
DESIGN APPROACH FOR SMALL SCALE CONTINUOUS, STEADY STATE PLANTS

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ABSTRACT

One of the major steps in the design of a process plant is development of an appropriate flowsheet. At this step, the technical solutions may be considered in relation to the social and economic factors. In this paper, guidelines for developing a flowsheet for continuous, steady state process in a developing country environment have been proposed. To illustrate the proposed guidelines, a case study involving design of an ethanol distillation plant has been used.

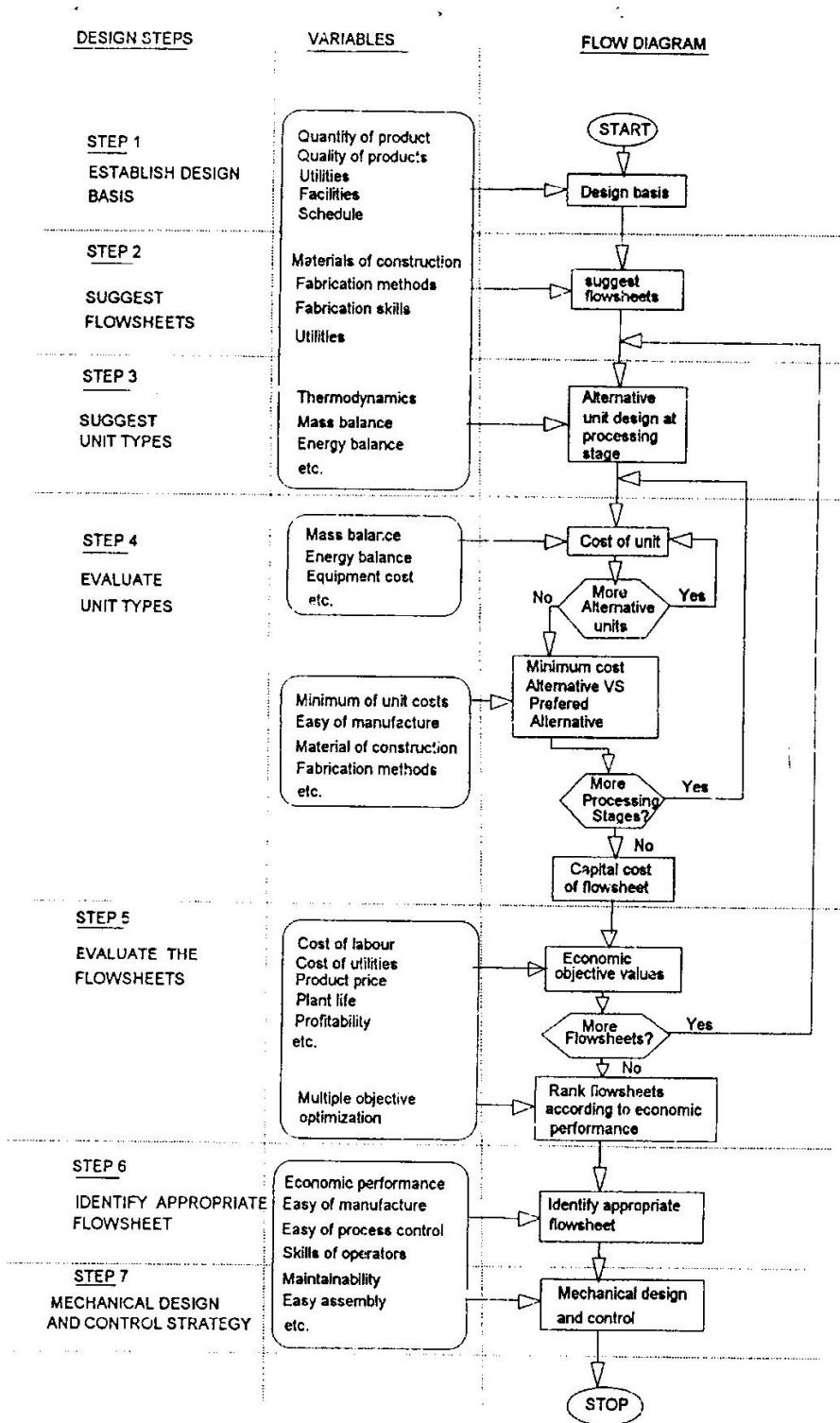
INTRODUCTION

Design of a continuous, steady state plant such as distillation plant involves deciding the type of unit for each processing stage and their interconnections, size and shape so as to produce the required product at required quality and rate. The total cost must be acceptably low and plant should operate satisfactorily.

