \cdot wts \cdot t + aa6 \cdot wts^3 + aa7 \cdot wts^4 + aa8 \cdot wts^5 + aa9 \cdot wts^6) \]

Dynamic Viscosity of the technical sugar solution
\[ y(wts, t) = \exp(bb0 + bb1 \cdot wts + bb2 \cdot t + bb3 \cdot wts^2 + bb4 \cdot t^2 + bb5 \cdot wts \cdot t + bb6 \cdot wts^3 + bb7 \cdot t^3 + bb8 \cdot t \cdot wts^2 + bb9 \cdot wts^4 + bb10 \cdot wts^5 + bb11 \cdot t^4 + bb12 \cdot t \cdot wts^3 + bb13 \cdot (wts^2) \cdot t^2 + bb14 \cdot wts^6) \]

y(w, t) = exp(g1 \cdot t + g2 \cdot w + g3 \cdot w^2 + g4 \cdot w \cdot t + g5 \cdot t \cdot w^2 + g6)

\[ \text{r}(x) = r1 \cdot x^3 + r2 \cdot x^2 + r3 \cdot x + r4 \]
\[ \text{s}(x) = s1 \cdot x^3 + s2 \cdot x^2 + s3 \cdot x + s4 \]
\[ \text{u}(x) = u1 \cdot x^3 + u2 \cdot x^2 + u3 \cdot x + u4 \]
\[ \text{v}(x) = v1 \cdot x^3 + v2 \cdot x^2 + v3 \cdot x + v4 \]
\[ \text{Roh}(x, t) = r(x) \cdot t + s(x) \cdot t^2 + u(x) \cdot t + v(x) \]
\[ \text{roh}(w, t) = 0.37589318 \cdot t + 342.05688 \cdot w + 130.630299 \cdot w^2 + 0.50738592 \cdot w \cdot t - 0.00811358 \cdot t^2 - 0.34276505 \cdot t \cdot w^2 + 0.00605366 \cdot (w \cdot t)^2 + 990.450799 \]
\[ \text{ppp}(\text{tem}, \text{pH1}, \text{dic}, \text{wt}, \text{f}) = 10^{(16.91 - 5670/(\text{temp} + 273.15) - \text{pH1} + \log_{10}(\text{dic} - \text{dic} \cdot \text{wt} \cdot \text{f})}) \]
\[ \text{xxx}(krv, tzm) = 1 - \exp(-krv \cdot tzm) \]

Enter the temperature and the dry matter content of the entering sugar solution

\[ \text{kw1}(x1) = 230/(x1^{0.305}) \]
\[ \text{kw2}(x1) = 10^{0.01823 \cdot (\log_{10}(x1)) \cdot 3 + 200967 \cdot (\log_{10}(x1))^2} \]
\[ \text{Kw3}(y1, y2) = 5.23E6/(y1^2 + y2^2 + 800) \]
The constants for these equations

Data a0 /2.50107833E6/,a1/1.88112E3/,a2/-1.3910397/
* a3/8.3904895E-3/,a4/-1.01047E-4/,a5/1.91E-7/
Data b0/-192.638655/,b1/4225.60835/,b2/-1.33198879/
* b3/1.5527375E-2/,b4 /-7.1709E-5/,b5/1.63E-7/
data a10/4165.7728/,a11//4458076/,a12/-6.705851E-3/
*a13/8.285391E-5/,a14/-1.020147E-7/
data c0/4185.71556/,c1/-2548.377463/,c2/-2.9312205E-3/
* c3/-12.087166/,c4/4.6306677E-5/,c5/7.539889/
Data aa0/-4.12760644/,aa1/17.337987/,aa2/7.39560963E-3/
Data aa3/-32.6333684/,aa4/1.1190419E-5/,aa5/-4.42240754E-3/
Data aa6/14.194842/,aa7/8.7896804/,aa8/-150.383061/
Data aa9/72.3593418/
Data bb0/-6.652123882/,bb1/5.18843795/,bb2/-2.67818425E-2/
Data bb3/-31.31311659/,bb4/2.94373718E-4/,bb5/-8.23035773E-2/
Data bb6/-8.91693562/,bb7/-2.01169503E-6/,bb8/.1879220322/
Data bb9/-1.096398079E-5/,bb10/27.74397121/,bb11/8.6192291E-9/
Data bb12/-0.345296603/,bb13/1.018113666E-3/,bb14/-2.30314387E-6/
open(7,file='hussein.doc')
1 write(*,*) 'Enter the number of effects of the evaporators n'
Read(*,*)n
write(*,*)'Enter  mass, dry matter content,  temperature and *purity of the solution entering the first body
*mle1, wtsle1,tle1,qr'
read(*,* ) mle(1),wtsle(1), tle(1),qr
write(*,*)'Enter the dry matter content of the syrup  wtsd's'
write(*,*)'wtsdl =,'
read(*,*) wtsdl
c write(*,*)'Geben Sie die Temperatur beim Eintritt und Austritt
* und die Masse der lösung beim Eintritt in den Wärmetauscher ein'
c Write(*,*)'mlwt=',tlwte('1,')=','tlwta('1,')=','wtslwt='
c read(*,* ) mlwt,tlwte(1),tlwta(1),wtslwt
c write(*,*)'Geben Sie die Temperatur von Heizdampf und Brüden in
c *der 1.Stufe ein'
do 11 I = 1,n
write(*,*)'Enter the temperature in the vapor space in each body
* starting from the first body'
write(*,*) 'tb(',i,')=',tb(i),pH(',i,')=',pH(i),tz(',i,')=',tz(i)
Read(*,*)  tb(i),pH(i),tz(i)
11 continue

