PCT WORLD INTELLECTUAL PROPERTY ORGANIZATION
International Bureau

INTERNATIONAL APPLICATION PUBLISHED UNDER THE PATENT COOPERATION TREATY (PCT)

(51) International Patent Classification ^{6} **:**
C13D 1/00, C13F 1/02, 1/00

A1

(11) International Publication Number: **WO 98/37244**

(43) International Publication Date: 27 August 1998 (27.08.98)

(21) International Application Number: PCT/GR98/00005

(22) International Filing Date: 17 February 1998 (17.02.98)

(30) Priority Data:
970100071 24 February 1997 (24.02.97) GR

(71)(72) Applicant and Inventor: TSITSIS, Ioannis [GR/GR]; 12 Notara Street, GR-106 83 Athens (GR).

(81) Designated States: IL, US, European patent (AT, BE, CH, DE, DK, ES, FI, FR, GB, GR, IE, IT, LU, MC, NL, PT, SE).

Published
With international search report.
Before the expiration of the time limit for amending the claims and to be republished in the event of the receipt of amendments.

(54) Title: A PROCESS FOR WHITE SUGAR PRODUCTION

(57) Abstract

The process consists in the simultaneous treatment or co-treatment of S.B.s. and R.S. and is differentiated from previous co-treatment processes already existing. According to it, the thin juice is evaporated in a 2-effect evaporation station to an intermediate concentration level, set conveniently at about 37° BX assuming thereafter its final concentration level, set conveniently in the range of 64–68° BX (mostly 65–67° BX) by dissolution of proper quality R.S., eventually affinated. The process can be applied to already existing plants, operating on the previous technique if properly modified/partly revamped, further to being applicable on grass roots plants and is characterized by significantly higher production and competitiveness characteristics.

FOR THE PURPOSES OF INFORMATION ONLY

Codes used to identify States party to the PCT on the front pages of pamphlets publishing international applications under the PCT.

AL	Albania
AM	Armenia
AT	Austria
AU	Australia
AZ	Azerbaijan
BA	Bosnia and Herzegovina
BB	Barbados
BE	Belgium
BF	Burkina Faso
BG	Bulgaria
BJ	Benin
BR	Brazil
BY	Belarus
CA	Canada
CF	Central African Republic
CG	Congo
CH	Switzerland
CI	Côte d'Ivoire
CM	Cameroon
CN	China
CU	Cuba
CZ	Czech Republic
DE	Germany
DK	Denmark
EE	Estonia
ES	Spain
FI	Finland
FR	France
GA	Gabon
GB	United Kingdom
GE	Georgia
GH	Ghana
GN	Guinea
GR	Greece
HU	Hungary
IE	Ireland
IL	Israel
IS	Iceland
IT	Italy
JP	Japan
KE	Kenya
KG	Kyrgyzstan
KP	Democratic People's Republic of Korea
KR	Republic of Korea
KZ	Kazakstan
LC	Saint Lucia
LI	Liechtenstein
LK	Sri Lanka
LR	Liberia
LS	Lesotho
LT	Lithuania
LU	Luxembourg
LV	Latvia
MC	Monaco
MD	Republic of Moldova
MG	Madagascar
MK	The former Yugoslav Republic of Macedonia
ML	Mali
MN	Mongolia
MR	Mauritania
MW	Malawi
MX	Mexico
NE	Niger
NL	Netherlands
NO	Norway
NZ	New Zealand
PL	Poland
PT	Portugal
RO	Romania
RU	Russian Federation
SD	Sudan
SE	Sweden
SG	Singapore
SI	Slovenia
SK	Slovakia
SN	Senegal
SZ	Swaziland
TD	Chad
TG	Togo
TJ	Tajikistan
TM	Turkmenistan
TR	Turkey
TT	Trinidad and Tobago
UA	Ukraine
UG	Uganda
US	United States of America
UZ	Uzbekistan
VN	Viet Nam
YU	Yugoslavia
ZW	Zimbabwe

A PROCESS FOR WHITE SUGAR PRODUCTION

1. DESCRIPTION

1.1 FIELD OF APPLICATION

The invention relates to a process for white sugar production marked by high production, productivity and competitiveness characteristics.

A limited number of abbreviations has been used, made necessary due to the frequent recalling of them in the text. These abbreviations are:

S.B.	for sugar beets
S.C.	for sugar cane
R.S.	for raw sugar
W.S.	for white sugar

Exception to this using of abbreviations was made in the "claims", where the full pronouncement of terms has been deemed necessary, due to the relative independence of this section from the rest of the text.

The process is applicable on existing S.B. treating plants by means of adequate modifications / partial revamping inserted to them, or on grass root plants.

In case of necessity, it can be replaced by the previous technique in production activities, as production lines of both the process and the previous technique will be co-existing in the same plant; this case of necessity, however is an extraordinarily unlikely case, as practically R.S. (to be also used in processing) is, in the international market, always in the reach of W.S. producers and at prices which very rarely (if at all) may become so prohibitive to reverse the large process economic superiority over the previous technique.

1.2 STATUS OF EXISTING TECHNOLOGY - INHERENT DISADVANTAGES OF IT

The traditional technology or previous technique, as it is known, is elaborated in its basic form, by utilizing S.B. (or S.C. in tropical countries) as the unique raw material and has a number of drawbacks frequently having to do with (low) production and productivity and also, from time to time, with low quality of W.S. produced. So W.S. manufactured must be based on qualitatively acceptable S.B. to produce satisfactory productivity and economic results. By lower quality beets, as eventually may be the case, W.S. production and productivity may drop to unacceptable law levels, extending even to negative economic results in extreme cases, while quality of W.S. is more than usually degraded.

On the other hand slightly reckless processing, may lead to sugar losses or juice qualitative deterioration (colour building) in so sensitive a technology producing similar type drawbacks as before (lower and qualitatively degraded production). And all this in an effective production time which can be as short as 100-120, or even less, operating days per year.

To closer investigate the main problems afflicting the previous technology we would mention that:

1. In the evaporation station, aiming at concentrating the juice to 60°BX, the juice is forced to stay, for a relatively long time, under high operating temperatures, the highest in the whole production chain (indicatively from 135 to 95 °C) thus creating the conditions for invert sugar formation, and invert sugar sub-product breakdown, leading to colour building.
2. With time going on, in a sugar campaign, colour gets worse as S.B. quality deteriorates in the mean time. Not only but thick juice starts getting less and less concentrated, when leaving the evaporation station, due to its decreased susceptibility to evaporate (increased difficulty of juice to evaporate) and the progressively increasing deposition of scale n the heat exchange surfaces. This has serious impacts on the whole production image. Less concentrated thick juice from the evaporation station in late campaign means less vapor produced in that and, therefore, less vapor sent to the rest of the plant and, especially, to crystallization station, which is a major consumer. In this latter finally three negative process events are cumulatively superimposed in that period, the lower concentration of thick juice, the inferior quality of it, and the lower quantities of vapor sent to it (from evaporation station). Therefore longer (significantly longer) crystallization cycles in the vacuum pans will be observed in the end period, leading to further colour building and production restraining.

3. Evaporation and crystallization problems as above, at the end of campaign (and even before that), have a strong ally also in filtration processes in the plant. With the general degradation of S.B. quality and some possible mistakes on part of the operating personnel in processing itself, in so sensitive a technology, filtration creates its own problems, getting slower and slower with the time. All this contributes in shortening the life of a campaign to quite brief periods, say of 100-120 operating days (if not less). If prolonged beyond that time length, campaign would have started producing negative economic results.
4. On top of all this comes the exceptionally bad campaign. When, due to unusually adverse climatic conditions or other circumstances, S.B. quality deteriorates to a degree that exceeds the acceptability limit of S.B.s, campaign results may get so bad as to determine even negative economic results over a part or even over the whole of it.
5. Salvation to the above problems has been sought among other in processing R.S. either at time periods different of the main campaign, or simultaneously with the S.B. processing.

   Outside or separate R.S. processing, which is mainly aimed at reffinade production, is out of our concern. On the contrary co-processing of R.S. with S.B. is the one we care about. In the previous co-processing techniques, R.S. is added to the juice purification phase, which means that production will be faced with the following problem: R.S. cannot be added (according to professor F. Schneider, ex. director of the technical institute of Braunschweig) at rates superior to 8-10% on S.B. as the whole R.S./S.B. juice blend has to be purified in the juice purification station and thereafter concentrated to 60°BX in the evaporation station by means of steam consumption.

   Purification through the juice purification station means overloading of its equipment and the following thickeners and filters with quite thicker juices.

   Concentration to 60 °BX in the evaporation station means exposing the juice to colour building and to bottlenecking phenomena (especially in the second half of the campaign) exactly in the same way as in the S.B. - only - based process due to the increased difficulty of juice to evaporate during the end period.

Therefore processing the S.Bs along with R.S. according to the above procedure is a half solution and a half salvation only.

1.3 ADVANTAGES OF THE INVENTION

The above disadvantages of the previous technique are reversed by the invention to a degree of becoming clear advantages. Leaving for the following chapter "Brief presentation of the invention", the technical description of it, we will concentrate here in simply enumerating these advantages:

1. Production is raised a few times over the rate associated with the previous technique. Not only but raw material input is stabilized and finished product output is nearly stabilized and at the same time qualitatively improved.

   According to the numerical examples 1 and 2 (treatment schemes A and B, pages 40-48), with an approximate modification cost to an existing plant, operating on the previous technique, of 24-28% (on the old plant equipment value), following rates will be obtained.

   On a daily basis entering sugar rate to the plant for processing is raised to 2,7-2,9 times the previous respective rate, while recovered sugar rate is raised to 3,0-3,1 times the previous rate.

   On a yearly basis (due to the possibility of the campaign to be extended under profit terms) entering sugar rate may be raised to 3,2-3,4 times the previous respective figure, while recovered sugar rate can be 3,5-3,7 times the respective previous one (see also tables 8 and 8' and further on tables 1Π, 1A and 2B pages 40-48 at the end of the text) provided that the extension rate of campaign is a 20% (this 20% being a mere assumption, which however, is quite logical, if not at all conservative, see table 4, note 1).
2. Bottlenecking phenomena, appearing at the end of campaign, mainly originated in the evaporation and crystallization station, but also in the various filtration facilities, are disappeared (or practically disappeared). Processed throughput rates, are at the end of campaign the same as in the beginning.
3. Daily processed rates of S.B. are also higher due to elimination of bottlenecking. According to the numerical examples (pages 40-48) which are based on campaign historical data, S.B. average daily throughput is in the range of 1,18 times the previous rate (18% increased) the 18% increase rate

reflecting the daily average of the 10 best days of the campaign as compared to the total campaign daily average. This 10 best days daily average is, by engineering judgement, very close to the theoretical capacity of the plant, if not literally coinciding with it.

On a yearly basis (due to the conservatively estimated extension of campaign by a 20%) processed rate of S.B.S. will be 1,1 8 x 1,2 = 1,42 times the previous rate (by full exploitation of production possibilities of the process).

4. W.S. produced is of better quality (lighter colour e.t.c.). This quality superiority is of course more evident if comparison with the previous technique is made in the second half and especially at the end of the campaign whereby sugar quality by the previous technique is usually clearly inferior.
5. W.S. unit production cost, is, as a rule, lower whereas cost effectiveness, referring to total campaign results, expressed as campaign gross profits, is in the totality of cases, by far higher, even several times higher (see further below pages 7-10).

1.4 BRIEF PRESENTATION OF THE INVENTION

In the proposed process thin juice is not any more concentrated to 60 °BX (in the evaporation station), but only to an intermediate, moderate concentration level, which, according to the numerical examples given later on in the text, is 37,1 °BX. This is accomplished in a reduced evaporation station, consisting now of two effects, eventually with slightly modified capacity. This semi-concentrated juice, which we will call from here on the semi-thick juice, has the definite advantage of being lighter coloured compared to the thick juice of the previous technique, **the comparison being considered at equal concentration level (colour test results brought to the same degrees BX.)**. The reason for that is to be sought in the significantly shorter time the semi-thick juice will stay now in the 2-effect evaporation station. Also in that temperature profile is lower with start temperatures in the range of 125-120°C and even below.

