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Effect of the Variation of the Level of Lactose Conversion in an Immobilized Lactase Reactor upon Operating Costs for the Production of Baker's Yeast from Hydrolyzed Cottage Cheese Whey Permeate

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ABSTRACT

Operating costs for the production of Baker's yeast from hydrolyzed permeate from the ultrafiltration of cottage cheese whey were calculated as a function of the level of lactose conversion in the immobilized lactase reactor. These costs were calculated for the case of 90% conversion of lactose in the reactor and compared to those that result when running the reactor at lower conversions with recycle of unreacted lactose. Total operating costs were estimated by combining individual operating costs for the immobilized enzyme reactor, costs associated with processing a lactose recycle stream, and energy costs associated with cooling the reactor feed stream and sterilizing the hydrolysate stream. It was determined that operating costs are minimized at about 9.9 ¢/lb. of lactose when the reactor is run at approx. 72% conversion. This represents a savings of 2.4 ¢/lb. of lactose over the case of a once-through 90% conversion of lactose in the reactor.

Index Entries: Economic analysis; immobilized enzyme reactor; baker's yeast production; whey processing; reactor operating strategy.

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INTRODUCTION

Whey is the liquid remaining after the coagulation of milk in the cheese making process. It is an aqueous solution containing 6–7% total solids. The major constituents of these solids are lactose (65%), protein (14%), and ash (14%). In order to meet the challenge of whey disposal (some 40 billion lbs. annually), the dairy industry is currently seeking more outlets for which the lactose rich streams that result from processing could be used as feedstock.

One method that has been given attention in recent years is to process the whey via ultrafiltration. In this process, whey is pumped under pressure across the surface of a semipermeable membrane that selectively passes or rejects molecules. The relatively large protein molecules are retained by the membrane, whereas both the small solute and the solvent molecules pass through the membrane, thereby fractionating the whey into a protein rich retentate and a permeate solution containing both lactose and salts. This process yields a whey protein concentrate (WPC) with excellent functional properties as the primary end product. The permeate resulting from the ultrafiltration is an aqueous solution containing about 5% lactose, 0.5% ash, and small quantities of other low molecular weight species. This permeate can be used in the production of refined lactose or animal feeds or as a feedstock for a number of fermentation processes.

In the early 1980s, Corning Glassworks entered into a joint venture with the Kroger Company to combine Corning's immobilized enzyme technology for hydrolyzing the lactose in whey permeate with Kroger's fermentation technology for producing Baker's yeast. The technique consists of processing cottage cheese whey from Kroger's dairy operations by ultrafiltration to produce WPC and a lactose rich permeate. The permeate can be hydrolyzed to produce a solution that can be used as a fermentation medium for production of Baker's yeast. Commercial viability of this process appears to be tied to the economics that result from internal integration of a raw material source, whey disposal, and end product utilization. Because of the in-house utilization of the whey, it is difficult to predict how process economics will be affected by the performance of the hydrolyzed permeate as a fermentation feedstock.

In this paper, operating costs for the production of Baker's yeast from hydrolyzed cottage cheese whey permeate have been calculated as a function of the level of lactose conversion in the immobilized enzyme reactor. In experimental studies, beta-galactosidase (lactase) immobilized on a porous silica carrier has been used to effect the hydrolysis of lactose contained in permeate that results from the ultrafiltration of cottage cheese whey. Previous articles have addressed the description of this immobilized enzyme system in terms of integral reactor analyses. Modeling of the packed-bed-immobilized enzyme reactor has been carried out starting with the description of a pore in a single catalyst particle and then expanded into an integral reactor description. The integral reactor model has been utilized to study the behavior of the immobilized enzyme under steady-state conditions (*1*), and under conditions of substantial decay in enzymatic activity (*2*), and to predict the lifetime of the immobilized enzyme catalyst under conditions that one might utilize in industrial applications (*3*). These previously cited models were used to determine the level of lactose conversion that should be achieved in the immobilized enzyme reactor to minimize overall operating costs for the production of Baker's yeast. Analysis of the Baker's yeast process is treated in three parts: process specification, calculation of costs, and discussion of results.