c      write(*,*)'n=',n,'mle(1)=',mle(1),',wtsdl=',wtsdl,',wtsle(1)='
c     * ,wtsle(1)
c       do 7 i=1,n
c       write(*,*) 'tb(',i,')=',tb(i),'Ph(',i,')=',Ph(i),'tz(',i,')=',tz(i),
c     *  mben(',i,')=',mben(i)
c   7  continue
write(*,*)' if you want to give the amount of vapour bled please
* enter 11 '
read(*,*) test
if(test .eq. 11) then
do 12 i =1,n
write(*,*) 'mben(',i,')='
read(*,*) mben(i)
12 continue
else
do 13 i= 1,n
write(*,*)'Enter the mass and dry matter content of the sugar
* solution entering and leaving the heat exchanger'
c      write(*,*)'mlwt=',mlwt,'wtslw=',wtslw,'
mlwt(',i,')=',',mlwt(',i,')=',
c      mlwt(',i,')=',mlwt(',i,')=',
c      mlwt(',i,')=',mlwt(',i,')=',
c read(*,*) mlwt,wtslw,mlwt(',i,')=mlwt(',i,')=mlwt(',i,')=
13  continue
endif
write(*,*)' if the flash vapour is sent back to the evaporators
* then enter 7'
read(*,*) wert
write(*,*)'Give the factor of heat losses vf=1+v from the
*evaporator'
Read(*,*) vf

C write(*,*) 'tol=',',','tol1='
C read(*,*) tol,tol1
C Start of the trial and error procedure
   e = wtsle(1)+.01
   ff = wtsdl
   Do 37 ws = e,ff,.0001
C Calculation of the first body
   wtsla(1) = ws
   tko(1) = (3*td(1)+tb(1))/4
   tko(2) = (3*td(2)+tb(2))/4
   dts(1) = delts(wtsla(1),tb(1))
   tla(1) = tb(1)+dts(1)
   hd(1) = hz(td(1))
   hko(1) = hk(tko(1))
   hb(1) = hz(tb(1))
   cko(1) = cw(tko(1))
   hle(1) = hl(tle(1),wtsle(1))
   hla(1) = hl(tla(1),Wtsla(1))
   mb(1) = mle(1)*(1-(wtsle(1)/wtsla(1)))
   tlwtm(1) = (tlwte(1) + tlwta(1))/2
   clwt(1) = Cl(tlwtm(1),wtswt)
   mla(1) = mle(1)- mb(1)
C Calculation of the steam consumption in the first body
   md(1)=vf*(mb(1)*hb(1)+mla(1)*hla(1)-mle(1)*hle(1))
   */(hd(1)-hko(1))
   write(*,*) 'md(1)=',md(1)
   write(*,*) 'wtsla(1) =',wtsla(1)
   write(*,*) 'mb(1)=', mb(1)
   if(test .eq.11) then
      goto 88
   else

mben(1) = mlwt*clwt(1)*(tlwta(1)-tlwte(1))
     /(hb(1) -hko(1))

endif

c  Calculation of the estimated mass of heating steam in the second body

mdgesch(2) = mb(1)-mben(1)+mbes(1)

t5 = wtsla(1)
do 27 i = 2,n
     t5 = wtsla(i-1)
     u5 = wtsdl+.0025
c      k(i) = 0