To the semi-thick juice, originated as above, proper quality R.S. (not end-product R.S., as a rule), is added at approximately 20% wt on beets, which is translated to a ratio: solid-matter-in-R.S. versus solid-matter-in-S.B. of 123-131/100 or to a ratio sugars-in-R.S. versus sugar-in-S.B. of 132-141/100 (according to the numerical examples). The resultant juice originated, called the super-thick or the final juice is now conveying (130/140+100)x1,18=271/283 weight units of sugar compared to 100 units of the previous technique as the S.B. input rate is, by the process, also higher, i.e. 1,18 times the previous rate (see advantages No1 and 3 of previous section); not only but it is, thanks to the upgrading/stabilizing effect of the R.S., also qualitatively and technically upgraded having now, as a result of this effect, a steady concentration of 66,5°BX (numerical example 1) and 63°BX (numerical example 2), obtained half-way by dissolution techniques, (by the addition of proper quality R.S.) and a nearly steady purity of 95% or 96,5% (according to same numerical examples) all the way through a sugar campaign (see advantage No 4 of previous section). In the previous technique the corresponding figures of thick juice are respectively 60-62 °BX, and 90-92% with the tendency of them to further decline in the second half and especially at the end of the campaign (i.e. 55° BX concentration and 87-88% purity).

The R.S. added is either of S.C. or S.B. origin. In the first case it is washed off to remove the excess invert sugar present (about 0,8%). In the second case washing is not, as a rule, indispensable, as invert sugar presence is always at lower levels. Super-thick or final juice produced, following the addition of R.S., is of lighter colour compared to the thick-juice of the previous technique, provided that washing of cane R.S. is carried out properly and that R.S. utilized is not of clearly inferior quality, i.e. not an end-product R.S. (see advantage No 4 of previous section).

So far as the evaporation station and the subsequent dissolution of R.S., to the semi-thick juice, is concerned. In the crystallization station, the other eminent fireside of invert sugar formation and colour building by the previous technique in late campaign things are again by far better in the new process. Colour building is contained to eminently lower levels due to the fact that boiling cycles are not but slightly only elongated at the end period, owing to the upgrading/stabilizing effect of the R.S. This determines a clear qualitative advantage of the process. Further to this, however, comes also the quantitative advantage, as production is no longer sticking to decisively, longer boiling cycles at the end of campaign, allowing thereby S.B. throughput to the plant to stay at the initial levels, something that will be further eased by a modest reinforcement of the crystallization station. This enables production to stay at a campaign average daily throughput of S.B. which is at least equal to that of the campaign 10 best days, therefore (as per the numerical examples)

1,18 times higher, as there will be no bottleneck any more to hamper it (see advantages No 2 and 3 of previous section).

At the same time total W.S. output is a number of times higher, due to the sugar content of R.S., also introduced to the plant, in addition to the sugar load from S.B.s. All of this sugar will be comfortably processed, following the modification of the evaporation station and the moderate revamping of the second part of the plant (crystallization station e.t.c.) assisted by the decisively higher process velocities now-achieved and the drastic curbing of idle times in the batch operating units of the plant (see tables that follow and especially tables 2 and 4, two bottom lines and 9.1, 9.2, 9.3, 9.4. tables, also table 5).

It is obvious that the resultant effect of higher sugar rates processed daily, and of the campaign extension towards longer periods of time, combined with a relatively low modification cost to make a plant operable on the new process, will determine a better cost effectiveness (unless in cases of extraordinarily high R.S. prices, which however can also be avoided by properly organizing R.S. acquisition activities, utilizing the modern information and computing techniques).

1. 5 ECONOMIC EFFECTIVENESS OF THE INVENTION IN COMPARISON TO THE PREVIOUS TECHNIQUE

As R.S. is a first order cost factor, when manufacturing W.S. by means of the process, it is quite important to quantitatively define its impact on W.S. production cost and on global economic results over an entire campaign since this is a process proposed for industrial application. In doing so, as a first step, a R.S./W.S. price ratio is selected, which according to historical background, is considered as being a statistically representative price ratio between the two sugar types. The percentage ratio of 75%, that is the price ratio R.S./W.S. of 75/100 is considered as being such a ratio, although on non-favorable assumption. A further approach to the subject will be determining the R.S. break-even price, i.e. that price level, above which economic attractiveness of the two processes may be reversed in favor of the previous technique. The R.S. break-even price matter is covered later on (see table 6 at the end). The statistically representative price ratio case, as defined above, is covered right here below (see also tables 1 and 3, production cost).

Cost effectiveness of the invention as compared to the previous technique by 75/100 R.S./W.S. price ratio-Reference period: The early '70s

Treatment scheme A as against the old technique.

Sub-case 1: Totally non-depreciated plant (both old equipment and modification equipment are still under depreciation).

Process

- W.S. ex-factory price: 9,44 drch/kg
- W.S. production cost[1] : 7,725 drch/kg (see table 3 at the end)
- R.S. acquisition price: 7,08 drch/kg CIF plant (75% of W.S. price)
- Campaign production rate in W.S.: 96,67 thousand tones

Therefore campaign profit before taxes:

(9,44-7,725)x96,67= **165,8** million drchs.

Previous technique

- W.S. ex-factory price: 9,44 drch/kg
- W.S. production cost: 8,223 drch/kg (see table 1, non-depreciated plant)
- Campaign production rate in W.S.: 25,74 thousand tones

Therefore campaign profit before taxes:

(9,44-8,223)x25,74 = 31,3 million drchs.

$$\frac{process}{previoustechnique}:\frac{165{,}8}{31{,}3} = \mathbf{5{,}3} \text{ or } \mathbf{530\%}$$

Cost effectiveness by lower R.S. price by 20%
(R.S. price 60% of W.S. price)

Process

Respective production cost: **6,785** drch/kg

Therefore campaign profit before taxes:

(9,44-6,785)x96,67 = **256,6 mill.drchs.**

---

(1) production cost in all cases as per below, includes also overheads.

Previous technique

Campaign profit before taxes **31,1 mill.drchs** (same as before)

$$\frac{process}{previoustechnique}:\frac{256,6}{31,3} = \mathbf{8,2} \text{ or } \mathbf{820\%}$$

Sub-case 2: **Partly, depreciated plant (old equipment depreciated, modification equipment still under depreciation).**

Process

- W.S. ex-factory price: 9,44 drch/kg
- W.S. production cost: 7,298 drch/kg (see table 3 partly depreciated plant)
- R.S. acquisition price: 7,08 drch/kg, CIF plant
- Campaign production rate in W.S.: 96,67 thousand tones

Therefore campaign profit before taxes

(9,44-7,298)x96,67 = **207,1** million drchs.

Previous technique

- W.S. ex-factory price: 9,44 drch/kg
- W.S. production cost: 6,585(1) drch/kg (see table 1 depreciated plant).
- Campaign production rate in W.S.: 25,74 thousand tones

Therefore campaign profit before taxes

(9,44-6,585)x25,74 = 73,5 million drchs.

$$\frac{process}{previoustechnique}:\frac{207,1}{73,5} = \mathbf{2,82} \text{ or } \mathbf{282\%}$$

---

(1) Low cost in this case, as old apparatus of the plant, referring to the previous technique, is depreciated. However, in spite of that, total profit before taxes, referring to the entire campaign, is higher in the process case due to the significantly higher production volume related.

To note again that above comparative results are valid at the assumption of a R.S./W.S. price ratio of 75/100 which quite often may be lower.

Cost effectiveness by lower R.S. price by 20%
(R.S. price 60% on W.S. price)

Process

Respective production costs **6,352** drchs/kg
Therefore campaign profit before taxes
(8,44-6,352)x96,.67 = **297,9 mill.drch.**

Previous technique

Campaign profit before taxes **73,5** drchs/kg (same as before)

$$\frac{process}{previoustechnique}:\frac{297{,}91}{73{,}5}=4{,}05 \text{ or } \mathbf{405\%}$$

1.6 DETAILED DESCRIPTION OF THE INVENTION - REFERENCE TO RESPECTIVE TABLES AND DRAWINGS

Having covered the topic of comparative economic effectiveness, which we deemed necessary in order to show the industrial or business importance of the invention, we will now concentrate on the detailed description of the process.

The new process develops in two parallel, alternative production schemes, scheme A or arithmetic example 1 and scheme B or arithmetic example 2 - as already mentioned earlier in the text.

In scheme A, the R.S.* as also earlier occasionally mentioned is added directly to the semi-thick juice without being subjected to previous washing (not affinated). This is the case, where (as mentioned) the R.S. is of S.B. origin, whereby the presence of invert sugar is quite limited to create colour building and colloid-associated/filtration problems. In case of necessity S.B. origin R.S. can, of course, undergo a preliminary washing (affination) treatment as well before being added to the semi-thick juice.

In scheme B, or arithmetic example 2, R.S.* is added after being affinated. This is in order to remove the excess invert sugar (contained in the order of magnitude of 0,8% WT) and also to remove the excess colloids thus freeing the process from the harmful effects of their presence.

In both schemes A and B (which represents ways of materialization of the new process) the new process is formally the same with the previous technique, up to the evaporation station, save the quantitative parameter of processed material, i.e. the higher daily rates of sugar beets processed in the process case. These daily rates are taken 1,18 times higher than those of the previous technique (to reflect facts stated, in page 4, par. 3 and elsewhere in the text).

The evaporation station is the big diversification point dictating a modification of its configuration, namely reduction of its size. According to this, the first two stages of a 4-stage station will remain untouched, with a possible adaptation of their capacities if it is necessary to produce the 37,1° BX

---

* Proper quality R.S. i.e. not end-product R.S. as a rule.

The rest of the effects 3rd and 4th are not any more necessary and are eliminated from operations, but they will still physically stay in their original position to be used in a 4-effect evaporation station, if usage of the previous technique, although extremely unlikely, might be required -see page 1, «Field of application», last 6 lines.

The total process equipment modifications required to make a plant operable on the new process are roughly estimated below:

S.B. storage and movement	intact
S.B. washing	‘’
S.B. cutting	‘’
extraction (diffusion)	‘’
juice purification and associated facilities(1)	‘’
evaporation	new capacity 52% about of previous (reduced size)
sugar affination/dissolution	new capacity about 650% ‘’
crystallization	10 vacuum pans instead of 7
crystallization	15 mixers instead of 9
crystallization	25 centrifuges instead of 14
sugar drying-conveying-bagging	new capacity about 270% of previous

The above is the strictly process considered equipment modification with a roughly estimated modification cost on total equipment, i.e. process, auxiliary, utilities e.t.c. equipment, of the old plant of 23-24%. If also some additional capacity for R.S. and W.S. storage and for the boiler station is considered, the modification cost will commensurately be increased but will always stay well low -27-28% on total old plant equipment, as an order of magnitude.

New steam/vapor production and consumption rates, emerging from the process, in comparison with those of the previous technique are shown in table 7 at the end. Steam requirements for the rest of the plant (units outside the evaporation station will be comfortably met as a result of the new arrangement, i.e. the 2-effect evaporation station.

With that we will come now to the analytical description of the two alternative treatment schemes of the process, mentioned above.

---

(1) such as limestone kiln, thickeners, filters, pumps, lime preparation system etc.

**Treatment scheme A or arithmetic example 1 (direct dissolution of R.S.)
Reference drawing 1, Reference table 1A (to be read in conjunction with table 1Π referring to the previous technique for comparison)**

It is noted that plain numbers below are referring to the drawing, while numbers accompanied by letter A are referring to the table 1A. It is also noted that description of process streams and, their function in operations, starts from the point where the new process is differentiated from the previous one, namely from evaporation station on. So:

a. Production of semi-thick juice of 37,1 °BX - process streams 1,2,3,4,5,6,7 and 1A,2A,3A,4A,5A,6A,7A respectively, (see explanation above for plain and letter-accompanied numbers).

b. Production of the super-thick juice or the final juice of 66,5 °BX after dissolution of the imported, proper quality, R.S. and of the recycled sugars B,C, and D of the plant to the 37,1 °BX juice in the dissolution tank, clarification and eventual decoloration with diatomaceous earth and decoloration coal - process streams 3,10,11,12 and 3A,10A,11A,12A respectively.

c. Crystallization of above in product A vacuum pans of the crystallization station; masquite A production - process streams 13,14,15,16 and 13A,14A,15A,16A respectively.

d. Stirring and ripening of masquite A in the masquite A mixers, for sugar crystal enlarging and increasing of yields - process stream 15 and 15A respectively.

e. Centrifugation of stirred masquite A in the product A centrifugals with following washing (rinsing) in situ, thereby production of sugar A, conveyed after drying to the bagging machines - process stream 17 and 17A respectively.

f. At the same time separation, from product A centrifugals, of syrup A, which feeds in the next product B circuit of the crystallization station - process stream 18 and 18A respectively.

g. Production, from circuit B, by similar procedure, of sugar B, not rinsed in situ, and of syrup B feeding in the next product C circuit -process streams 18,19,20,21,22 and 18A,19A,20A,21A,22A respectively.

h. Production from circuit C by similar procedure, of sugar C, not rinsed in situ, and at the same time of syrup C, feeding in the next product D circuit - process streams 22,23,24,25,26 and 22A,23A,24A,25A, 26A respectively.

i. Production, as per above, of sugar D from circuit D, also not rinsed in situ, and recycling of sugars B,C and D to the dissolution tank in order for them to be dissolved together with the imported R.S. At the same time separation from D product centrifugals of molasses - process streams 27,28,29,30,11 and 27A,28A,29A,30A,11A respectively.