PROCESS SPECIFICATION

Engineers at Kroger Foods and Corning Glassworks have proposed that hydrolyzed permeate be used to produce Baker's yeast via a fermentation process. This section describes the simplified process flowsheet for the production of Baker's yeast from the hydrolyzed permeate. The process feed is assumed to be permeate from the ultrafiltration of cottage cheese whey. The throughput of the process is assumed to be 30,120 gal of permeate/d (12,000 lbs./d of equivalent lactose). The process is assumed to be in operation 95% of the time and to consume 90% of the lactose in the fresh feed. Figure 1 is a schematic diagram of the simplified process flowsheet. Table 1 contains a description of the contents of each of the process streams shown in Fig. 1.

Permeate from the ultrafiltration of acid whey (4.8% lactose, .6% ash) is assumed to be provided by the cottage cheese manufacturer at a rate of 1,255 gal/h (equivalent to processing 526 lbs. lactose/h). This corresponds to processing 12,000 lbs. lactose/d, assuming that the system is in operation 95% of the time. The process feed is mixed with the lactose recycle stream and heated (or cooled) to the reaction temperature before introduction into the immobilized lactase reactor. The lactose recycle stream is assumed to contain the same concentrations of lactose and ash as are present in the process feed. After hydrolysis of the lactose to the specified level of conversion, the hydrolysate stream is sterilized by heating to 90°C for 30 s and then cooled to 30°C. The stream is then fed to the yeast fermentation process of the type described by Edwards (*4*). The hydrolyzed permeate is inoculated with a culture of *Saccharomyces cerevisiae*. Acceptable fermentation conditions include temperatures in the range of 28–32°C and a pH in the range of 4.60–5.20. Hydrolyzed permeate and yeast inoculum are continuously introduced into a yeast growth zone in a fermenter. A portion of the fermentation liquor equal to the flow rate of the incoming inoculum is continuously harvested and transferred into aerated, temperature controlled storage tanks. The residence time of the inoculated liquid in the fermenter is such that a substantial quantity of

Table 1
Description of the Streams Shown in the Process Flowsheet

Stream no.	Contents	Comments
1	Lactose, ash, water	Process inlet stream; Permeate from the ultrafiltration of cottage cheese whey
2	Lactose, ash, water	Combined recycle and process inlet stream
3	Lactose, ash, water	Reactor feed stream
4	Lactose, galactose, glucose, ash, water	Hydrolyzed permeate
5	Lactose, galactose, glucose, ash, water	Stream 4 sterilized for production of yeast
6	Diammonium and monoammonium phosphate, ammonium sulfate, biotin, thiamine	Nutrients for production of yeast
7	Lactose, yeast, ash, yeast metabolites	Yeast rich stream
8	Lactose, yeast, ash, yeast metabolites	Retentate from ultrafiltration; Product stream
9	Lactose, ash, water	Permeate from ultrafiltration; Lactose recycle stream
10	Water	Permeate from reverse osmosis
11	Lactose, ash, water	Retentate from reverse osmosis
12	Ash	Ion exchange process; Retained by resin until regeneration
13	Lactose, ash, water	Treated recycle stream; Has the same composition with respect to lactose, ash, and water as the process inlet stream

yeast is produced before removal (typical residence times are less than 15 min). Substantially complete carbohydrate utilization is accomplished by allowing the fermentation liquor to remain in the storage tanks for four h.

The liquor containing the yeast and yeast metabolites is then ultrafiltered. The retentate is the product stream for the overall process. This stream contains yeast, yeast metabolites, water, ash, and an amount of lactose that is equal to 10% of the lactose entering the process (52.6 lbs. lactose/h). The permeate from the ultrafiltration of the yeast rich stream constitutes the lactose recycle stream. The lactose recycle stream must be concentrated by reverse osmosis to remove water and subjected to partial deashing via ion exchange to create a stream with the same lactose and ash levels as the fresh feed to the process. The recycle and fresh feed streams are combined to constitute the reactor feed stream.