A general design approach as proposed by Westerberg et. al.[1] involves three steps; first the synthesis, where flowsheets are chosen, the second is analysis of the synthesized flowsheet and the third is optimisation. Synthesis, analysis and optimization have reached a high level of sophistication in recent years largely because of the need for simulation of complex plants.

Developing countries however need simple plants for processing of agricultural raw materials. These makes use of the well known processing techniques (e.g. milling, fermentation, distillation). Thus as opposed to the synthesis problem of generating an entirely new flowsheet, the prob-

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lem in developing countries is more about the choice of an appropriate flowsheet from among the alternatives, and design of the individual process units taking into account of the local circumstances such as availability of materials of construction and construction methods, level of skills of the operators and more of such variables. It is the intention of this paper to propose design guidelines for such an approach, and to illustrate it with a case study.

PROPOSED DESIGN APPROACH.

This paper proposes to approach the design of a small scale continuous steady state plant as a sequence of several interacting activities (Fig 1). In this figure the left column shows the design steps. The middle column lists examples of variables to consider for each design step. The right hand side column shows a flow diagram as a series of design decisions inter-linked with feedback loops. A further elaboration of the steps in the proposed design process is as follows;

Step 1. Establish design basis

The design process will first require specification of quantity and quality of the finished product. There are no "golden rules" as to how these values should be chosen, instead understanding of the prevailing conditions in the environment in which the plant will operate plays a very important role. Information concerning the demand of finished product should be gathered through consultation with the population (end users, policy makers, financiers). In addition some knowledge about the raw material availability and quality, utilities on site, operating schedule should also be considered.

Step 2. Suggest flowsheets.

There could be several ways of transforming the raw material into finished product. Having decided on the product specifications, possible flowsheets are proposed at this step. Some alternatives can be dropped at this step by judging from past experience, heuristics or thermodynamic arguments.

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Step 3. Suggest unit types.

One or more alternative designs of accomplishing a given processing task in the flowsheet may be possible. At this step these alternatives should be listed down.

Factors such as availability of materials of construction, methods of fabrication, and easy maintenance need to be considered in making suggestions and short listing the alternatives.

Step 4. Evaluate unit types.

Mass and energy balance calculations enable specification of the expected performance of unit at each processing stage. Capital cost estimates for alternative types of unit should then be worked out. This saves as one of the initial criteria for screening the alternatives. However other variables which may not easily be quantified such as, ease of manufacture, cost of maintenance and replacement must be taken into account to finalise the choice of type of unit to use for each processing stage.

Step 5. Evaluate the flowsheets.

In step 4, the type of unit to use at each processing stage and capital cost associated with the unit would have been worked out. At this step (step 5) capital cost for each alternative flowsheets is estimated by taking the sum of cost of individual units. Also other economic factors of interest should be considered at this stage. These may include the operating cost, annualized cost or profitability such as the percentage rate of return of investment associated with each flowsheet.

This step should be concluded by ranking the flowsheet alternatives in the order of decreasing economical attractiveness. In the event that there is no flowsheet that is better than the other in every economic objective then multiple objective optimization methods[3,4] may save as useful tools for ranking alternatives.

Step 6. Identify appropriate flowsheet.

The most economical flowsheet as found from the ranking at step 5 may

not necessarily be the most appropriate alternative when taking into account the total environment. Thus in addition to the economic objectives the designer should at this stage take into account the social implications the proposed flowsheet will have on the society.

To be considered here are factors such as safety of the plant, required skill to operate the plant, maintenance requirements, process control and more of such factors. Here too, the problem may be viewed as a multiple objective optimization problem. This step may require application of optimization technique based on qualitative factors rather than quantitative methods. Examples of such methods have been discussed by Sfeir-Younis and Bromley[4].