Treatment scheme B or arithmetic example 2. (dissolution of R.S. after affination)

Reference drawing 2, reference table 2B (to be read in conjunction with table 1Π for comparison)

Same notes apply concerning reference numbers of process streams and start point of description of the process as with treatment scheme A (see notes above). To be also noted that washing of R.S. will be carried out by means of a mixture of syrups from the crystallization station (syrup A and B) added by smaller quantities of dilution water (to the dissolution tank) and rinsing water (to the centrifugals) as described in relative bibliography. The syrup separated at the centrifugal from affinated R.S. will be directed, along with product A syrup, to product B circuit (product B vacuum pans) for concentration and crystallization. The temperature of affination will be as required (usually in the range of 55°-60 °C). So, similarly to treatment scheme A:

a. Affination (washing off) of R.S. and recycled sugar B,C and D of the plant and following centrifugation of washed sugar - process streams 12,48,37,15,16,17,25 and 12B,48B,37B,15B,16B,17B,25B respectively, (see explanation at the beginning of «treatment scheme A» for plain numbers and letter - accompanied numbers).

b. Production of the 37,1°BX semi-thick juice - process streams 1,2,3,4,5,6,7 and 1B,2B,3B,4B,5B,6B,7B respectively.

c. Production of the 62,9 °BX final juice (steady concentration to the last day of campaign) by dissolution of the affinated R.S. and recycled sugars B.C. and D to the semi-thick juice of 37,1 °BX. Clarification/decoloration of it with diatomaceous earth and decoloration coal - process streams 3,17,18 and 3B,17B,18B respectively.

d. Evaporation/crystallization of super-thick juice in product A vacuum pans - process streams 22,22' and 22B, 22'B respectively.

e. Centrifugation of product A masquite and following rinsing of separated sugar A on the centrifugals. Forwarding of sugar A (W.S.) thus produced to the bagging machines after drying-process, stream 23 and 23.B respectivelly.

f. Simultaneously to sugar A, production also of product A syrup, feeding in the next product B circuit - process streams 24,26,27 and 24B,26B,27B respectively.

g. Production, in product B circuit, of sugar B, not washed in situ, and of syrup B, feeding in the next product C circuit - process streams 28,31,32,33,34,35,36 and 28B,31B,32B,33B,34B,35B,36B respectively.

h Production, in the same way, of sugar C, from product C circuit, not washed on the centrifugals, and at the same time of syrup C, feeding in the next product D circuit - process streams 35,35',40,41,46 and 35B,35'B,40B,41B,46B respectively.

i. Production, by same procedure, of sugar D, from product D circuit, not washed on the centrifugals, and recycling of all non-washed plant sugars B,C and D to a common mixer, where they will be washed off, along with the imported R.S. At the same time productions of molasses - process streams 41,41',44,45,48,47 and 41B,41'B,44B,45B,48B,47B respectively.

TABLES - DIAGRAMS - DRAWINGS
A QUICK REVIEW OF THEM

What so far expressed in the previous sections of the text are summarized in terms of numbers, sketches and brief comments in the following tables, diagrams and drawings:

Tables 1 and 2, showing W.S. production cost analysis and equipment capacity utilization with regard to the previous technique.

Tables 3 and 4 , showing the same as above with regard to the process.

Diagrams 1 and 2, illustrating production schemes in relation to the previous technique and the process respectively.

Table 5, showing how an approximate 3-fold production is possible by means of the process provisions, though at a low modification / revamping level.

Chart 6, showing calculations for determination of R.S. break-even-price,. namely the price by which the process may loose its large economic superiority over the previous technique.

Table-sketch 7, illustrating the new steam balance as compared to the previous one, especially around the evaporation and crystallization stations which are key producers/consumers of vapor/steam.

Tables 8 and 8', showing key production indicators with regard to treatment scheme A and B, as compared to the previous technique.

Tables 9, namely **9.1,** **9.2,** **9.3** and **9.4,** showing detailed calculations with regard to the crystallization station operation and production, proving in effect the decisively higher production and productivity possibilities of the new process.

Table 1Π, showing process stream rates and properties referring to the previous technique.

Tables 1A and **2B**, showing process stream rates and properties referring to treatment schemes A and B respectively.

Drawings 1 and 2, illustrating the modified part of the plant in correspondence to treatment scheme A and treatment scheme B of the process.

Table 1

Previous Technique
Raw Material: S.B.s
Cost Analysis⁽¹⁾
S.B.s treated in campaign: 218,131 tones
Operations period: 110 days
W.S. produced: 25.740 tones
Processed S.B.s versus W.S. produced: 8,474 kg/1 kg
S.B. price, CIF manufacturing plant: 0,78 drchs/kg

		Non-depreciated plant			**Depreciated plant**		
Cost participation of	S.B.s.	6,61	drchs/kg	W.S.	6,61	drchs/kg	W.S.
"	personnel	0,839	"	"	0,839	"	"
"	administration	0,135	"	"	0,135	"	"
"	fuel/power	0,389	"	"	0,389	"	"
"	packing	0,070	"	"	0,070	"	"
"	auxiliary material/maint	0,712	"	"	0,712	"	"
"	depreciation	1,638	"	"	-		
"	relief from by products(2)	(2,170)	"	"	(2,170)	"	"
Total production cost		8,223	drchs/kg	W.S.	6,585	drchs/kg	W.S.

(1) From historical data referring to the early '70s.

(2) Molasses and pulp.

Table 2
Previous Technique
Raw Material: S.B.s
Utilization Factor of equipment
110 operating days

campaign factor $\frac{110}{365}$ =0,30

	Capacity utilization factor in framework of a sugar campaign			Capacity utilization factor in framework of calendar year		
	Begin, main part of campaign	**End of campaign**	**Average campaign**	**Begin, main part of campaign(2')**	**End of campaign(2')**	**Average campaign(2')**
Beet stor. Beet wash. Beet cutting	0,852(2)	0,767(2)	0,84(1)	0,255	0,230	0.252
Juice extract. juice purific.	0,852	0.767	0,84(3)	0,255	0,230	0.252
Evaporation	0,852	0,703(3)	0,832(3)	0,255	0,211	0.249
Dissolution of sugar	0.852	0.767	0,84	0,255	0,230	0,252
Drying, conveying bagging of sugar.	0,852	0,767	0,84	0,255	0,230	0,252
Crystallization	0,475(4)	0,428(5)	0,468(6)	0,142	0,128	0.140
	Time utilization factor of crystallization station					
	0,475(7)	**0,7496(7)**	**0,512(8)**	**0,142**	**0,225**	**0,154**

Notes:

(1) From real campaign measurements has resulted that daily average beet throughput rate was 85% of that of the 10 best campaign days, or inversely 100/85=118%, the average daily beet throughput rate of the 10 best days over the campaign average. The reason for having a drop of the campaign average in the previous technique is to be sought in the bottlenecking phenomena escalating in the second half and especially at the end of campaign period In the new process due to striking out of bottlenecks the campaign average will be raised, very close to maximum capacity all the way through to the last day of the campaign. We assume real throughput or maximum achieved capacity, to be 99% of the theoretical capacity (see also table 4).

Therefore the campaign average beet throughput rate in the previous technique will be according to above

0,99x0,85=0,84 versus theoretical capacity

(2) End period throughput measured for 15 days out of the 110 total operating days has resulted in approx. rates of 0,89% of the beginning and main part ones; taken at 0,9%. Therefore throughput factor at beginning/main part (95 days) and at end of campaign (15 days), called X and 0,9X respectively:

$$\frac{[95.X]+[15.(0{,}9.X)]}{110} = 0{,}84 \text{ , and } \begin{cases} X=0{,}852 \text{ (beginning and main part of campaing)} \\ 0{,}9X=0{,}767 \text{ (end of campaign)} \end{cases}$$

(2') The right handside figures multiplied by the campaign factor 0,30.

(3)— Evaporation station is a special case concerning capacity utilization factor. The end period lower throughput (0,767), as in the rest of the plant units, is further combined here with lower concentration of thick juice. Taken at 55°BX (instead of 60 °BX).

Therefore capacity utilization at end period (measured by vapor production as compared with theoretical vap. production equal to $1{,}00\text{x}\,\frac{60}{60}=1{,}00$).

$$0{,}767x\frac{55}{60} = 0{,}703$$

Consequently average capacity utilization over a campaign

$$\frac{(0{,}852.95)+(0{,}703.15)}{110} = 0{,}832$$

— Crystallization station (being a major batch operation station) is even more complicate as far as capacity utilization and time factor (time utilization) are concerned.

(4) Measured occupancy (productive time) of product A vacuum pans at beginning (and main part) of campaign: 47,5% (or 0,475). This coincides numerically with capacity utilization factor, (therefore c.u.f. 47.5%) as in the beginning period vacuum pans (and in general the whole crystallization station) have their maximal productivity therefore their theoretical capacity (boiling cycle, at product A vacuum pans, minimum, i.e. 3.5 hours for said measurements).

(5) In the end period, irrespective of longer occupancy of equipment (74,96%) capacity utilization is lower since lower are the throughput rates of raw material, (same capacity utilization decrease as raw material throughput decrease i.e. $0{,}475\text{x}\ \frac{0{,}767}{0{,}852}=0{,}428$)

(6) Average capacity utilization throughout the whole campaign: $\frac{(0{,}475.95)+(0{,}428.15)}{110}=0{,}468$

(7) In beginning the observed occupancy (time factor) 0,475 corresponding to **3,5 hours boiling cycle**. At the end as per calculations, based **on 5,5 hours boiling cycles** time factor (estimated accupancy) is 0,7496. (see calculation sheets of tables 8 further below, sheet 982).

(8) Campaign average as per the usual calculations: 0,512.

Table 3

New process
Raw Materials: S.B.s + R.S.
Cost analysis (based on real, previous technique campaign data, properly elaborated to reflect the requirement's of the process)
S.B.s treated in campaign: 308,880 tones
R.S. treated in campaign: 54,284 tones
Operations period: 132 days (110 days of prev. technique x1,2)
Total W.S. produced: 96.670 tones
S.B.S. treated versus W.S. produced: 3,19 kg/1 kg
R.S. treated versus W.S. produced: 0,665 kg/1 kg
S.B. price, CIF manufacturing plant: 0,78 drchs/kg
R.S. price, CIF manufacturing plant: 7,08 drchs/kg
(75% of W.S. price)

		Totally non-depreciated plant			Partly depreciated plant		
Cost participation of	S.B.s.	2,49	drchs/kg	W.S.	2,49	drchs/kg	W.S.
‘’	R.S.	4,71	‘’	‘’	4,71	‘’	‘’
‘’	total raw materials	7,2	‘’	‘’	7,2	‘’	‘’
‘’	personnel(2)	0,289	‘’	‘’	0,289	‘’	‘’
‘’	administration	0,046	‘’	‘’	0,046	‘’	‘’
‘’	fuel/power	0,158	‘’	‘’	0,158	‘’	‘’
‘’	packing	0,070	‘’	‘’	0,070	‘’	‘’
‘’	auxiliary materials maintenance	0,297			0,297		
‘’	deprecation	0,556	‘’	‘’	0,129(3)		
‘’ relief from byproducts(4)		(0,891)	‘’	‘’	(0,891)	‘’	‘’
Total production cost		7,725	drchs/kg	W.S.	7,298	drchs/kg	W.S.

(2) Estimated increase of personnel 30% compared to the old plant operating on the previous technique, therefore 172x1,3=224, i.e. 52 persons more, whereby 172 persons the manpower refercing to a previous technique based operation.

(3) Only modification equipment non-depreciated

(4) molasses, pulp.