The operating costs for the process shown in Fig. 1 are a function of the extent of conversion of lactose achieved in the immobilized enzyme

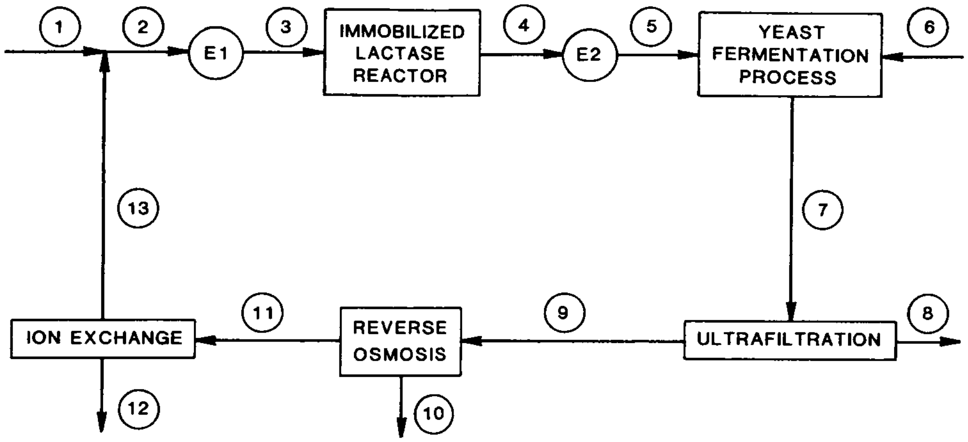

Fig. 1. Process flowsheet for the production of Baker's yeast from the permeate of ultrafiltered cottage cheese whey (contents of numbered streams are listed in Table 1).

reactor. These costs were computed assuming 50, 60, 65, 70, 75, 80, and 90% conversion of lactose in the reactor. A detailed mass balance for one of these seven cases is given in the next section. The cost of operation of the reactor is approximated by calculating the cost of the catalyst that is necessary to maintain a given level of production of hydrolysate. This aspect of the analysis is discussed in a previous publication (3).

CALCULATION OF COSTS

This section summarizes the results of the material balance calculations for the production of yeast from the permeate resulting from the ultrafiltration of cottage cheese whey. The system is assumed to have a throughput of 30,120 gal/d of permeate (12,000 lbs. of lactose/d), while remaining in operation 95% of the time. This throughput corresponds to a flow rate of 526.3 lbs. lactose/h. While the conversion of lactose in the immobilized lactase reactor is varied from 50 to 90%, the recycle of unreacted lactose permits the overall process conversion of lactose to be maintained constant at 90% of the lactose entering in the feed stream. In order to permit recycle of part of the unconverted lactose, the product stream must be subjected to additional processing.

In the cases where lactose recycle is necessary, the product stream from the yeast fermentation process is ultrafiltered. Permeate is removed to an extent that reduces the amount of lactose in the retentate stream to 10% of the amount of lactose that enters the process in the fresh feed stream. The permeate stream that contains the remainder of the unreacted lactose is first treated by reverse osmosis to effect removal of water and then by ion exchange to effect removal of ionic constituents. By removing appropriate amounts of water and ash, the composition of the recycle stream can be adjusted so as to contain the same water-to-lactose-to-ash ratio as that present in the fresh feed to the process. Table 2 summarizes the material balance calculations for the system for a per pass lactose conversion in the immobilized lactase reactor of 70%. The flow rates listed in Table 2 are used to determine loads on the processing equipment. These loads are then used to determine the operating costs for the several steps in the process.

Processing costs include those associated with the immobilized lactase reactor (ILR), the ultrafiltration system (UF), the reverse osmosis system (RO), and the ion exchange system (IE), as well as those arising from system energy requirements. For each aspect of the process the operating costs were determined for conditions where the ILR was run at levels of 50, 60, 70, 80, and 90% conversion of lactose.