Step 7. Mechanical design and control strategy

Physical dimensions of every unit and the structure of the entire plant are presented in the form of design drawings which can be used for manufacturing and installation of the plant units. To be supplied also in the set of drawings are the specifications of the auxiliary equipment such as pumps, the control systems and instruments, piping and the supporting structures for the plant.

APPLICATION TO THE DESIGN OF DISTILLATION UNIT

The Institute of Production Innovation and the Faculty of Engineering both of the University of Dar es salaam have developed a village level technology for small scale production of crystalline sugar. The by-products of this process i.e. molasses and bagasse are potential sources as the raw material and fuel respectively for production of valuable ethanol by distillation. Thus the approach for designing a distillation plant - as an extension to the existing sugar production plant, would be as follows;

Step 1. Establish design basis.

In this particular case production capacity of the distillation plant can be specified taking into account of the rate of material production from the existing sugar production process, and the operating schedule of the integrated plants. Further, product of reasonably high

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Table 1. Design basis of distillation plant [2]

Molasses usage	960 kg/day*
Ethanol production	50 kg/day
Ethanol in distillate	80% wt/wt
Ethanol in feed	6% wt/wt
Ethanol Yield	70% wt/wt
Operating days	105 days/year
Operating hours	8 hours/day

* 960 kg/day of fermented molasses is equivalent to 240 kg/day raw molasses.

concentration of ethanol should be specified so as to cater for wider range of possible uses of ethanol in Tanzania. This may include using ethanol as a reagent in the laboratories in schools, fuel for lighting homes, fuel in agricultural machinery, and as a sterilising agent in hospitals.

For the sake of this paper, Table 1 shows the design basis.

Step 2. Suggest flowsheets

In any general discussion of distillation, a distinction can be made between batch and continuous operation.

In batch wise operation the amount to be distilled is charged into the still and vaporised through the column. Part of the condensate is returned as reflux liquid and the remainder is collected as product. Batch operation can be broken further into;

- variable reflux operation, which gives product of constant composition, or
- constant reflux operation, which gives product of differing concentration.

In continuous distillation feed is continuously charged into the column, simultaneously withdrawing top and bottom products at constant compositions. The condition of feed is important in deciding the operating reflux ratio. consequently influencing the size of column and heat supply in the

reboiler and condenser capacity. For simple operations, as is the situation in this case study we may assume feed enters the column as either

- cold feed at its room temperature, or
- preheated to about its boiling point .

The distillation process is a heavy energy consumer. In the case of continuous operation (Fig 2a), it is possible to incorporate heat recovery options so as to achieve the required separation with reduced energy demand. Utility heat for preheating the feed to about its boiling point may be reduced by considering matching process streams according to the following options;

- Matching the feed with vapour and utility heat (Fig 2b)
- Matching the feed with effluent and utility heat (Fig 2c)
- Matching the feed with both vapour and effluent (Fig 2d)

In this case the heat exchanger network (HEN) may be derived according to the Pinch method developed by Linnhoff and Hindmarsh [5].

To summarise, alternative flowsheets of the distillation in this case study may be generated according to the “tree” shown in Fig 3.