do 17 wtsa = v5, u5,.0001
     wtsla(i) = wtsa
     tko(i) = (3*td(i)+tb(i))/4
     tlwtn(i) = (tlwte(i) + tlwta(i))/2
     td(i) = tb(i-1)
     dts(i) = delts(wtsla(i),tb(i))
     tla(i) = tb(i)+dts(i)
     tle(i) = tla(i-1)
     hb(i) = hz(tb(i))
     hd(i) = hz(td(i))
     mle(2) = mla(1)
     wtsle(2) = wtsla(1)
     mle(i)= mla(i-1)
     wtsle(i)= wtsla(i-1)
     hko(i)= hk(tko(i))
     hla(i)= hl(tla(i),wtsla(i))
     hLe(i)= hl(tle(i),wtsle(i))
c  Calculation of the mass of vapour
     mb(i) = mle(i)*(1 - wtsle(i)/wtsla(i))
     clwt(i)= Cl(tlwtn(i),wtslwt)
     mla(i) = mle(i) - mb(i)
     cko(i) = cw(tko(i))
c  Calculation of the mass of the heating steam up the 2nd body
     md(i) =vf*(mb(i)*hb(i)+mla(i)*hla(i)-mle(i)*hle(i))/
*(hd(i)-hko(i))
mko(i) = md(i)
c Calculation of the mass of the condensates which will be flashed
mbes(1) = .0
mkes(1) = .0
if(i .lt. n) then
    mkes(i) = mkes(i-1)+mko(i)-mbes(i-1)
else
    mkes(i) =0
endif
c k(i) = K(i) + 1
c Calculation of the flashed vapor
mbes(i) = mkes(i)*(tko(i)-tb(i))*cko(i)/(hb(i)-hk(tb(i)))
c Calculation of the bled vapor
tkowt(i) = (6*tb(i)+tlwte(i)+tlwta(i))/8
hkowt(i) = hk(tkowt(i))
if(test .eq. 11) then
goto 99
else
    mben(i) = mlwt*clwt(i)*(tlwta(i)-tlwte(i))/(hb(i)-hko(i))
endif
c Calculation of the mass of estimated heating steam up the 2nd body
99 If(wert .eq. 7) then
    mdgesch(i) = mb(i-1)-mben(i-1)+mbes(i-1)
else
    mdgesch(i) = mb(i-1)-mben(i-1)
endif
c Comparison of the estimated heating steam with calculated one
tol = .1
    IF(ABS(mgesch(i)-md(i)).Le. tol) then
goto 19
endif
17 continue
goto 37
write(*,*) 'md(',i,')=',md(i)
write(*,*) 'mb(',i,')=',mb(i)
write(*,*) 'mla(',i,')=',mla(i)
write(*,*) 'mle(',i,')=',mle(i)
write(*,*) 'wtsle(',i,')=',wtsle(i)
write(*,*) 'wtsla(',i,')=',wtsla(i)
write(*,*) 'mben(',i,')=',mben(i)
write(*,*) 'mbes(',i,')=',mbes(i)
write(*,*) 'mdgesch(',i,')=',mdgesch(i)

continue

c  Comparison of the dry matter content of the syrup leaving the last effect

c  obtained by calculation with the desired ones

tol1 =.005
d=wtsdl+.0025
IF(Abs(d-wtsla(n)).Le.tol1) then
  goto 47
endif

continue

goto 67

47  write(*,*) 'all conditions are satisfied'

write(*,*) 'Enter the crushing capacity of the factory  mr t/h'
read(*,*) mr
f=mr*10/3600
Write(*,*)'Which equation do you want to use for the calculation
*of the heat transfer coefficient U values? enter 1, if you
*want to use  U =230/eta**.305 enter 2 if you want to use
*lg(U)=a.(lg(mu)**3+b.(lg(mu))**2, or enter 3, if you
* want to use U= 5.23E6/(wtsla**2+wtsle**2+800),or Enter  5
*if you want to enter the values of U from key board'
read(*,*)wahl
do 55 j = 1,n
wtslm(j)= (wtsle(j)+ wtsla(j))/2
deltats(j) = delts(wtslm(j),tb(j))
deltat(j)= td(j) - deltats(j)-tb(j)
\( tl(j) = tb(j) + \text{deltats}(j) \)
\( aq(j) = \text{roh}(\text{wtslm}(j), tl(j)) \)
\( \text{eta}(j) = \text{y}(\text{wtslm}(j), tl(j)) \)
\( \text{mu}(j) = \frac{\text{Eta}(j)}{aq(j)} \)

If\\(\text{wahl .eq. 1})\) then
\( k(j) = \text{kw1}(\text{eta}(j)) \)
\( Q(j) = \text{md}(j)\times(\text{hd}(j)-\text{hko}(j)) / \text{vf} \)
\( A(j) = \frac{f\times Q(j)}{\text{deltat}(j)\times k(j)} \)

elseif\\(\text{wahl .eq. 2})\) then
\( k(j) = kw2(\text{mu}(j)) \)
\( Q(j) = \text{md}(j)\times(\text{hd}(j)-\text{hko}(j)) / \text{vf} \)
\( A(j) = \frac{(Q(j)\times f)}{\text{deltat}(j)\times k(j)} \)

elseif\\(\text{wahl .eq. 3})\) then
\( \text{xxa}(j) = 100\times \text{wtsla}(j) \)
\( \text{xxe}(j) = 100\times \text{wtsle}(j) \)
\( k(j) = kw3(\text{xxa}(j), \text{xxe}(j)) \)
\( Q(j) = \text{md}(j)\times(\text{hd}(j)-\text{hko}(j)) / \text{vf} \)
\( A(j) = \frac{(Q(j)\times f)}{\text{deltat}(j)\times k(j)} \)

elseif\\(\text{wahl .eq. 5})\) then
\( \text{write}(*,*) 'geben Sie die k-Werte ein' \)
\( \text{write}(*,*) k('j',')=' \)
\( \text{read}(*,*) k(j) \)
\( Q(j) = \text{md}(j)\times(\text{hd}(j)-\text{hko}(j)) / \text{vf} \)
\( A(j) = \frac{f\times Q(j)}{\text{deltat}(j)\times k(j)} \)
Endif