Production cost by lower R.S. price

If R.S. cheaper by, say 20% (which is quite possible to occur) production cost will be:
6,785 drch/kg and 6,358 drch/kg
for non-depreciated/partly depreciated plant respectively

Table 4

New Process
Raw Materials S.B.s + R.S.
Utilization factor
of equipment

132(1) operating days - campaign factor: $\frac{132}{365} = 0{,}36$

	Utilization factor of equipment capacity in the framework of a sugar campaign			Utilization factor of equipment capacity in the framework of calendar year		
	Utilization factor, begin and main body of campaign	**Utilization factor, end of campaign**	**Utilization factor, campaign average**	**Utilization factor, begin and main body of campaign(2')**	**Utilization factor, end of campaign(2')**	**Utilization factor, campaign average(2')**
Beet stor. Beet wash. Beet cutt.	0,99(2)	0,99(2)	0-,99(2)	0,356	0,356	0.356
Juice extract. juice purific.	0,99	0,99	0,99	0,356	0,356	0,356
Evaporation	0,99	0,99(3)	0,99	0.356	0,356	0.356
Dissolution of sugars	0,99	0,99	0,99	0,356	0.356	0.356
Drying, conveying bagging of W.S.	0,99	0,99	0,99	0,356	0.356	0.356
Crystallization	0,729(4)	0,729(5)	0,729	0,262	0,262	0,262
	Time utilization factor of crystallization station					
	0,729(6)	**0,8840(7)**	**0,750(7')**	**0,262**	**0,318**	**0,279**

(1) Extended campaign, as compared to the previous-technique-based campaign, as even second hand beets can be processed now, under profit terms, because of the stabilizing (upgrading) effect of R.S. Conservatively assumed 20% campaign extension, therefore 110x1,2=132 productive days.

(2) See note (1) of table 2.

(2') The right handside figures multiplied by the campaign factor 0,36.

(3) Contrary to the previous technique the evaporation capacity of the evaporation station (vapors produced per hour) will remain firm throughout the whole campaign, as juice will be concentrated moderately (37,1°BX as per the arithmetic examples) which will be easy to obtain in all instances to the last camp. day (the rest of the concentration to 66,5°BX being obtained by R.S. dissolution). Therefore capacity utilization of the modified (reduced) evaporation station at the end will be the same as in the beginning, assumed 99%.

(4) Calculated occupancy (productive time) of product A vaccum pans at beginning (and main part) of campaing 72,9% or 0,729. This coincides numerically with capacity utilization factor (therefore c.u.f. 72,9%) as in the beginning period vacuum pans (and in general the whole crystallization station) have their maximal productivity, therefore their theoretical capacity

(boiling cycle at product A vacuum pans, 2,5 hours as per estimation),(see calculations on calculation sheet 9.3 of the series 9 further below).

(5) In the end period, regardless of the esstimated slightly higher time utilization factor of equipment (20% higher only related to 3,0 hours boiling cycle instead of 2,5 hours in the beginning, as a result of the upgrading/stabilizing effect of the R.S.) capacity utilization is the same as in the beginning of campaign, as raw material throughput (S.B.s. and R.S.) is the same with that of the beginning (bottlenecks elimination being, one of the main characteristics of the process). Therefore capacity utilization at the end: 72,9% (or 0,729).

(6) In beginning the estimated occupancy (time factor) is at 72,9% coinciding numerically with the utilization factor (see above), see also calculation sheet 9.3, in the series 9.

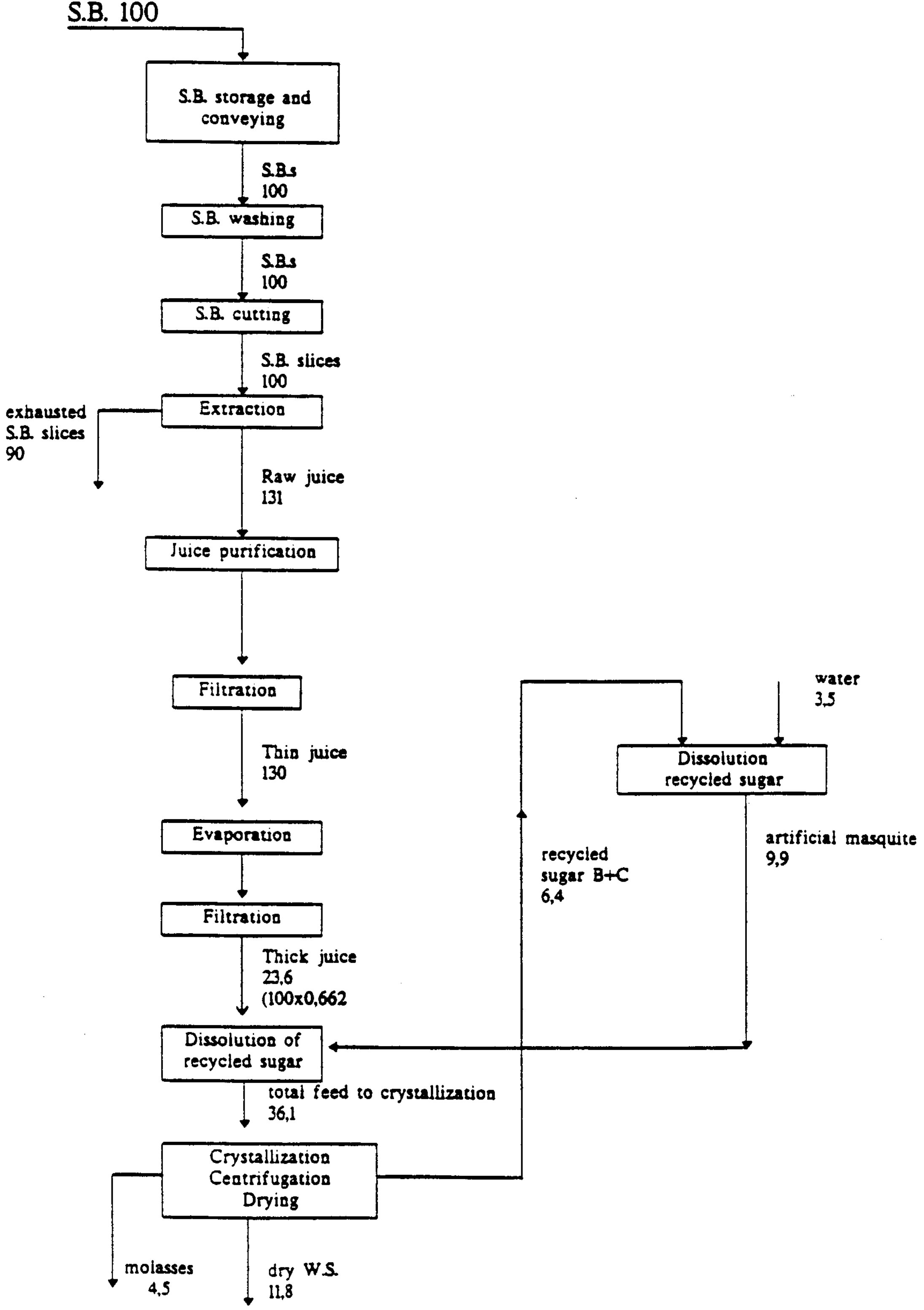
(7) At the end slightly increased occupancy (time factor) estimated at 88,4% or 0,884 (see calculation sheet 9.4 of the series 9 further below).

(7') Campaign average time factor: $\frac{(0{,}729.114)+(0{,}884.18)}{132}=0{,}750$

Schematic Production Diagram 1

Previous Technique
110 productive days

All process streams are weight rates referring to S.B. throughput taken equal to 100



Schematic Diagram 2

New Process-Treatment scheme B
Raw Material : S.B. + R.S.
132 productive days (110x1,2=132)

All process streams refer to S.B. throughout (weight rate) further referring to diagram 1 S.B. throughput, taken equal to 100. Therefore diagram 2 S.B. throughput:

$$100x\frac{99}{84}x1{,}2=141$$ **weight rate (see also tables 2 and 4 and also 2B, treatment scheme B)**

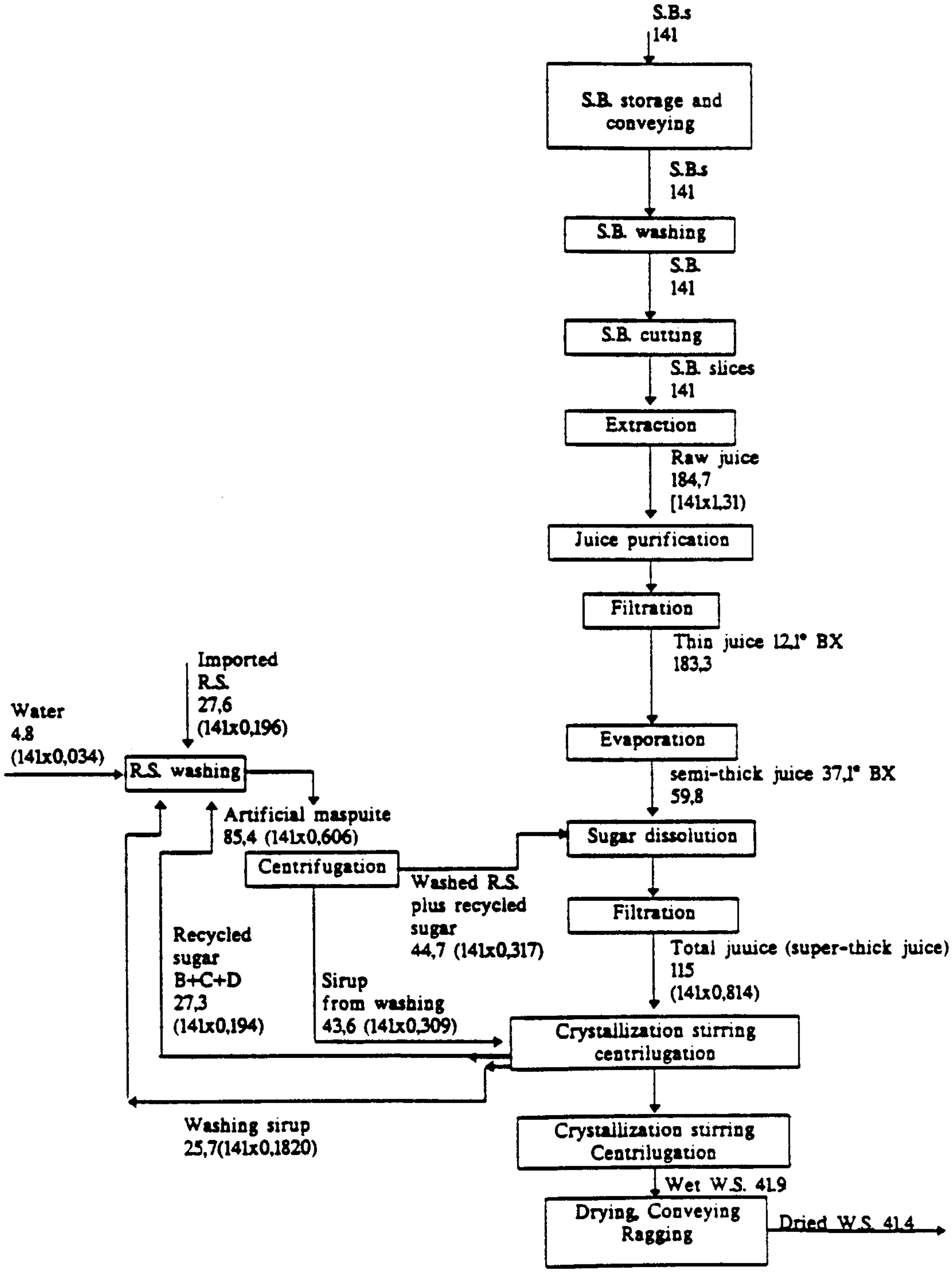


Table 5
Crystallization station
Product A vacuum pans

Production volume in the new process as against that of the previous technique.
How an approximate 3-fold production is possible, even by modest modification / revamping
(see also table 9.2, pag. 36, note 1, table 9.3, page 37, note 1, table 9.4, page 38, note 1 and page 12)

	a. New production volume as against the old one due only to increased process velocity in the new process		b. New production volume as against the old one due only to increased time utilization		c. New production volume as against the old one due only to equipment revamping	
	Beginning of campaign	**End of campaign**	**Beginning of campaign**	**End of campaign**	**Beginning of campaign**	**End of campaign**
1. New process	(1)3,5/2,5 (1)=1,4(2)	(1)3,5/3,0 (1)= 1,17(2)	73,6%	88,4%	1,33(4)	1,33(4)
2. Prev. technique	(1)3,5/3,5 (1) =1,0(2)	(1)3,5/5,5 (1)= 0,64 (2)	47,5%	75%	1,00(4)	1,00(4)
	(3) $\frac{1,4}{1,0}$ =1,4	(3) $\frac{1,17}{0,64}$ =1,83	$\frac{73,6}{47,5}$ =1,55	$\frac{88,4}{75,0}$ =1,18	$\frac{1,33}{1,00}$ =1,33	

Production volume in the new process
as against that of the previous technique
as a combined effect of all 3 above production parameters a, b, c.