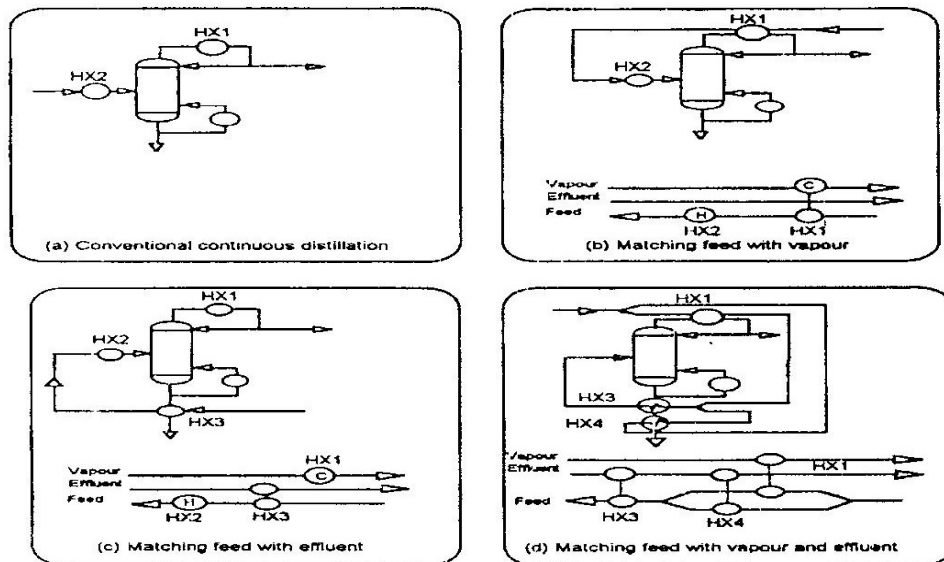


Fig. 2. Heat recovery options for single continuous column

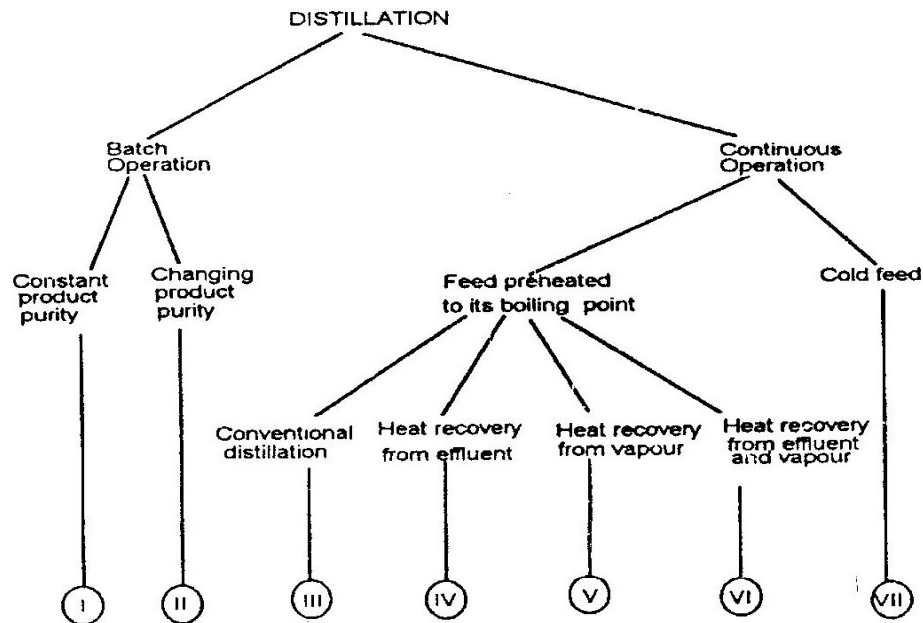


Fig. 3: "Tree" structure for generating distillation plant alternative flowsheets

Screening the alternatives

At this stage already some of the alternatives shown in Fig 3 may be discarded on the following grounds;

- Calculation of mass and energy balance suggests that the sugar process will produce more than enough bagasse which can be used by the distillation process [2] . This justifies dropping those alternatives which employs heat recovery options, i.e. alternatives IV, V and VI. Further, due to increased number of process streams, these alternatives would require rather complicated control systems and instrumentation.
- Cold feed to the column (alternative VII), would condense some of the rising vapour to provide sufficient heat to bring the feed to its boiling point. Thus there will be an increase in reflux flow below the feed plate and hence a larger diameter for the stripping

section of the column, and an increased heat consumption in the reboiler. On the other hand preheating the feed to its boiling point (alternative III) would lower the energy demand in the reboiler at the expense of capital cost of the preheater.

A detail cost analysis would be necessary to compare alternatives III and VII. However as a rule of thumb the former is assumed to a better alternative than the latter provided that the effluent need not be cooled before disposal.

- With reference to batch operations, operating a column to give constant product purity (Alternative I) requires adjusting the reflux continuously throughout the batch cycle, and stopping the fractionation when the reflux has climbed to some value considered to be uneconomical. An alternative policy for operating a column is with fixed reflux ratio. This gives product of high purity at the beginning of the fractionation and diminishing quality with time (Alternative II). In this case fractionation is stopped when the average composition of top product in the receiver has reached the product specification. Constant reflux operation is obviously the simplest and easiest to implement. This operating policy does not require installation of an automatic reflux control equipment and/or operator's attention to maintain constant product quality. However, to implement such a policy, a lower product composition (below the azeotropic composition) must be specified.