55 continue

c Print out of the results

\( \text{xxr}=0 \)
\( \text{do 44} i=1, n \)
\( \text{wtsm}(i)= \frac{(\text{wtsle}(i)+\text{wtsla}(i))}{2} \)
\( \text{den}(i)= \text{Roh}(\text{wtsm}(i), \text{tla}(i)) \)
\( \text{dich}(i)= \frac{\text{den}(i)}{1000.0} \)
\( \text{kr}(i)= \text{ppp}(\text{tla}(i), \text{pH}(i), \text{dich}(i), \text{wtsm}(i), \text{qr}) \)
xr(i)= xxx(kr(i),tz(i))
xxr= xxr+xr(i)

44 continue
mz=mle(1)*wtsle(1)*qr
zverlust=xxr*mz*mr*10

do 57 i = 1,n
write(7,'(1x,a8,i1,a,f8.3,a3)')'td(',i,')=',td(i),'C°'
write(7,'(1x,a8,i1,a,f8.3,a3)')'tb(',i,')=',tb(i),'C°'
write(7,'(1x,a8,i1,a,f8.3,a3)')'tla(',i,')=',tla(i),'C°'
do 57 i = 1,n
write(7,'(1x,a8,i1,a,f7.3,a4)')'md(',i,')=',md(i),'kg/dtR'
write(7,'(1x,a8,i1,a,f7.3,a4)')'mb(',i,')=',mb(i),'kg/dtR'
write(7,'(1x,a8,i1,a,f7.3,a4)')'mle(',i,')=',mle(i),'kg/dtR'
write(7,'(1x,a8,i1,a,f7.3,a4)')'mla(',i,')=',mla(i),'kg/dtR'
write(7,'(1x,a8,i1,a,f7.3,a4)')'wtsle(',i,')=',wtsle(i),'g/g'
write(7,'(1x,a8,i1,a,f7.3,a4)')'wtsla(',i,')=',wtsla(i),'g/g'
write(7,'(1x,a8,i1,a,f7.3,a4)')'mben(',i,')=',mben(i),'kg/dtR'
write(7,'(1x,a8,i1,a,f7.3,a4)')'mkes(',i,')=',mkes(i),'kg/dtR'
write(7,'(1x,a8,i1,a,f7.3,a4)')'mbes(',i,')=',mbes(i),'kg/dtR'
write(7,'(1x,a8,i1,a,f7.3,a4)')'mdges(',i,')=',mdgesch(i),'kg/dtR'
write(7,'(1x,a7,i1,a,e10.3,a5)')'k(',i,') =',k(i),'W/m2.K'
Write(7,'(1x,a7,i1,a,e9.3,a6)')'Q(',i,') =',Q(i),' J/dtR'
write(7,'(1x,a7,i1,a,f9.3,a3)')'A(',i,') =',A(i),'m2'
write(7,'(1x,a8,i1,a,f7.3,a2)')'deltats(',i,')=',deltats(i), 'K'
write(7,'(1x,a8,i1,a,f7.3,a2)')'Deltat(',i,')=',deltat(i), 'K'
write(7,'(1x,a8,i1,a,e10.3,a5)')'Eta(',i,')=',eta(i), 'Pa.s'
write(7,'(1x,a8,i1,a,f8.3,a6)')'aq(',i,')=aq(i),'kg/m3'
write(7,'(1x,a8,i1,a,e9.3,a6)')'kr(',i,')=',kr(i),'min-1'
write(7,'(1x,a8,i1,a,e9.3,a4)')'xr(',i,')=',xr(i),'g/g'
write(7,'(1x,a8,i1,a,f6.4,a6)')'dich(',i,')=',dich(i),'g/cm3'
write(7,'(1x,a11,f7.3,a5)')'zverlust=',zverlust,'kg/d'
57 continue
goto 77

67 write(*,*) 'It is impossible to a solution'
77 write(*,*)'Do you want to run the program again:answer yes
*or no'
Read(*,*) option
  if(option .eq. 'yes') then
    goto 1
  endif
stop
end
Appendix B

Output of the simulation program

Cas I
-Configuration I
The crushing capacity of the factory mR = 200.000 t/h
No of bodies = 4
Body (1):
Temperature of steam td (1) = 124.000 °C
Temperature of vapor tb (1) = 115.000 °C
Temperature of vapor tle (1)= 115.000 °C
Temperature of solution leaving the evaporator tla (1) = 115.502 °C
Amount of steam consumption md (1)= 25.543 kg/dtR + 5.3077 = 30.8507

-Configuration II
The crushing capacity of the factory mR = 200.000 t/h
No of bodies = 4
Body (1):
Temperature of steam td (1) = 124.000 °C
Temperature of vapor tb (1) = 115.000 °C
Temperature of solution leaving the evaporator tle (1) = 115.000 °C
Temperature of solution leaving the evaporator tla (1) = 115.479 °C
Amount of steam consumption md (1) = 23.445 kg/dtR + 5.3077 = 28.7527