Beginning of campaign	New process	: 1,4x0,736x1,33	= 1,37	1,37/0,475	= 2,88:1
	Prev. technique	: 1x0,475x1	= 0,475		
End of campaign	New process	: 1,17x0,884x1,33	= 1,375	1,375/0,48	= 2,86:1
	Prev. technique	: 0,64x0,75x1	= 0,48		

(1) Crystallization periods, 3,5/5,5 hours previous technique, 2,5/3,0 hours new process, begin/end campaign. For more information see page behind.
(2) Relative process velocities
(3) Relative velocities ratio
(4) Relative capacity in crystallization vacuum pans

More explanation on table 5 - Note (1)
(see 1st and 2nd column of upper part of table)

(1)	3,5/2,5:	crystallization cycle,	begin period,	prev. technique,	versus cryst. cycl.,	begin period	new process
	3,5/3,0:	“ “	“ “	“ “	“ “ “	end period	“ “
	3,5/3,5:	“ “	“ “	“ “	“ “ “	begin period	prev. technique
	3,5/5,5:	“ “	“ “	“ “	“ “ “	end period	“ “

Table 6

Calculation of critical raw sugar (R.S.) prices at which the new process may loose its economic superiority, over the previous technique (break-even-point)

Symbols:

Δ_2, Δ_1	Unit profit (before taxes) in drchs/kg W.S., referring to the new process and the previous technique respectively, for a totally non-depreciated plant (both the old equipment and the modification equipment are not depreciated).
Δ'_2, Δ'_1	Unit profit in drchs/kg W.S., referring to the new process and the previous technique respectively for a partly depreciated plant (old equipment depreciated, modification equipment not depreciated).
Q_2, Q_1	W.S. produced by either of the two technologies the new process and the previous technique.
C_2, C_1	Unit production-and-administration cost in drch/kg W.S., referring to the new process and the previous technique respectively, for a totally non-depreciated plant.
C'_2, C'_1	Unit production-and-administration cost in drch/kg W.S., referring to the new process and the previous technique respectively for a partly depreciated plant.

Numerical figures below refer to early '70s

1st Case-totally non-depreciated plant

The break-even-point whereby economic superiority of the process, compared to the previous technique, is reversed, is expressed by the equation:

$$\Delta_1 x Q_1 = \Delta_2 x Q_2$$

From above it is obtained that:

$$\Delta_2 = \Delta_1 x \frac{Q_1}{Q_2}$$

For previous technique (see also table 1, non-depreciated plant) it is:

$$\Delta_1 = 9{,}44 - C_1 = 9{,}44 - 8{,}223 = 1{,}217 \text{ drch/kg W.S.}$$

Therefore for the process it is:

$$\Delta_2 = 1{,}217 x \frac{Q_1}{Q_2} = 1{,}217 x \frac{25{,}74}{96{,}67} = 0{,}324 \text{ drch/kg W.S.}$$

Consequently the unit production-and-administration cost in the new process, at the break-even-point, will be:

$$C_2 = 9{,}44 - 0{,}324 = 9{,}116 \text{ drchs/kg W.S.}$$

Since it is (see also table 3 «Totally non-depreciated plant):

$$2{,}49+0{,}66X+0{,}289+0{,}046+0{,}158+0{,}070+0{,}297+0{,}556-0{,}891 = 9{,}116 \text{ drch/kg W.S.}$$

at the break-even-point, where 0,712 kg of X priced R.S. are required to produce 1 kg of W.S., it will finally be:

$$X = 9{,}174 \text{ drch/kg R.S.,}$$

which means that the R.S. must be priced at 9,174 drch/kg CIF production plant or that it will be ~ 97% of the W.S. price in order to have equal campaign profits in the two cases (B.E.P.).

By lower W.S. price, say of 8,7 drch/kg the break-even R.S. price results to be 8,453 drch/kg i.e. 89,5% on the W.S. price, which is again pretty close to W.S. price.

2nd Case-partly depreciated plant

From same equation as in beginning of case 1 (see symbols)

$$\Delta'_1 x Q_1 = \Delta'_2 x Q_2$$

it is obtained that $\Delta'_2=\Delta'_1 x \frac{Q_1}{Q_2}$

In the case of the previous technique it is (see also table 1, «Depreciated plant»)

$$\Delta'_1=9,44-C'_1= 9,44-6,585 = 2,855 \text{ drch/kg W.S.}$$

(the respective unit profit), where 6,585 drch/kg W.S. the corresponding production-plus-administration cost, with no depreciation burden for the old plant.

Therefore:

$$\Delta'_2=2,855 \text{ x} \frac{25,74}{96,67} = 0,760 \text{ drch/kg W.S.}$$

the respective unit profit in the case of the new process (at the break-even-point). Consequently the corresponding production-plus-administration cost in the new process will be:

$$C'_2=9,44-0,760=8,680 \text{ drch/kg W.S.}$$

at the break-even-point status.

Since it is (see also table 3 «Partly depreciated plant»)

$$2,49+0,665X+0,289+0,046+0,158+0,070+0,297+0,129+0,891 = 8,680 \text{ drch/kg W.S.}$$

in that point, where 0,712 kg of X' R.S. are required to produce 1 kg of W.S., it will finally be:

$$X'=9,160 \text{ drch/kg R.S.}$$

which means that the R.S. must be priced practically at same price as in the previous case, i.e. 9,160 drch/kg, CIF production plant, or that it should be ~97% of the W.S. price in order to have equal campaign profits in the two cases (B.T.P. or $\Delta'_1xQ_1=\Delta'_2xQ_2$).

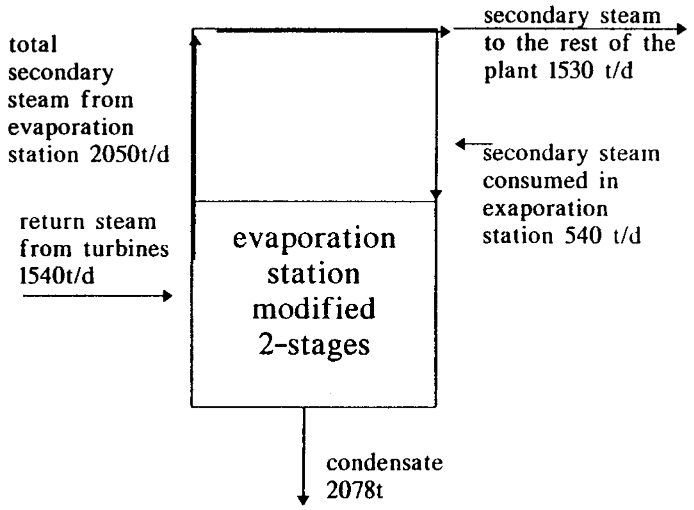
Table 7

STEAM BALANCE - DAILY BASIS CAMPAIGN AVERAGES

New Process

Steam rates corresponding to 2340 daily treated tones of sugar beets, in a modified (reduced) evaporation station, schematically shown below.

campaign average: 2340 tone/day (or 118% see also note (1) of table 2)



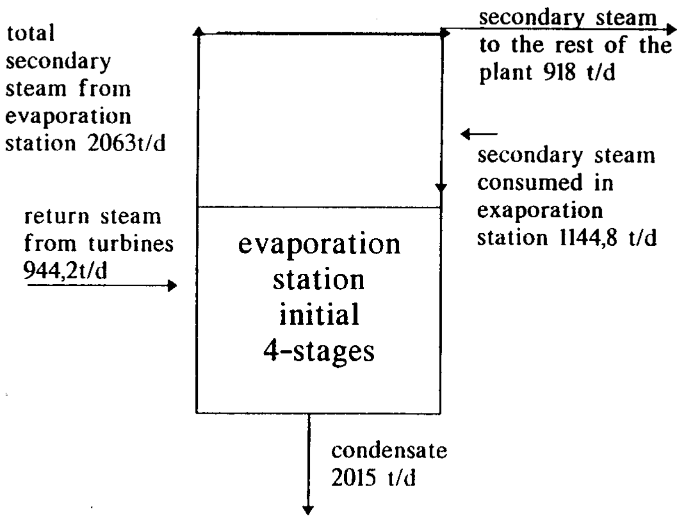
Steam to crystallization vacuum pans:

Product	A	550,2 t
product	B	63,4 t
product	C	18,8 t
product	D	5,2 t
Total		637,6 t

Previous Technique

Steam rates corresponding to 1983 daily treated tones of suga beets in the initial four-stageevaporations station, schematically shown below.

campaign average: 1983 tones/day (or 100%)



Steam to crystallization vacuum pans:

Product	A	258,0 t
product	B	18,8 t
product	C	6,2 t
Total		283,0 t

Table 8
Production Indicators

New Process, Treatment Scheme A, versus Previous Technique

(1) Beets entering the plant as a daily campaign average, in the previous technique (where production is bottlenecked by inherent process handicaps especially in the second half of the campaign). Taken equal to 100 (see also «Status of existing technology»). Also see note on page behind.

(2) Beets entering the plant as a daily campaign average in the new process, where production is no more bottlenecked, are taken equal to 118 (100x1,18).
The ratio 118:100 is from real campaign data where 118 indicates the beet throughput rates of the 10 best days of campaign, whereas 100 indicates the overall average of campaign. In the new process 118 throughput units, can be comfortably treated.

(3) Operating days in the case of previous technique: 110.

(4) In the case of the new process 132 days, whereby 132 represents a conservative estimation, (see text «Advantages of the invention» and «Brief presentation of the invention»).

Previous Technique

	Daily basis	Campaign basis
Sugar entering the plant	100x0,14(1) = 14,00	100x0,14x110 = 1540
Sugar recovered	14,00x0,843 = 11,80	1540x0,843 = 1298

New Process-Treatment scheme A

	Daily basis	Campaign basis
a) Sugar entering the plant with sugar beets	118(2)x0,14 = 16,52	118x0,14x132(3) = 2181
b) Sugar entering the plant with raw sugar	20,8x1,18x0,961 = 23,59	20,8x1,18x0,961x132 = 3113
Total sugar entering the plant	= 40,11	= 5294
Total sugar recovery	40,11x0,916 = 36,74	5294x0,916 = 4849

(1) Real sugar content of the beets in early '70s. By those years sugar beets had not the content of later years (14,5-16%) comparative picture, by accepting the 14% content does not alter but only marginally

(2) 118 = 1,18 x 100

(3) 132 operating days in the new process

New Process (scheme A) versus Previous Technique

	Daily basis	Campaign basis
Total sugar entering the plant	$\frac{newprocess}{previoustechnique} = \frac{40,11}{14,00} = 2,86$	$\frac{newprocess}{previoustechnique} = \frac{5294}{1540} = 3,44$
Total sugar recovered	$\frac{newprocess}{previoustechnique} = \frac{36,74}{11,8} = 3,11$	$\frac{newprocess}{previoustechnique} = \frac{4849}{1298} = 3.74$

Table 8'

Production Indicators

New Process, Treatment Scheme B versus Previous Technique

(same remarks and notes as with table 8)

Previous Technique

	Daily basis	Campaign basis
Sugar entering the plant	100x0,14 = 14,00	100x0,14x110 = 1540
Sugar recovered	14,00x0,843 = 11,80	1540x0,843 = 1298

New Process-Treatment scheme B

	Daily basis	Campaign basis
a) Sugar entering the plant with sugar beets	118x0,14 = 16,52	118x0,14x132 = 2181
b) Sugar entering the plant with raw sugar	19,6x1,18x0,961 = 22,23	19,6x1,18x0,961x132 = 2934
Total sugar entering the plant	= 38,75	= 5115
Total sugar recovery	38,75x0,902 = 34,95	5115x0,902 = 4614