To summarise the foregoing discussion the alternatives which should be carried forward for further analysis are;

- alternative II, i.e. batch plant with changing product purity, and
- alternative III, i.e. continuous conventional distillation plant layout.

Step 3. Suggest unit types

At this step suggestions for alternative design of major process units in the distillation plant are made. Specifically these are the column and the heat exchangers.

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Column internals.

The alternatives for the design of the column are to use either packing or trays for the column internals. Example of commercially available trays include sieve tray, bubble caps, and valve trays. Likewise packing of different types are available and they include Raschig rings, Pall rings, and Saddle packing. In general Raschig rings are recommended for the design of small scale distillation plant[6]. However, if it is unavoidable to use trays, then of all the available tray designs, the sieve tray may be considered ideal for the distillation plant. Sieve trays are simple in design and can be produced locally in a small workshop.

Heat exchangers

Different heat exchangers designs are known, but consideration of fabrication methods in the developing countries and maintenance requirements may be used to short list the alternatives to only three. These are;

- Coil in tank heat exchanger (CTK)
- Double pipe heat exchanger (DPP)
- Shell and tube (STB)

Fig 4 shows the distillation plant flowsheets with possible heat exchanger types for each heat exchanger task. Unfortunately there is possibly no substitute for copper and stainless steel tubes in heat exchangers, however there can be a choice of the sizes of these materials.

Step 4. Evaluate unit types

On the basis of plant specifications as in Table 1 capacity of the column and heat duty of each heat exchanger in the two alternative flowsheets - batch and continuous column, are first worked out. This may employ the usual mass and energy balance calculations. Examples of results are as shown in Table 2 and Table 3 for batch and continuous distillation plant respectively.

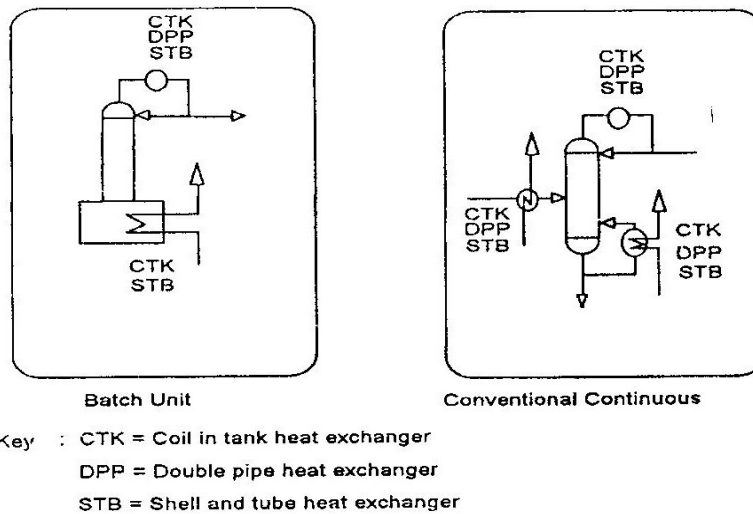


Fig.4; Distillation plant flowsheets with possible heat exchanger geometries

Table 2. Mass and energy requirement (batch distillation plant)

Charge	317 kg/batch
Number of batches	3 batches/day
Batch time	2hr: 40 min
Condenser load	28.3 kW
Reboiler load	86.32 kW

Table 3. Mass and energy requirements (continuous distillation plant)

Column feed rate	119 kg/hr
Condenser load	3.51 kW
Reboiler load	6.78 kW
Preheater load	10.35 kW

Two approaches would be possible to estimate the capital investment for each unit in the plant;

- i) using the power rule [7]
- ii) cost estimates on the basis of actual volume of materials of construction used.

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The second method is used in this case since with this method the effect of design alternatives on cost for the same performance of the unit can be realised.

Table 4 shows typical available materials of constructions and their costs in Tanzania[2].