-Configuration III
The crushing capacity of the factory mR = 200.000 t/h
No of bodies = 4
Stufe(1):
Temperature of steam td (1) = 124.000 °C
Temperature of vapor tb (1) = 115.000 °C
Temperature of solution leaving the evaporator tle (1) = 115.000 °C
Temperature of solution leaving the evaporator tla (1) = 115.491 °C
Amount of steam consumption md (1) = 24.541 kg/dtR + 1.516 = 26.057
-Configuration IV
The crushing capacity of the factory $m_R = 200.000$ t/h
No of bodies = 4
Body (1):
Temperature of steam $td (1) = 124.000$ C°
Temperature of vapor $tb (1) = 115.000$ C°
Temperature of solution leaving the evaporator $tle (1) = 115.000$ C°
Temperature of solution leaving the evaporator $tla (1) = 115.469$ C°
Amount of steam consumption $md (1) = 22.472$ kg/dtR $+1.516 = 23.988$

Case 2
-Configuration I
The crushing capacity of the factory $m_R = 200.000$ t/h
No of bodies = 4
Body (1):
Temperature of steam $td (1) = 127.000$ C°
Temperature of vapor $tb (1) = 120.000$ C°
Temperature of solution leaving the evaporator $tle (1) = 120.000$ C°
Temperature of solution leaving the evaporator $tla (1) = 120.530$ C°
Amount of steam consumption $md (1) = 25.945$ kg/dtR $+6.09565 = 32.0363$

-Configuration II
The crushing capacity of the factory $m_R = 200.000$ t/h
No of bodies = 4
Body (1):
Temperature of steam $td (1) = 127.000$ C°
Temperature of vapor $tb (1) = 120.000$ C°
Temperature of solution leaving the evaporator $tle (1) = 120.000$ C°
Temperature of solution leaving the evaporator $tla (1) = 120.506$ C°
Amount of steam consumption $md (1) = 23.908$ kg/dtR $+6.09565 = 30.00365$

-Configuration III
The crushing capacity of the factory $m_R = 200.000$ t/h
No of bodies = 4
Body (1):
Temperature of steam $td (1) = 127.000$ C°
Temperature of vapor \( tb \) (1) = 120.000 \( ^\circ \text{C} \)
Temperature of solution leaving the evaporator \( tle \) (1) = 120.000 \( ^\circ \text{C} \)
Temperature of solution leaving the evaporator \( tla \) (1) = 120.518 \( ^\circ \text{C} \)
Amount of steam consumption \( md \) (1) = 24.957 kg/dtR +2.2834=27.2404

-Configuration IV
The crushing capacity of the factory \( mR \) = 200.000 t/h
No of bodies = 4
Body (1):
Temperature of steam \( td \) (1) = 127.000 \( ^\circ \text{C} \)
Temperature of vapor \( tb \) (1) = 120.000 \( ^\circ \text{C} \)
Temperature of solution leaving the evaporator \( tle \) (1) = 120.000 \( ^\circ \text{C} \)
Temperature of solution leaving the evaporator \( tla \) (1) = 120.496 \( ^\circ \text{C} \)
Amount of steam consumption \( md \) (1)= 22.919 kg/dtR +2.2834= 25.2024

Case 3
-Configuration I
The crushing capacity of the factory \( mR \) = 200.000 t/h
No of bodies = 4
Body (1):
Temperature of steam \( td \) (1) = 130.000 \( ^\circ \text{C} \)
Temperature of vapor \( tb \) (1) = 125.000 \( ^\circ \text{C} \)
Temperature of solution leaving the evaporator \( tle \) (1) = 125.000 \( ^\circ \text{C} \)
Temperature of solution leaving the evaporator \( tla \) (1) = 125.557 \( ^\circ \text{C} \)
Amount of steam consumption \( md \) (1)= 26.055 kg/dtR +6.881265=32.936265

-Configuration II
The crushing capacity of the factory \( mR \) = 200.000 t/h
No of bodies = 4
Body (1):
Temperature of steam \( td \) (1) = 130.000 \( ^\circ \text{C} \)
Temperature of vapor \( tb \) (1)= 125.000 \( ^\circ \text{C} \)
Temperature of solution leaving the evaporator \( tle \) (1) = 125.000 \( ^\circ \text{C} \)
Temperature of solution leaving the evaporator \( tla \) (1) = 125.532 \( ^\circ \text{C} \)
Amount of steam consumption \( md \) (1) = 24.000 kg/dtR +6.881265= 30.881265
-Configuration III
The crushing capacity of the factory $m_R = 200,000$ t/h
No of bodies = 4
Body (1):
Temperature of steam $t_{d(1)} = 130,000$ C°
Temperature of vapor $t_{b(1)} = 125,000$ C°
Temperature of solution leaving the evaporator $t_{le(1)} = 125,000$ C°
Temperature of solution leaving the evaporator $t_{la(1)} = 125,544$ C°
Amount of steam consumption $m_d(1) = 25.043 \text{ kg/dtR} +3.1834 = 28.2264$

-Configuration IV
The crushing capacity of the factory $m_R = 200,000$ t/h
No of bodies = 4
Body (1):
Temperature of steam $t_{d(1)} = 130,000$ C°
Temperature of vapor $t_{b(1)} = 125,000$ C°
Temperature of solution leaving the evaporator $t_{le(1)} = 125,000$ C°
Temperature of solution leaving the evaporator $t_{la(1)} = 125,520$ C°
Amount of steam consumption $m_d(1) = 22.985 \text{ kg/dtR} +3.1834 = 26.1684$
Appendix C