New Process (scheme B versus Previous Technique

	Daily basis	Campaign basis
Total sugar entering the plant	$\frac{newprocess}{previoustechnique} = \frac{38,75}{14,00} = 2,77$	$\frac{newprocess}{previoustechnique} = \frac{5115}{1192} = 3,32$
Total sugar recovered	$\frac{newprocess}{previoustechnique} = \frac{34,95}{11,8} = 2,96$	$\frac{newprocess}{previoustechnique} = \frac{4614}{1298} = 3,55$

Table 9

ANALYTICAL INVESTIGATION OF THE OPERATIONS OF CRYSTALLIZATION STATION VACUUM PANS SECTION

9.1

PREVIOUS TECHNIQUE BEGINNING AND MAIN BODY OF THE CAMPAIGN PRODUCT A-3 VACUUM PANS

Engagement time of a vacuum pan (from real measurements during a campaign)		47,5%	
Engaged (productive) hours of a vacuum pan per day	0,475x24=	11,4	hours
Boiling cycle of vacuum pan (from real measurements excluding end period)		3,5	hours
Volume of boiled product of 92°BX leaving a vacuum pan		33,47	m^3
Weight of feed to a vacuum pan, if feed at 61°BX		73,05(1)	tones
Daily weight rate of feed to all product A vacuum pans (see table 1Π, process stream 1Π)		714,9	tones
Number of vacuum pans allocated to product A		3	
Daily feed to each product A vacuum pan		238,3	tones
Number of boiling cycles per vacuum pan per day	238,3/73,05=	3,26	
Therefore productive time of a vacuum pan per day	3,26x3,5=	11,41 or 47,5%	hours

(1) 92° BX → 1,494 gr/cm^3, 61°BX → 1,2924 gr/cm^3

$(33,47 . 1,494)+(X . 1)=(33,47+X) . 1,2924$

$X=23,05$ m^3, $33,47+X=56,52$ m3, $56,52 . 1,2924=73,05$ tones

9.2
PREVIOUS TECHNIQUE END OF CAMPAIGN PRODUCT A-3 VACUUM PANS

Engagement (productive) time of vacuum pans		74,96%(1) (see behind)
Engaged (productive) hours of a pan per day.	0,7496x24=	17,99(1) hours
Boiling cycle of a vacuum pan(1) (from real measurements during the end period)		5,5(1) hours
Volume of boiled product of 92°BX leaving a vacuum pan		33,47 m^3
Weight of above	33,47x1,494=	50,0 tones
Weight of feed(2) to a vacuum pan if feed at 54,8°BX*		81,01 tones (see behind)

(Thin juice by the end of campaign leaves the evaporation station at a thinner state taken at 52°BX. Therefore its daily rate will be raised to 520x60/52=600 (see table 1Π, stream αΠ).(3) Artificial masquite produced from recycled sugar dissolution, is of steady concentration, taken at 63,4°BX and of steady rate of 194,9 tones (table 1Π, stream γΠ). Therefore total rate 600+194,4=794,9 tones. Concentration of it will be (table 1Π, streams αΠ, γΠ, 1Π).

$$\frac{(520x\frac{60}{52}x52)+(194{,}9x63{,}4)}{794{,}9}=\frac{(600x52)+(194{,}9x63{,}4)}{794{,}9}=54{,}8^{o}BX\ )$$

Daily weight rate of feed to all product A vacuum pans (see above)	794,9 tones
Number of vacuum pans allocated to product A	3
Daily feed to each product A vacuum pan	265,0 tones

Number of boiling cycles per vacuum pan, per day	265/81,01=	3,27
Therefore productive time of a vacuum pan per day	3,27x5,5=	17,99(1) or 74,96%

(1) These data are experimental data based on the largely predominant boiling cycle of 5,5 hours. Substantial deviations, however, of actual boiling cycles have also been observed towards higher figures, say 6, 7 e.t.c. hours, forcing product A vacuum pans to operate at higher utilization rates, which sometimes may approach the 100% utilization degree.

(2) 92°BX → 1,494 gr/cm^3, 54,8°BX → 1,2564 gr/cm^3

$(33,47\ .\ 1,494) + (X\ .\ 1) = (33,47+X)\ .\ 1,2564$

$X=31,01$ m^3, $33,47+X=64,48$ m^3, $64,48\ .\ 1,2564 = 81,01$ tones

(3) More accurate calculations, taking into account, sugar beet throughput decrease at the end of campaign, might also be presented. These would result in a slightly lower engagement time (than 74,96%). To avoid further complication of calculations this corrective imput is ommitted as it is not but of minor importance.

9.3

NEW PROCESS
BEGINNING AND MAIN BODY OF CAMPAIGN
PRODUCT A-4 VACUUM PANS

Engagement (productive) time of vacuum pans (calculated, see below)		73,6%	
Engaged (productive) hours of vacuum pans per day	0,736x24=	17,67	hours
Boiling cycle of a vacuum pan, estimated		2,5	hours(1)
Volume of boiled product of 92°BX leaving a vacuum pan		33,47	m^3
Weight of above	33,47x1,494=	50,0	tones
Weight of feed to a vacuum pan if feed at 66,5°BX (numerical example 1)		67,34(2)	tones
Daily weight rate of feed to all product A vacuum pans if feed at 66,5°BX (total feed from dissolution of imported R.S. to semi-thick juice plus dissolution of recycled plant sugar to same, see table 1A, stream 12A)		1904,9	tones
Sugar weight rate conveyed by the above (table 1A, stream 12A)		1208,0	tones
Number of vacuum pans allocated to product A (3 pans from initial plant plus one added)		4(3)	
Daily feed to each product A vacuum pan		476,2	tones
Number of boiling cycles per vacuum pan per day	476,2/67,34=	7,07	
Therefore productive time of a vacuum pan per day	7,07x2,5=	17,7	hours or 73,6%

(1) The total feed to product A vacuum pans results substantially upgraded if compared to that of the previous technique. (see behind)

	New Process	Previous Technique
Concentration	66,5°BX	62-55 °BX
Purity	95,5%	91-88%

The boiling cycle in the main part of campaign, in the previous technique is at 3,5 hours with usual juices. In refineries (where only raffinades are produced, not W.S.) boiling cycles of 1,5-2,0 hours are encountered. In our process, with qualitative characteristics, not far below those of sugar refineries, we have assumed a boiling cycle of 2,5 hours. We think this is a conservative assumption.

(2) 92°BX → 1,494 gr/cm^3, 66,5°BX → 1,3255 gr/cm^3

$(33{,}47 . 1{,}494) + (X . 1) = (33{,}47+X) . 1{,}3255$

$X=17{,}33$ m^3, $33{,}47+X=50{,}8$ m^3, $50{,}8 . 1{,}3255 = 67{,}34$ tones

(3) If, instead of 4 vacuum pans, we had used 3 of them, as in the previous technique, the engagement time would have been 73,6x4/3=98,1% which means that 4 vacuum pans are practically indispensable.

9.4
NEW PROCESS
END OF CAMPAIGN
PRODUCT A-4 VACUUM PANS

Engagement (productive) time of vacuum pans, calculated (see below)		88,38%	
Engagement (productive) hours of vacuum pans	0,8838x24=	21,21	hours
Boiling cycle of a vacuum pan, estimated		3,0(1)	hours
Volume of boiled product of 92°BX leaving a vacuum pan		33,47	m^3
Weight of above	33,47x1,494=	50,0	tones
Weight of feed to a vacuum pan if feed at 66,5°BX (numerical example 1)		67,34(2)	tones
Daily weight rate of feed to all product A vacuum pans, if feed at 66,5°BX (total feed from dissolution of imported R.S. to semi-thick juice plus dissolution of recycled plant sugar to same, see table 1A, stream 12A)		1904,9	tones
Number of vacuum pans allocated to product A		4(3)	
Daily feed to each product A vacuum pan		476,2	tones
Number of boiling cycles per vacuum pan per day	476,2/67,34=	7,07	
Therefore production time of a vacuum pan per day	7,07x3,0=	21,21 88,38%	hours or

(1) This is an unfavorable estimation.
In the previous technique with a thick juice concentration of 60-62°BX and a purity of 90-92%, in the main part of the campaign, product A boiling cycles of 3.5 hours are experienced. (see behind)

In the new process, end period of campaign, with a total feed of 66,5°BX and a purity slightly only inferior to 95% (due to the strongly stabilizing effect of the R.S.) we unfavorably estimated a boiling cycle in the range of 3,0 hours.

(2) Calculation as in note (2) of table 9.3.

(3) 4 vacuum pans are sufficient for the end period requirements. Should, however, longer boiling cycles than the 3,0 hours estimated be experienced, so a 5th vacuum pan might come into consideration. Our conviction, anyhow, is that boiling cycle requirements at the end of campaign, will be lower than the 3 hours used considering the technical data of the total juice as compared to those of the thick juice of the previous technique (new process end period: concentration 66,5°BX, purity slightly inferior to 95% as against concentration of 60-62°BX and purity of 90-92% in the main part of campaign of the previous technique requiring 3,5 hours of boiling cycle).

1 of 2

TABLE 1Π
PREVIOUS TECHNIQUE
PROCESS STREAMS AND THEIR PROPERTIES

				PRODUCT A							PRODUCT B		
	αΠ	βΠ	γΠ	1Π	5Π	7Π	4Π	6Π	8Π	9Π	12Π	14Π	13Π
	Thick juice	Water for dissolution of recycled plant sugar	Artificial masquite (dissolved recycled sugar)	Total feed to crystallization	Masquite A	Vapor from masquite A	Steam to product A crystallization	Product A Condensate	Sugar A	Syrup A	Masquite B	Vapor from masquite B	Steam to product B crystallization
tones/day	520,0	68,7	194,9	714,9	467	247,9	258	258	234,0	233,0	220,7	12,3	12,6
°BX	60,1		63,4	61,0	93,0				98,8	88,1	93,0		
Pol	54,1		61,5	56,1	85,9				98,3	73,5	77,6		
Q	90,0		97,1	92,0	92,0				99,5	83,5	83,5		
solid matter tones/day	312,8		123,5	436,4	436,4				231,2	205,2	205,2		
sugars tones/day	281,5		119,9	401,4	401,4				230,1	171,3	171,3		
non-sugars tones/day	31,3		3,7	35,0	35,0				1,1	33,9	33,9		
water tones/day	207,2	68,7	71,4	278,5	30,6	247,9	258	258	2,8	27,8	15,5	12,3	12,6

2 of 2

TABLE 1Π
PREVIOUS TECHNIQUE
PROCESS STREAMS AND THEIR PROPERTIES

PRODUCT B			PRODUCT C							
13Π	15Π	16Π	18Π	20Π	17Π	19Π	21Π	28Π	29Π	30Π
Steem to Product B crystallization	Sugar B	Syrup B	Masquite C	Vapor from masquite C	Steam to product C crystallization	Product C condensate	Sugar C	Molasses	Recycled plant Sugar B+C	Total vapors from vacuum pans of crystallization stations
12,6	81,2	139,5	133,6	5,9	6,2	6,2	45,0	88,6	126,2	266,1
	98,0	90,0	94,0				97,5	92,2	97,8	
	96,5	66,7	69,6				92,2	58,1	95,0	
	98,5	74,0	74,0				94,5	63,0	97,0	
	79,6	125,6	125,6				43,9	81,7	123,5	
	78,4	93,0	93,0				41,5	51,5	119,9	
	1,2	32,6	32,6				2,4	30,2	3,6	
12,6	1,6	13,9	8,0	5,9	6,2	6,2	1,1	6,9	2,7	266,1

1 of 3

TABLE 1A

NEW PROCESS

TREATMENT SCHEME A - DIRECT DISSOLUTION OF IMPORTED RAW SUGAR TO SEMI-THICK JUICE PROCESS STREAMS AND THEIR PROPERTIES

SEMI-THICK JUICE-IMPORTED RAW SUGAR AND RECYCLED PLANT SUGAR DISSOLUTION

	1A	2A	3A	4A	5A	6A	7A	8A	9A	10A	11A
	Thin juice	Return steam from turbines to evaporation station	Semi-thick juice 37,1°Bx	Evaporation station condensate	Total vapor from evaporation station	Vapor from evaporation consumed locally	Vapor from evaporation to the rest of the plant	canceled	canceled	Imported raw sugar	Recycled plant sugar B+C+D
tones/day	3042	1065	992	2078	2050	1013	1036			487	425,9
°BX	12,1		37,1							98,9	98,2
Pol	10,9		33,4							96,1	95,4
Q	90		90							97,2	97,1
solid matter tones/day	368		368							481,6	418,4
sugars tones/day	331,5		331,5							468	406,5
non-sugars tones/day	36,5		36,5							13,6	11,9
water tones/day	2674		624	2078	2050	1013	1036			5,4	7,5