Table 4: Assumed costs of materials of construction and utilities

Material	Cost Tsh/kg	Possible use
Copper tube	16400	Heat exchanger tubes
Brass plate	16400	Heat exchanger shell
MS plate	7600	Sieve trays, column shell
Ms pipe	7600	Column shell, heat exchangers
Raschig rings	7600	Column packings
Bagasse	0.24	Energy source
Molasses	0.42	Raw material
Water	0.33	Condenser cooling fluid
Ethanol	100	Final product

Density of steel = 7830 kg/m³

Density of brass = 8950 kg/m³

Bulk density of Raschig rings = 740 kg/m³

Process stream temperatures and heat transfer coefficients for heat exchangers of different types are assumed (typical values may be found in Perry and Green[8], Cculson and Richardson[9]) as shown in Table 5. Hence, capital investment for alternative designs of the column and heat exchangers (results from simulation) are tabulated (Table 6) and plotted (Fig 5) for the batch distillation plant. Similarly results for continuous distillation are as shown in Table 7 and plotted in Fig 6.

Table 5: Assumed stream temperatures and overall heat transfer coefficients in heat exchangers

	Tube side	Shell side	U _o (W/m ² .K)
Preheater	Hot air	Molasses	850
	600 - 450 °C	25 - 100 °C	
Condenser	Water	Vapours	1000
	25 - 50 °C	80 - 80 °C	
Reboiler	Hot air	Molasses	850
	600 - 450 °C	25 - 100 °C	

It should be emphasized that the type of unit which gives lowest capital cost does not necessarily mean the best design. For this matter, preferred