Equations Used in the Program

Sugar solution

Density

\[ \rho = a.t^3 + b.t^2 + c. t +d \]

\( \rho \): density in kg/m³

\( w_{TS} \): dry matter content ing/100g solution

\( t \) : temperature in ºC

\[ a = a_1. w_{TS}^3 + a_2. w_{TS}^2 + a_3. w_{TS} + a_4 \]

\[ b = b_1. w_{TS}^3 + b_2. w_{TS}^2 + b_3. w_{TS} + b_4 \]

\[ c = c_1. w_{TS}^3 + c_2. w_{TS}^2 + c_3. w_{TS} + c_4 \]

\[ d = d_1. w_{TS}^3 + d_2. w_{TS}^2 + d_3. w_{TS} + d_4 \]

\[ a_1 = -7.225949 .10^{-11} \]

\[ a_2 = 6.36639 .10^{-9} \]

\[ a_3 = -3.55879 .10^{-7} \]

\[ a_4 = 1.31672 .10^{-5} \]

\[ b_1 = 8.44748 .10^{-9} \]

\[ b_2 = -6.50649 .10^{-7} \]

\[ b_3 = 7.64846 .10^{-5} \]

\[ b_4 = -5.44367 .10^{-3} \]

\[ c_1 = -1.54849 .10^{-8} \]

\[ c_2 = -2.63869 .10^{-6} \]

\[ c_3 = -6.85707 .10^{-3} \]

\[ c_4 = -6.01137 .10^{-3} \]

\[ d_1 = 2.69936 .10^{-5} \]

\[ d_2 = 1.46409 .10^{-2} \]

\[ d_3 = 3.94325 \]

\[ d_4 = 1.00045.10^{-3} \]

validity range 0< \( w_{TS} < 69\% \)

10< \( t <130 \) ºC

q = 100 %

Source of material: Schneider et al. (1963)

Boiling point rise

\[ \ln \beta = a_0 + a_1 . w_{TS} + a_2.t + a_3. w_{TS}^2 + a_4. t^2 + a_5. w_{TS}.t + a_6. w_{TS}^3+ a_7. w_{TS}^4+ a_8. w_{TS}^5 + a_9. w_{TS}^6 \]

\( \beta \) = Boiling point rise in ° C
w_{TS}: dry matter content in g/100g solution

\[ t: temperature \in ^\circ C \]

\[ a_0 = -4,12760644 \hspace{1cm} a_8 = -4.42240754 \times 10^{-3} \]
\[ a_1 = 17.337987 \hspace{1cm} a_6 = 14.194842 \]
\[ a_2 = 7.39560963 \times 10^{-3} \hspace{1cm} a_7 = 87.896804 \]
\[ a_3 = 32.6333684 \hspace{1cm} a_8 = -150.383061 \]
\[ a_4 = 1.1190419 \times 10^{-5} \hspace{1cm} a_9 = 72.3593418 \]

validity range \(0 < w_{TS} < 0.8\)

\[ 65 < t < 130 \, ^\circ C \]
\[ q = 90 \% \]


Specific heat capacity

\[ c = a_0 + a_1 \cdot w_{TS} + a_2 \cdot t + a_3 \cdot w_{TS}^2 + a_4 \cdot t^2 + a_5 \cdot w_{TS} \cdot t \]

\[ c: \text{Specific heat capacity in J/kg .K} \]

\[ w_{TS}: \text{dry matter content in g/100g solution} \]

\[ t: \text{temperature in } ^\circ C \]

\[ a_0 = 4185.71556 \hspace{1cm} a_3 = -12.087166 \]
\[ a_1 = -2548.377463 \hspace{1cm} a_4 = 4.6306677 \times 10^{-5} \]
\[ a_2 = -2.9312205 \times 10^{-3} \hspace{1cm} a_5 = 7.539889 \]

validity range \(0 < w_{TS} < 0.7\)

\[ 30 < t < 140 \, ^\circ C \]
\[ q = 90 \% \]


Specific enthalpy

\[ h = \{ t - 0.183 - [14.155 + 0.60627 \cdot (t-30) - 0.0009 \cdot (t^2-900)] \cdot w_{TS}\} \cdot 4186.8 \]

\[ h: \text{Specific enthalpy in J/kg} \]

\[ w_{TS}: \text{dry matter content in g/100g solution} \]

\[ t: \text{temperature in } ^\circ C \]

validity range \(0 < w_{TS} < 0.7\)

\[ 0 < t < 130 \, ^\circ C \]
\[ q = 94.3 \% \]

Thermal conductivity
\[ \lambda = a_0 + a_1 \cdot w_{TS} + a_2 \cdot t + a_3 \cdot w^2_{TS} + a_4 \cdot t^2 + a_5 \cdot w_{TS} \cdot t \]
\( \lambda \): Thermal conductivity in W/m K
\( w_{TS} \): dry matter content in g/100g solution
\( t \): temperature in °C

\( a_0 = 0.568234263 \)
\( a_1 = -0.341773136 \)
\( a_2 = 1.874240163 \cdot 10^{-3} \)
\( a_3 = -3.19310954 \cdot 10^{-4} \)
\( a_4 = -7.622765304 \cdot 10^{-6} \)
\( a_5 = -2.089174088 \cdot 10^{-4} \)

validity range \(0 < w_{TS} < 0.6\)
\(0 < t < 110 \) °C
\( q = 100\% \)