2 of 3

TABLE 1A

TREATMENT SCHEME A - DIRECT DISSOLUTION OF IMPORTED RAW SUGAR TO SEMI-THICK JUICE PROCESS STREAMS AND THEIR PROPERTIES

	CRYSTALLIZATION PRODUCT A						CRYSTALLIZATION PRODUCT B			
12A	13A	14A	15A	16A	17A	18A	19A	20A	21A	22A
Total feed to crystallization	As stream 12A decolared	As stream 12A filtered	Masquite A	Vapor from mesquit A	Sugar A	Syrup A	Masquite B	Vapor from masquite B	Sugar B	Syrup B
1904,9	Same	Same	1378	526,9	732,3	645,7	584,7	61	262,1	322,6
66,56	properties	properties	92		98,8	84,3	93,1		98,5	88,7
63,3	except	except	87,5		98,5	75,1	82,9		97,3	71,2
95,1	colour	colour	95,1		99,7	89,0	89,0		98,9	80,3
1268		and clarity	1268		723,5	544,5	544,5		258,2	286,3
1206			1206		721,3	484,7	484,7		255	229,7
62			62		2,2	59,8	59,8		3,2	56,6
636,9			110	526,9	8,8	101,2	40,2	61	3,9	36,3

3 of 3

TABLE 1A

NEW PROCESS

TREATMENT SCHEME A - DIRECT DISSOLUTION OF IMPORTED RAW SUGAR TO SEMI-THICK JUICE

PROCESS STREAMS AND THEIR PROPERTIES

CRYSTALLIZATION		PRODUCT C		CRYSTALLIZATION		PRODUCT D		
23A	24A	25A	26A	27A	28A	29A	30A	11A
Masquite C	Vapor from masquite C	Sugar C	Syrup C	Masquite D	Vapor from masquite D	Sugar D	Molasses	Recycled plant sugar B+C+D
304,5	18,1	100,6	203,9	198,7	5,2	63,2	135,5	425,9
94,0		98,0	92,0	94,4		98,6	92,5	98,4
75,4		96,9	64,8	66,5		88,1	56,4	95,2
80,3		98,9	70,4	70,3		89,4	60,9	97,4
286,3		98,6	187,7	187,7		62,3	125,4	419,1
229,7		97,5	132,1	132,1		55,7	76,4	408,2
56,6		1,1	55,6	55,6		6,6	49,0	10,9
18,2		2,0	16,2	11,0		0,9	10,1	6,8

1 of 4

TABLE 2B
NEW PROCESS
TREATMENT SCHEME B-WASHING (AFFINATION) OF RAW SUGAR PRIOR TO DISSOLUTION TO SEMI-THICK JUICE
PROCESS STREAMS AND THEIR PROPERTIES

	IB	**2B**	**3B**	**4B**	**5B**	**6B**	**7B**
	Thin juice	**Return steam from turbines to evaporation station**	**Semi-thick juice 37,1°Bx**	**Condensate**	**Total vapor from evaporation station**	**Vapor from evaporation station consumed locally**	**Vapor from evaporation station to the rest of the plant**
tones/day	3042	1065	992	2078	2050	1013	1037
°BX	12,1		37,1				
Pol	10,9		33,4				
Q	90		90				
solid matter tones/day	368		368				
sugars tones/day	331,5		331,5				
non-sugars tones/day	36,5		36,5				
water tones/day	2674	1065	624	2078	2050	1013	1037

TABLE 2B

NEW PROCESS

TREATMENT SCHEME B - WASHING OF RAW SUGAR PRIOR TO DISSOLUTION TO SEMI-THICK JUICE

PROCESS STREAMS AND THEIR PROPERTIES

	WASHING OF RAW SUGAR								DISSOLUTION OF RAW SUGAR	
	12B	48B	37B	15B	16B	17B	17'B	25B	17B	18B
	Imported raw sugar	Recycled plant sugar B+C+D	Washing syrup	Dissolution water to artificial measure	Artificial masquite	Washed off sugar	Rinsing water to centrifugal	Washing syrup from washed raw sugar	Washed raw sugar (repeated)	Total thick juice feed to crystallization station
tones/day	459,45	453,45	426,0	79,6	1418,5	742,3	56,7	723,9	742,3	1734,3
°BX	98,6	98,6	87,0		89,6	97,4		74,8	97,4	62,9
Pol	95,35	96,15	73,6		83,75	97,2		63,7	97,2	60,7
Q	96,8	97,6	84,5		93,5	99.8		85,1	99,8	96,5
solid matter tones/day	453,1	446,9	370,9		1270,9	722,8		548,1	722,8	1090,8
sugars tones/day	438,5	436,0	313,5		1188,0	721,4		466,6	721,4	1052,9
non-sugars tones/day	14,6	10,9	57,4		82,9	1,4		82,0	1,4	37,9
water tones/day	6,35	6,55	55,1	79,6	147,6	19,5	56,7	184,8	19,5	643,5

TABLE 2B

NEW PROCESS

TREATMENT SCHEME B - WASHING OF RAW SUGAR PRIOR TO DISSOLUTION TO SEMI-THICK JUICE

PROCESS STREAMS AND THEIR PROPERTIES

CRYSTALLIZATION - PRODUCT A

CRYSTALLIZATION - PRODUCT B

22B	23B	24B	27B	26B	25B	28B	22'B	31B	32B	33B	34B
Masquite A	Sugar A	Total syrup A	Syrup A to raw sugar washing	Balance of syrup A to product B crystalliz.	Syrup from raw sugar washing	Combined syrup to product B crystalliz.	Vapor from masquite A	Masquite B	Vapor from masquite B	Sugar B	Total syrup B
1174,3	694,7	479,6	213,0	266,6	732,9	999,5	560,0	829,9	169,6	288,7	541,2
92,9	98,8	84,3	84,3	84,3	74,8	77,3		93,1		98,8	90,1
89,65	98,5	76,8	76,8	76,8	63,7	67,2		80,9		97,6	71,9
96,5	99,7	91,1	91,1	91,1	85,1	86,8		86,9		98,8	79,9
1090,8	686,5	404,3	179,5	224,8	548,1	772,9		772,9		285,2	487,7
1052,9	684,4	368,5	163,7	204,8	466,6	671,4		671,4		281,8	389,6
37,9	2,1	35,8	15,8	20,0	81,5	101,5		101,5		3,4	98,1
83,5	8,2	75,3	33,5	41,8	184,8	226,6	560,0	57,0	169,6	3,5	53,5

TABLE 2B

NEW PROCESS

TREATMENT SCHEME B - WASHING OF RAW SUGAR PRIOR TO DISSOLUTION TO SEMI-THICK JUICE

PROCESS STREAMS AND THEIR PROPERTIES

CRYSTALLIZATION - PRODUCT B CRYSTALLIZATION - PRODUCT C CRYSTALLIZATION - PRODUCT

36B	35B	32B	35'B	40B	41B	46B	41'B	44B	45B	47B (1)
Syrup B to raw sugar washing	Balance of syrup B to product B crystalliz.	Vapor from masquite B	Masquite C	Sugar C	Syrup C to product D crystalliz/	Vapor from masquite C	Masquite D	Sugar D	Molasses	Vapor from masquite D
213	328,2	169,6	314,6	101,75	212,85	13,6	207,9	63,1	144,8	4,95
90,1	90,1		94,0	97,9	92,2		94,4	98,6	92,6	
71,9	71,9		75,0	96,9	64,5		66,1	88,2	56,5	
79,8	79,8		79,8	99,0	70,0		70,0	89,5	61,0	
191,9	295,8		295,8	99,6	196,2		196,2	62,1	134,1	
153,1	236,0		236,0	98,6	137,4		137,4	55,6	81,8	
38,8	59,8		59,8	1,0	58,8		58,8	6,5	52,3	
21,1	32,4	169,6	18,8	2,15	16,65	13,6	11,7	1,0	10,7	4,95

(1) Total recycled plant sugar B+C+D under stream number 48B, sheet 2 of 4.

2. CLAIMS

1. A process for white sugar production, whereby the thin juice is obtained as per the usual procedures of the previous technique and, wherein, that juice is thereafter concentrated in two distinct, consecutive steps; one first step by evaporation, utilizing return or exhaust steam from the electricity generating unit of the plant, reaching a moderate concentration level conveniently set in the range of 37° Bx; and one second step, by dissolution in the above juice, of proper quality raw sugar, reaching a final concentration level conveniently set in the range of 63-68° Bx, most preferably 65-67° Bx; and whereby manufacturing process, from this point on, proceeds on a typically similar pattern, as in the previous technique, which, however, is profoundly differentiated as far as its technical, production and productivity aspects are concerned, by virtue of the far reaching repercussions of the change at the evaporation station; which repercussions further dictate a purposeful reinforcement of process equipment in sugar dissolution crystallization and transportation apparatus to cope with the production needs in their totality.
2. A process as per claim 1, whereby the bottleneck situation, created around the evaporation station in late campaign in the preceding technique, wherein thinner, darker and durtier, and above all, quantitatively inferior thick juice is produced, is eliminated; this occurring thanks to the breakdown of the total previous concentration process, declining with time, into the two distinct phases of claim 1, each one of which is a fairly stable, equally speedy at all times procedure, therefore into an overall fairly independent from time, procedure.
3. A process as per claim 1, whereby the number of effects in the evaporation station is reduced to 2 only, as against the usually 4 effects of the previous technology.
4. A process as per claims 1 and 3, whereby the moderately concentrated juice of 37° Bx, or so, is decisively lighter coloured, as compared to the previous technique thick juice -the colour results being expressed at equal Bx levels- due to the essentially shorter residence time in the 2-effect evaporation station.
5. A process as per claims 1 and 3, whereby the 37° Bx, or so, juice, is further lighter coloured, compared to the previous technique thick juice, due to the

lower temperature profile, applicable in the 2-effect station, starting from temperatures as low as 125-120°C and even below that, sufficing to produce the right temperature levels at the exit secondary steam from this station, i.e. the second effect produced secondary steam, which is in the close vicinity of 100°C, on either side of it, according to the particular, case-by-case requirements.

6. A process as per claim 1, whereby the added rate of raw sugar, being a non-end-product (proper quality) raw sugar, affinated if necessary, is in the range of 20% wt on beets, carrying a sugar load 1,3 - 1,4 times (130-140%) the sugar load carried by the sugar beets themselves, co-treated in the same process, this imparting a deep correction to the quality and technical characteristics of the total blended stuff processed; i.e. a higher concentration and a higher purity, being set conveniently in the range of 63-68° Bx, most preferably 65-67° Bx, the former, and at 94-97%, preferably 95-96% the latter, maintained practically or literally constant to the last campaign day; as against the 60-62° Bx thick juice concentration, tending to decline, to even below 55° Bx, and 90-92% thick juice purity tending to decline to even below 87-88% in the previous technique in late campaign.
7. A process as per claims 1 and 6, whereby the other major bottleneck, created in the previous technique in late campaign around the crystallization station, mainly as a result of the longer boiling cycles at this station vacuum pans, because of the degraded beet quality, is also eliminated. The boiling cycles at product A vacuum pans being, as per the improved technical properties in the new process, at 3,0 hours in late campaign, as against 5,5 hours, and more, in the previous technique; and the boiling cycles of the other sugar products of sugar grades being also shorter in proportion (see «Brief presentation of the invention» pages 5-7 in the text, also tables 9.1, 9.2, 9.3, 9.4 pages 34-38; further tables 8, 8", pages 31-33).
8. A process as per claim 1 and 7, whereby colour status is further improved, as compared to the respective colour of the previous technique, as a result of the eminently shorter boiling times at the crystallization station vacuum pans, especially in late campaign.
9. A process as per claims 1, 2, 6 and 7, whereby all related process and production figures, expressed as daily averages over the whole campaign, as against the previous-technique-associated respective figures, are increased. So:

a) the beet throughput and the associated polarimetric sugar contained in that, being in the range of 1,15 - 1,25 times (115-125%) the previous rate (see table 8, note 2, page 31)

b) the total polarimetric sugar input to the plant for processing (sugar beets plus raw sugar) being in the range of 2,7 - 2,9 times (270-290%) the previous rate, (see tables 8 and 8", «new process» rates, compared to the «previous technique» ones. Also table 1Π, column αΠ, table 1A, column 1A and 10A, and table 2B, column 1B and 12B).

c) the white sugar recovered from processing, being in the range of 2,9-3,1 times (290-310%) the previous rate, as recovery yields by the process are superior to the previous technique respective yields (see table 1Π, 1A, and 2B, besides the above mentioned columns also columns 8Π, 17A and 23B respectively).