The operating cost for the ILR was assumed to arise only from the costs associated with the catalyst in the reactor. The analysis presented in (*3*) indicates that the catalyst bed would last for 90 d before requiring replacement. The amount of catalyst needed to perform the required extent of lactose conversion was determined by multiplying the appropriate reactor space time by the volumetric flow rate of the reactor feed. Catalyst costs were determined using a bulk selling price of $32/lb. of catalyst (*5*). The cost per pound of lactose was then determined by dividing the cost of the catalyst bed by the amount of lactose processed in 90 d of operation. Table 3 contains the data utilized to determine the cost of the catalyst per pound of lactose. The tabulated results indicate that the reactor operating costs increase by a factor of six as the level of lactose conversion per pass through the reactor increases from 50 to 90%. In addition, the capital costs associated with the ILR would increase because of the larger reactor volume that would be needed to contain the immobilized enzyme.

The operating cost of the ultrafiltration system was determined by multiplying the load on the apparatus by a per unit load cost factor. The cost of ultrafiltration was determined by multiplying the permeate flow rate calculated in the mass balance by the cost per pound of permeate. The cost for concentrating the retentate from 6.3 to 10.8% total solids by ultrafiltration is approximately 0.044 ¢/lb. of permeate (*6*). The cost of ultrafiltration per pound of lactose was determined by dividing the hourly cost by the input rate of lactose to the process (526.3 lbs. lactose/h). Table 4 summarizes the estimated operating costs for the UF system. The cost of operating the ultrafiltration system decreases by a factor of about 2.6 as the conversion in the immobilized lactase reactor is increased from 50 to 80%. The ultrafiltration system was not needed in the case of 90% conversion because no lactose recycle stream would then be required. Usually, ultrafiltration operations yield retentates that contain less than 12% total solids. Examination of the operating conditions re-

Table 2
Mass Balance for 70% Hydrolysis through the Immobilized Lactase Reactor

Stream No.	Lactose, lb/h	Glucose, lb/h	Galactose, lb/h	Water, lb/h	Ash, lb/h	Yeast, lb/h	Metabolites, lb/h	Nutrients, lb/h	Total, lb/h	Temp., °C	% Total solids
1	526	—	—	10,372	66	—	—	—	10,964	49	5.40
2	676	—	—	13,336	85	—	—	—	14,097	Tm[a]	5.40
3	676	—	—	13,336	85	—	—	—	14,097	Tr[b]	5.40
4	203	249	249	13,311	85	—	—	—	14,097	Tr	5.58
5	203	249	249	13,311	85	—	—	—	14,097	30	5.58
6	—	—	—	—	—	—	—	21	21	30	100.00
7	203	—	—	13,311	85	424	95	—	14,118	30	5.72
8	53	—	—	3,452	22	424	95	—	4,045	30	14.70
9	150	—	—	9,860	63	—	—	—	10,073	30	2.11
10	—	—	—	6,896	—	—	—	—	6,896	30	—
11	150	—	—	2,964	63	—	—	—	3,177	30	6.69
12	—	—	—	—	44	-	—	—	44	30	100.00
13	150	—	—	2,964	19	—	.	—	3,133	30	5.40

[a]Tm is the mixing temperature of the fresh feed and the recycle stream.
[b]Tr is the required reaction temperature.

Table 3
Catalyst Costs as a Function of the Level of Lactose Conversion in the Immobilized Lactase Reactor

% Conversion of lactose in the ILR	Feed rate, ft^3/sec	Reactor space time, sec	Reactor volume, ft^3	Catalyst cost, ¢/lb lactose
50	.083	110	9.2	1.25
60	.073	175	12.8	1.75
70	.063	277	17.4	2.38
80	.055	515	28.4	3.87
90	.046	1200	55.6	7.58

veals that the total solids content of the retentate is too great for normal ultrafiltration operations for the cases of 50 and 60% conversion. The operating conditions for the case of 70% conversion are not too far out of line and the conditions at 80% conversion are well within the normal range of ultrafiltration operations.

The operating cost of the reverse osmosis system was determined by multiplying the required rate of water removal by the cost of removal per unit weight. The costs of reverse osmosis operations per hour were determined by multiplying the water removal rate by the cost of 0.0528 ¢/lb of water removed (*6*). The cost per pound of lactose was determined by dividing the operating cost per hour by the rate of lactose input to the process (526.3 lb/h). Table 5 summarizes the cost estimates for operation of the reverse osmosis system. The operating costs for the reverse osmosis system decrease by a factor of 1.6 as the level of conversion in the immobilized lactase reactor is increased from 50 to 80%. In all of the cases shown, the amount of total slides in the retentate is well within the normal operating conditions for reverse osmosis (usually below 10% total solids).