Dynamic viscosity
\[ \ln \eta = a_0 + a_1 \cdot w_{TS} + a_2 \cdot t + a_3 \cdot w^2_{TS} + a_4 \cdot t^2 + a_5 \cdot w_{TS} \cdot t + a_6 \cdot w^3_{TS} + a_7 \cdot t^3 + a_8 \cdot w^2_{TS} \cdot t^2 + a_9 \cdot w_{TS} \cdot t^2 + a_{10} \cdot w^4_{TS} + a_{11} \cdot t^4 + a_{12} \cdot w^3_{TS} \cdot t + a_{13} \cdot w^2_{TS} \cdot t^2 + a_{14} \cdot w_{TS} \cdot t^3 \]
\( \eta \): Dynamic viscosity in Pas
\( w_{TS} \): dry matter content in g/100g solution
\( t \): temperature in °C

\( a_0 = -6.652123882 \)
\( a_1 = 5.18843795 \)
\( a_2 = -2.67818425 \cdot 10^{-2} \)
\( a_3 = 1.21211659 \)
\( a_4 = 2.94272718 \cdot 10^{-4} \)
\( a_5 = -2.23035773 \cdot 10^{-2} \)
\( a_6 = -8.91693562 \)
\( a_7 = -2.01169503 \cdot 10^{-6} \)
\( a_8 = 0.1879220322 \)
\( a_9 = 27.74397121 \)
\( a_{10} = 8.6192291 \cdot 10^{-9} \)
\( a_{11} = -0.3452966035 \)
\( a_{12} = 1.018111366 \cdot 10^{-3} \)
\( a_{13} = -2.30314387 \cdot 10^{-6} \)

validity range \(0 < w_{TS} < 0.8\)
\(0 < t < 80 \) °C
\( q = 90\% \)

Water

Density
\[ \rho = a_0 + a_1 t + a_2 t^2 + a_3 t^3 + a_4 t^4 \]
\( \rho \) : density in kg/m³
\( t \) : temperature in °C

\[ a_0 = 996,571028 \quad a_3 = 4,591676 \times 10^{-5} \]
\[ a_1 = 0,1879371 \quad a_4 = -1,1105351 \times 10^{-7} \]
\[ a_2 = -9,20491 \times 10^{-3} \]
validity range 50< t <150 °C

Specific heat capacity
\[ c = a_0 + a_1 t + a_2 t^2 + a_3 t^3 + a_4 t^4 \]
c : Specific heat capacity in J/kg.K
\( t \) : temperature in °C

\[ a_0 = 4165,7728 \quad a_3 = 8,285391 \times 10^{-5} \]
\[ a_1 = 0,4458076 \quad a_4 = -1,020147 \times 10^{-7} \]
\[ a_2 = -6,705851 \times 10^{-3} \]
validity range 50< t <150 °C

Specific enthalpy
\[ h = a_0 + a_1 t + a_2 t^2 + a_3 t^3 + a_4 t^4 + a_5 t^5 \]
h : Specific enthalpy in J/kg
\( t \) : temperature in °C

\[ a_0 = -192,638655 \quad a_3 = 1,5527375 \times 10^{-2} \]
\[ a_1 = 4225,60835 \quad a_4 = -7,1709 \times 10^{-5} \]
\[ a_2 = -1,33198879 \quad a_5 = 1,63 \times 10^{-7} \]
validity range 20< t <160 °C
Boiling point
\[ t = a_0 + a_1 \cdot k + a_2 \cdot k^2 + a_3 \cdot k^3 + a_4 \cdot k^4 + a_5 \cdot k^5 + a_6 \cdot k^6 + a_7 \cdot k^7 + a_8 \cdot k^8 + a_9 \cdot k^9 + a_{10} \cdot k^{10} + a_{11} \cdot k^{11} \]

\[ k = \ln \left( \frac{p}{0.98067} \right) \]

\[ t \quad : \text{temperature in } ^\circ \text{C} \]
\[ p \quad : \text{pressure in bar} \]

\[ a_0 = 99.0927119 \quad a_6 = -3.73934842 \cdot 10^{-4} \]
\[ a_1 = 27.8542421 \quad a_7 = -1.74177519 \cdot 10^{-5} \]
\[ a_2 = 2.37535764 \quad a_8 = 2.20717117 \cdot 10^{-5} \]
\[ a_3 = 0.210778046 \quad a_9 = 1.53437313 \cdot 10^{-6} \]
\[ a_4 = 2.12968201 \cdot 10^{-2} \quad a_{10} = -4.26856851 \cdot 10^{-7} \]
\[ a_5 = 1.32837729 \cdot 10^{-3} \quad a_{11} = -2.29246029 \cdot 10^{-8} \]

validity range \( 0.02 < p < 5 \) bar


Thermal conductivity
\[ \lambda = a_0 + a_1 \cdot t + a_2 \cdot t^2 + a_3 \cdot t^3 + a_4 \cdot t^4 \]