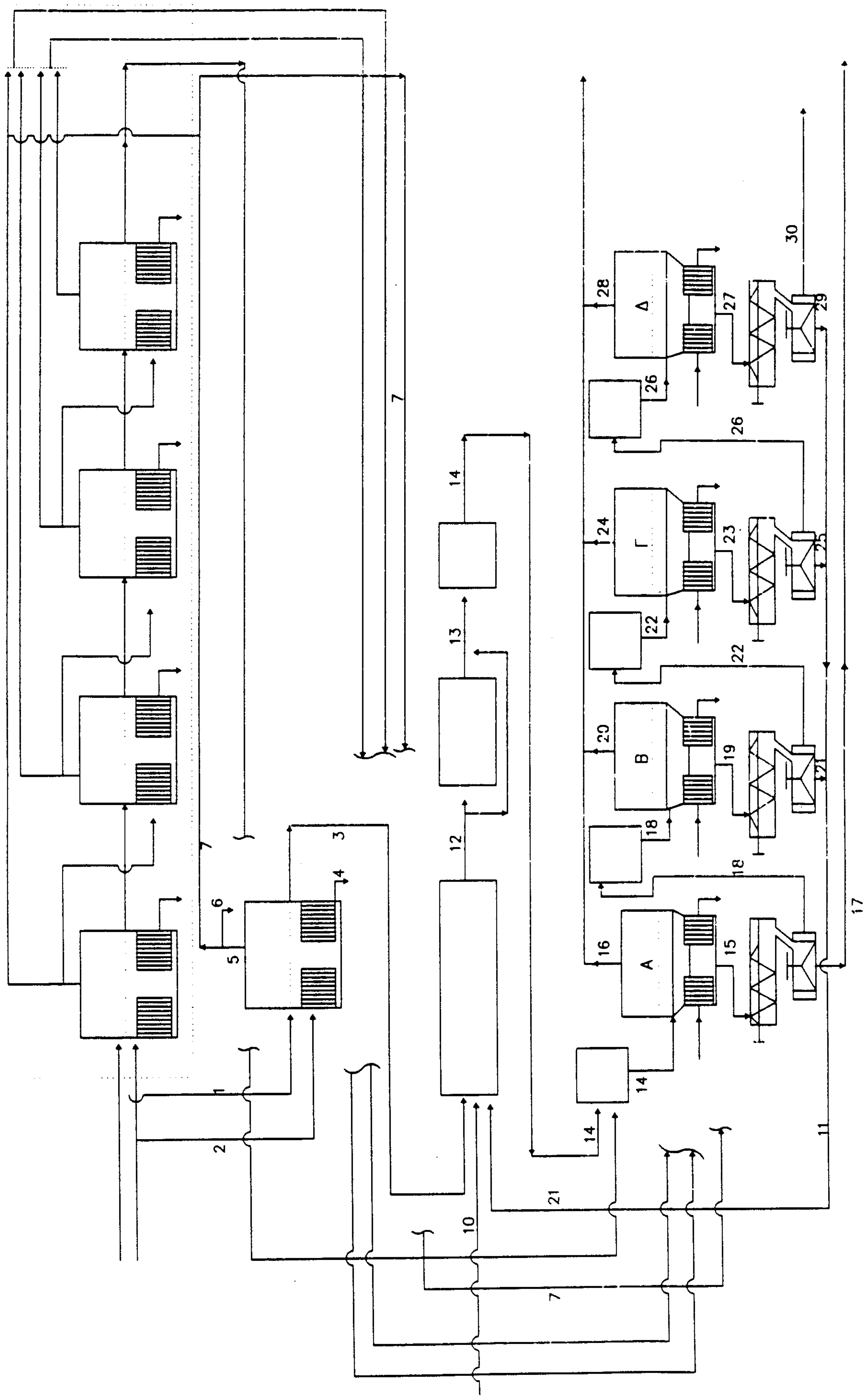
d) the campaign period being extensible by an estimated 20-30% on profit making terms (see page 4, par.3).

10. A process as per claims 1, 2, 6, 7 and 9, whereby daily increased process and production rates, claimed previously, are achieved by a modest modification / partial revamping of the process apparatus, being in the estimated order of 24-28%, most probably 25-27% of the old apparatus value, as a result of the combined higher process velocities and the reduced idle times in batch operating units of the plant, plus the fact that apparatus modification / reinforcement is limited to a number of units only of the plant (see table 5, page 26).

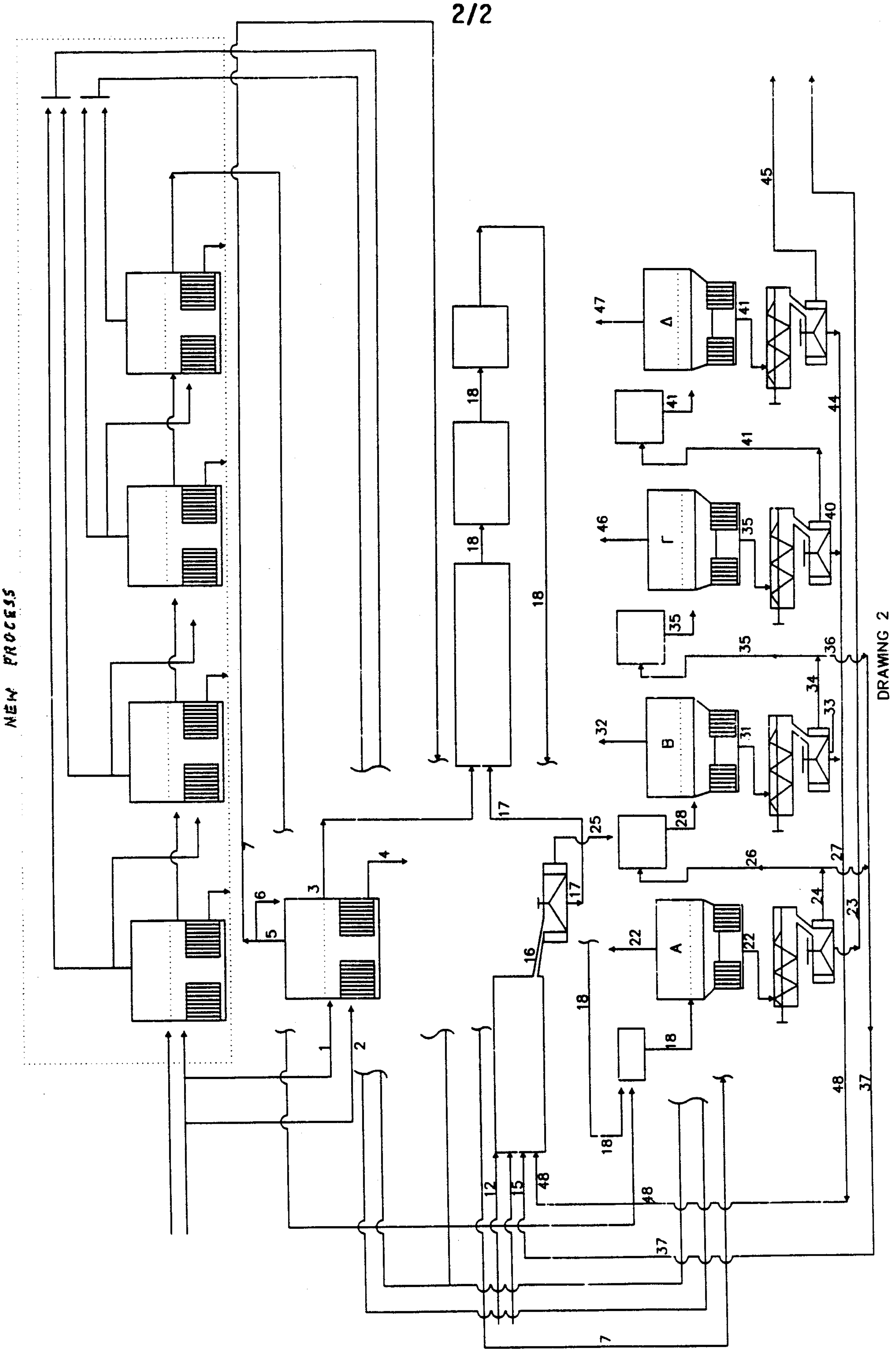
11. A process as per all previous claims, being implemented by either of the following two schemes: a) Scheme A, whereby the outside raw sugar is added directly, i.e. without previous affination, to the moderately concentrated juice of 37° Bx or so and b) Scheme B, whereby the raw sugar is added to the 37° Bx, or so, juice, after being affinated, to remove the excess invert sugar present, as in the case of a sugar-cane-origin raw sugar.

12. A process as per all previous claims being differentiated from previously devised co-treatment processes, in that those processes called for addition of the raw sugar in the juice purification phase, with the raw-sugar-enriched juice, so formed, having to be concentrated by evaporation to 60° Bx, thus not avoiding the disadvantages characterizing the previous technique (see page 3, upper part, of the text).

NEW PROCESS



DRAWING 1


NEW PROCESS
1
2
3
4
5
6
7
12
15
16
17
18
22
23
24
25
26
27
28
A
B
Γ
Δ
31
32
33
34
35
36
37
40
41
44
45
46
47
48
DRAWING 2

INTERNATIONAL SEARCH REPORT

Inte ional Application No
PCT/GR 98/00005

A. CLASSIFICATION OF SUBJECT MATTER
IPC 6 C13D1/00 C13F1/02 C13F1/00

According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)
IPC 6 C13D C13F

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the international search (name of data base and, where practical, search terms used)

C. DOCUMENTS CONSIDERED TO BE RELEVANT

Category °	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
A	EP 0 737 753 A (ERIDANIA) 16 October 1996 see the whole document ---	1-12
A	EP 0 110 315 A (DANSKE SUKKERFABRIKKER) 13 June 1984 see the whole document ---	1-12
A	GB 2 287 952 A (IRISH SUGAR RESAEARCH AND DEVELOPMENT) 4 October 1995 see the whole document ---	1-12
A	DE 27 29 192 B (SÜDDEUTSCHE ZUCKER) 2 November 1978 see the whole document -----	1-12

☐ Further documents are listed in the continuation of box C.

☒ Patent family members are listed in annex.

° Special categories of cited documents :

"A" document defining the general state of the art which is not considered to be of particular relevance
"E" earlier document but published on or after the international filing date
"L" document which may throw doubts on priority claim(s) or which is cited to establish the publication date of another citation or other special reason (as specified)
"O" document referring to an oral disclosure, use, exhibition or other means
"P" document published prior to the international filing date but later than the priority date claimed

"T" later document published after the international filing date or priority date and not in conflict with the application but cited to understand the principle or theory underlying the invention
"X" document of particular relevance; the claimed invention cannot be considered novel or cannot be considered to involve an inventive step when the document is taken alone
"Y" document of particular relevance; the claimed invention cannot be considered to involve an inventive step when the document is combined with one or more other such documents, such combination being obvious to a person skilled in the art.
"&" document member of the same patent family

Date of the actual completion of the international search 18 June 1998	Date of mailing of the international search report 26/06/1998
Name and mailing address of the ISA European Patent Office, P.B. 5818 Patentlaan 2 NL - 2280 HV Rijswijk Tel. (+31-70) 340-2040, Tx. 31 651 epo nl, Fax: (+31-70) 340-3016	Authorized officer Van Moer, A

INTERNATIONAL SEARCH REPORT

.nformation on patent family members

Intei ɔnal Application No
PCT/GR 98/00005

Patent document cited in search report		Publication date	Patent family member(s)			Publication date
EP 737753	A	16-10-1996	IT	MI950777	A	14-10-1996
EP 110315	A	13-06-1984	DK	523782	A	25-05-1984
			DE	3377144	A	28-07-1988
			US	4534800	A	13-08-1985
GB 2287952	A	04-10-1995	IE	940281	A	04-10-1995
DE 2729192	B	02-11-1978	AT	357122	B	10-06-1980
			BE	867938	A	02-10-1978
			DK	202478	A	29-12-1978
			FR	2396084	A	26-01-1979
			GB	1603679	A	25-11-1981
			GB	1603678	A	25-11-1981
			NL	7806914	A	02-01-1979
			SE	426360	B	17-01-1983
			SE	7805358	A	29-12-1978
			US	4290818	A	22-09-1981