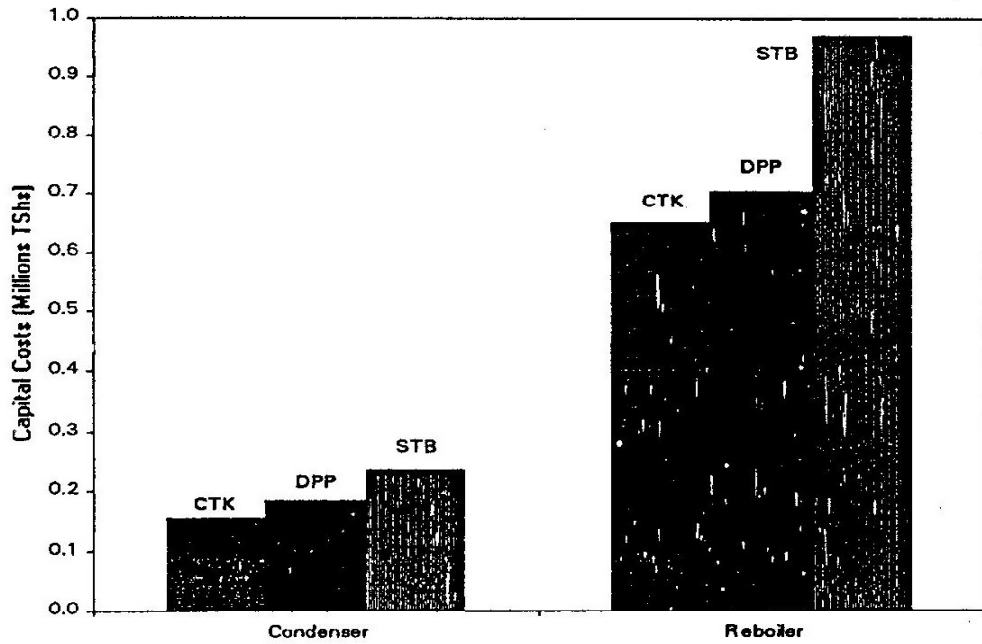


Fig. 5: Capital costs of alternative heat exchanger design - batch distillation

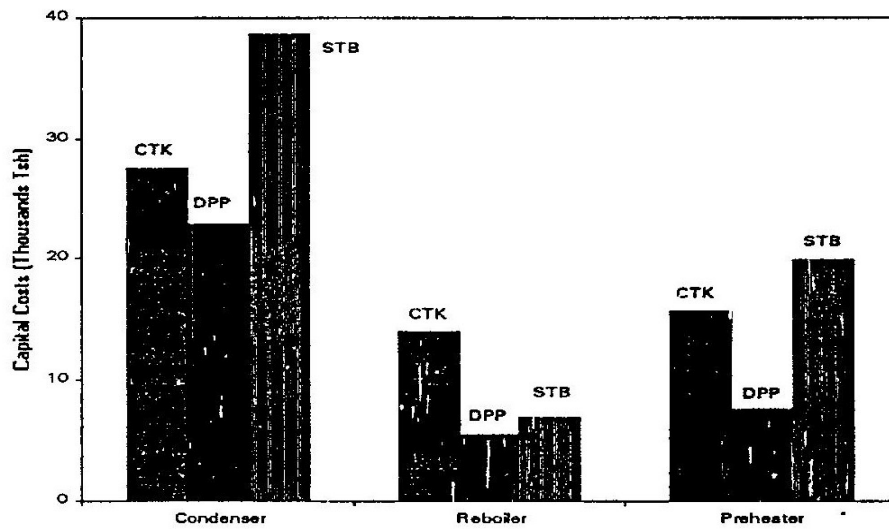


Fig. 6: Capital costs for alternative heat exchanger designs - continuous plant

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type of unit for each process task is also indicated in Table 6 and 7.

Table 6: Capital costs of units and flowsheet (batch distillation plant)

Unit name	Unit type	Area (m ²)	Capital cost (Tsh)	Minimum unit costs (Tsh)	Cost of preferred design. (Tsh)
Column+Still	Packed	n.a	582500	582600	582600
	Tray	n.a	808800		
Condenser	CTK	0.54	155800	155800	155800
	DPP	0.54	184700		
	STB	0.54	238000		
Reboiler	CTK	0.205	65100	70200	97000
	DPP	0.205	70200		
	STB	0.205	97000		
			Total cost	808700	835400

Table 7: Capital costs of units and flowsheets (continuous distillation plant)

Equipment name	Equipment type	Area (m ²)	Capital cost (Tsh)	Minimum unit costs (Tsh)	Cost of preferred design. (Tsh)
Column	Packed	n.a	194400	194400	194400
	Tray	n.a	345800		
Condenser	CTK	0.067	27600	22900	27600
	DPP	0.067	22900		
	STB	0.067	38700		
Reboiler	CTK	0.016	13900	5500	7000
	DPP	0.016	5500		
	STB	0.016	7000		
Preheater	CTK	0.022	15600	7600	19900
	DPP	0.022	7600		
	STB	0.022	19900		
			Total Cost	230400	249000

Thus with regard to column design, packings are preferred for column internals instead of trays. Much have been said on the advantages of packing over the trays and vice versa. However, in the context of designing for the developing countries the following advantages of packing over the trays appears to be most relevant;

- packings offer low capital cost at smaller column diameters,
- packings are preferred for foaming systems,

- more corrosion resistance with a range of materials of construction available.

For the sake of argument, where trays are used, the advantages they may offer is that trays are easier to clean than packings.

With regard to heat exchangers, a shell and tube heat exchangers would be preferred for the preheater and reboiler. This is based on the assumption that heat supply will be by flue gases as is the case with the existing sugar plant, hence the tubes will need regular cleaning (mechanical cleaning) due to severe fouling caused by the ash/smoke remains in the tube. For a condenser, a coil in tank design is preferred. In the condenser one may chose placing water in the tube side and therefore fouling will not be so critical. In addition, as opposed to the double pipe heat exchanger design, the shell and tube and coil in tank design offer compact design.

Step 5. Evaluate the flowsheets

At this step, in addition to capital costs calculated in step 4, the operating costs may be estimated based on the quantity of utilities (molasses, bagasse and water) requirements for each flowsheet. Also plant life may be assumed to be known which allows estimation of annualized cost associated with each flowsheet. Further, by assuming the cost of product ethanol is known, the percentage rate of return of investment (for simplicity depreciation is not accounted for) is worked out. Example of results are entered in Table 8, and plotted in Fig 7. For the underlying set of assumptions this particular case study shows that a continuous plant is economically more attractive than the batch plant.

Table 8: Ranking of the distillation plant alternatives on economic basis

Flowsheet	Capital cost (Tsh)	Energy cost (Tsh/yr)	Annualised cost (Tsh/yr)	Revenue (Tsh/yr)	% Return of Investment*	Ranking
Batch plant	835400	315900	483000	535500	26	2
Continuous	249000	76480	125900	535500	185	1

*cost of ethanol product = 100 Tsh/kg, plant life = 5 years.