\[ \lambda \quad : \text{Thermal conductivity in W/m K} \]
\[ t \quad : \text{temperature in } ^\circ \text{C} \]

\[ a_0 = 0.557385464 \quad a_3 = 6.9971321 \cdot 10^{-8} \]
\[ a_1 = 2.428389975 \cdot 10^{-3} \quad a_4 = -1.61136222 \cdot 10^{-10} \]
\[ a_2 = -1,72874778 \cdot 10^{-5} \]

validity range \( 20 < t < 150 \) °C


Dynamic viscosity
\[ \eta = a_0 + a_1 \cdot t + a_2 \cdot t^2 + a_3 \cdot t^3 + a_4 \cdot t^4 \]

\[ \eta \quad : \text{Dynamic viscosity in Pas} \]
\[ t \quad : \text{temperature in } ^\circ \text{C} \]

\[ a_0 = 1.398387 \cdot 10^{-3} \quad a_3 = -1.40322764 \cdot 10^{-9} \]
\[ a_1 = -2.78953077 \cdot 10^{-5} \quad a_4 = 2.8098933 \cdot 10^{-12} \]
$a_2 = 2,7923869 \cdot 10^{-7}$  
validity range $50 < t < 150 \, ^\circ C$  
Steam  
Density  
\[ \rho = a_0 + a_1 t + a_2 t^2 + a_3 t^3 + a_4 t^4 + a_5 t^5 \]  
$\rho$ : density in kg/m$^3$  
t : temperature in °C  
$a_0 = 5,31816778 \cdot 10^{-3}$  
a_1 = 2,53577316 \cdot 10^{-4}$  
a_2 = 1,41854076 \cdot 10^{-5}$  
a_3 = 8,18744927 \cdot 10^{-8}$  
a_4 = 2,74761584 \cdot 10^{-7}$  
a_5 = 6,8231375 \cdot 10^{-12}$  
validity range $40 < t < 170 \, ^\circ C$  
Specific heat capacity  
\[ c = a_0 + a_1 t + a_2 t^2 + a_3 t^3 + a_4 t^4 \]  
c : Specific heat capacity in J/kg .K  
t : temperature in °C  
a_0 = 1866,61248  
a_1 = 0,2292992  
a_2 = 1,0065478 \cdot 10^{-2}$  
a_3 = 1,731646 \cdot 10^{-5}$  
a_4 = 2,641961 \cdot 10^{-7}$  
validity range $20 < t < 160 \, ^\circ C$  
Specific enthalpy  
\[ h = a_0 + a_1 t + a_2 t^2 + a_3 t^3 + a_4 t^4 + a_5 t^5 \]  
h : Specific enthalpy in J/kg  
t : temperature in °C  
a_0 = 2,50107833 \cdot 10^6$  
a_1 = 1,88112 \cdot 10^3$  
a_2 = -1,3910397  
a_3 = 8,904895 \cdot 10^{-3}$  
a_4 = -1,01047 \cdot 10^{-4}$  
a_5 = 1,91 \cdot 10^{-7}$
validity range $20 < t < 160 \, ^\circ C$


Saturation pressure

$$p = a_0 + a_1 \cdot t + a_2 \cdot t^2 + a_3 \cdot t^3 + a_4 \cdot t^4 + a_5 \cdot t^5 + a_6 \cdot t^6$$

$\text{p}$ : pressure in bar

$t$ : temperature in $^\circ C$

$a_0 = 6,071346336 \cdot 10^{-3}$

$a_1 = 4,70344177 \cdot 10^{-4}$

$a_2 = 1,14074016 \cdot 10^{-5}$

$a_3 = 3,8375975 \cdot 10^{-7}$

$a_4 = 6,173605 \cdot 10^{-10}$

$a_5 = 4,592779 \cdot 10^{-11}$

$a_6 = -5,86980873 \cdot 10^{-14}$

validity range $30 < t < 160 \, ^\circ C$


Thermal conductivity

$$c = a_0 + a_1 \cdot t + a_2 \cdot t^2 + a_3 \cdot t^3$$

$c$ : Thermal conductivity in W/m K

$t$ : temperature in $^\circ C$

$a_0 = 1,83205026 \cdot 10^{-3}$

$a_1 = 5,5365096 \cdot 10^{-5}$

$a_2 = 1,06831 \cdot 10^{-8}$

$a_3 = 8,1532183 \cdot 10^{-10}$

validity range $20 < t < 170 \, ^\circ C$


Dynamic viscosity

$$\eta = a_0 + a_1 \cdot t + a_2 \cdot t^2 + a_3 \cdot t^3 + a_4 \cdot t^4 + a_5 \cdot t^5$$

$\eta$ : Dynamic viscosity in Pas

$t$ : temperature in $^\circ C$

$a_0 = 6,07570478 \cdot 10^{-6}$

$a_1 = 1,59108519 \cdot 10^{-7}$

$a_2 = -2, 8133159 \cdot 10^{-9}$

$a_3 = 3,2004751 \cdot 10^{-11}$

$a_4 = -1,75165413 \cdot 10^{-13}$

$a_5 = 3,6816396 \cdot 10^{-16}$

validity range $50 < t < 150 \, ^\circ C$