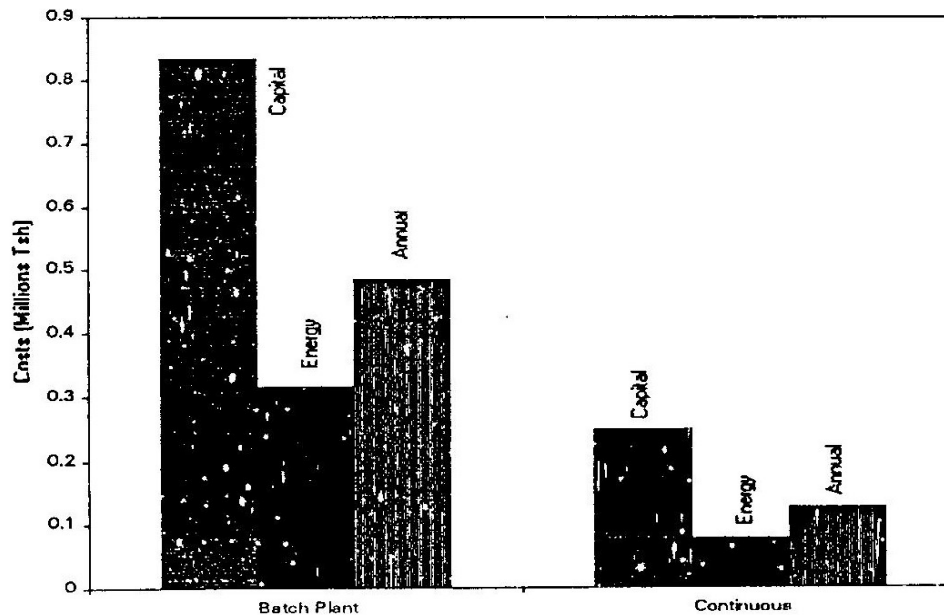


Fig. 7: Comparison of costs Batch vs Continuous distillation plants

Step 6. Identify appropriate flowsheet

At this step the most appropriate flowsheet is identified not only from the view point of economics of the process (step 5) but also from the appropriateness of the process to the total environment. Thus in the context of this case study comparison of batch versus continuous operation would be as follows;

Advantages of batch (disadvantages of continuous) plant;

- various feed compositions can be readily accommodated in a batch operation.
- batch column can separate a number of components into its pure components. To achieve the same separation with a continuous distillation system requires $n-1$ columns for a system of n components (Rose, 1985).
- batches must be replenished and sizes of fractions monitored, switched on and discarded, manpower requirement are high. This would offer employment for many people in the villages and may be considered as an advantage.

Advantages of continuous (disadvantages of batch) plant

- Continuous distillation is easier to control. In batch operation conditions change with time hence it will be difficult to control.
- Manpower costs are low as compared to batch plants.

To conclude this step the economic performance of each flowsheet (step 5) and the non quantifiable factors in (step 6) are considered in their totality, from which it is concluded that continuous process appears to be most appropriate for the prevailing set of assumptions.

Step 7. Mechanical design and control strategy.

Mechanical design considerations/Plant layout

It was concluded in the foregoing sections that a continuous distillation plant with packed column is the most appropriate design for the set of assumptions made on the conditions on site. However the assumptions made in reaching this conclusion may be different from the actual conditions on site, or, some of the decisions made in the design steps are subjective in the sense that another designer may have a different opinion.

For that matter, preparation of the detailed engineering drawings will not be presented in this paper, instead it is left to an interested reader / designer as a recommendation for future work. However, it need not be over emphasized that mechanical design and control of the distillation plant should be made in a way to match the level of rural labour that will operate it. Also, the design should be simple so as to facilitate its production by the small machine shops.

CONCLUSION

Design guidelines that will enable the designer to develop an optimum flowsheet for a small scale continuous, steady state production process has been proposed. The proposed guidelines are considered appropriate for the prevailing economic, social and/or cultural level typical of a developing country. However, some of the variables that enter the design process (refer Fig 1) may not fall into a specific step. It is easier to examine these cases to highlight complex issues which need to be resolved. In line with this, the design approach of a distillation plant to match with the village level technology for the production of sugar has been used to dem-

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onstrate several decisions that may enter the various stages of the design process.

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