The operating costs for the ion exchange system were determined by multiplying the rate of ash removal per hour by the unit cost of removal.

Table 4
Operating Costs for the Ultrafiltration System as a Function of the Level of Lactose Conversion per Pass in the Immobilized Lactase Reactor

% Conversion	Permeate flow rate, lb/h	Total solids of retentate, W/W In	Out	Operating costs (¢/lb lactose)
50	17,099	5.63%	22.04%	1.44
60	13,290	5.67%	18.54%	1.12
70	10,072	5.72%	14.70%	.85
80	6,576	5.79%	10.48%	.55
90	—	—	—	.00

Table 5
Operating Costs for the Reverse Osmosis System as a Function of the Level of Lactose Conversion in the Immobilized Lactase Reactor

% Conversion in the ILR	Water removal rate, lb/h	Total solids in the retentate In	Total solids in the retentate Out	Operating costs, ¢/lb lactose
50	8,275	3.08%	5.96%	.83
60	7,758	2.60%	6.24%	.78
70	6,896	2.11%	6.69%	.69
80	5,173	1.63%	7.62%	.52
90	—	—	—	.00

The cost of removal is approximately 2.63 ¢/lb. of ash (*7*). The cost of ash removal per pound of lactose was determined by dividing the cost of ion exchange per hour by the input rate of lactose to the process (526.3 lb/h). Table 6 summarizes estimates for the operating costs of the ion exchange system. The cost of operating the ion exchange system decreases by a factor of about 1.6 as the level of conversion in the immobilized lactase reactor is increased from 50 to 80%.

Costs were also calculated for the two energy inputs shown in Fig. 1. The first energy input (depicted by E1) is required to adjust the temperature of the feed stream entering the reactor. This energy input is variable because the required reactor operating temperature increases as a function of time on stream to maintain a constant rate of hydrolysis. The two streams that comprise the reactor feed are at different temperatures. The process feed is assumed to be at 49°C (120°F) and the recycle stream is assumed to be at 30°C. When the required reactor temperature is below the mixing temperature of the streams comprising the reactor feed, the energy input at E1 is in the form of cooling. When the required reactor temperature above the mixing temperature, the energy input at E1 is in the form of heating. The total energy cost (heating and cooling) at E1 was

Table 6
Operating Costs for the Ion Exchange System as a Function of the Level of Hydrolysis in the Immobilized Lactase Reactor

% Conversion in the ILR	Ash removal rate, lb/h	Operating costs, ¢/lb lactose
50	53	.26
60	50	.25
70	44	.22
80	33	.16
90	—	.00

determined for a 90-d cycle (catalyst lifetime) and then divided by the associated input of lactose to the system to obtain an average energy cost per pound of lactose. The energy requirements were plotted as a function of the number of days of operation and graphically integrated to obtain the total amount of heating and cooling for the 90-d cycle. Figure 2 contains plots of the required heating and cooling loads on E1 as a function of time for the cases of 50, 70, and 90% conversion. The plots indicate that the majority of the energy requirements for the reactor feed are in the form of cooling. No heating is required at the 80 and 90% conversion levels because the process inlet temperature (49°C) causes the reactor feed temperature to always be above the prescribed value.

The energy input at E2 arises from the necessity for sterilization of the hydrolysate before introduction into the yeast producing fermenter. The hydrolysate stream is rapidly heated to 90°C and held at the temperature for 30 s and then rapidly cooled to 30°C. The hydrolysate stream is then introduced into the fermenter. These energy costs were also determined on the basis of a 90-d cycle. This approach is necessary because the input temperature (initial temperature of the hydrolysate) varies as a function of time-on-stream. The cooling costs can be determined from a knowledge of the flow rate of the hydrolysate stream as the initial and final temperatures of the stream remain constant throughout production.

The energy requirements for heating were plotted as a function of the number of days of operation and then graphically integrated to determine the total heating requirements for the 90-d cycle. Figure 3 contains the plots for the heating load on E2 as a function of the number of days in

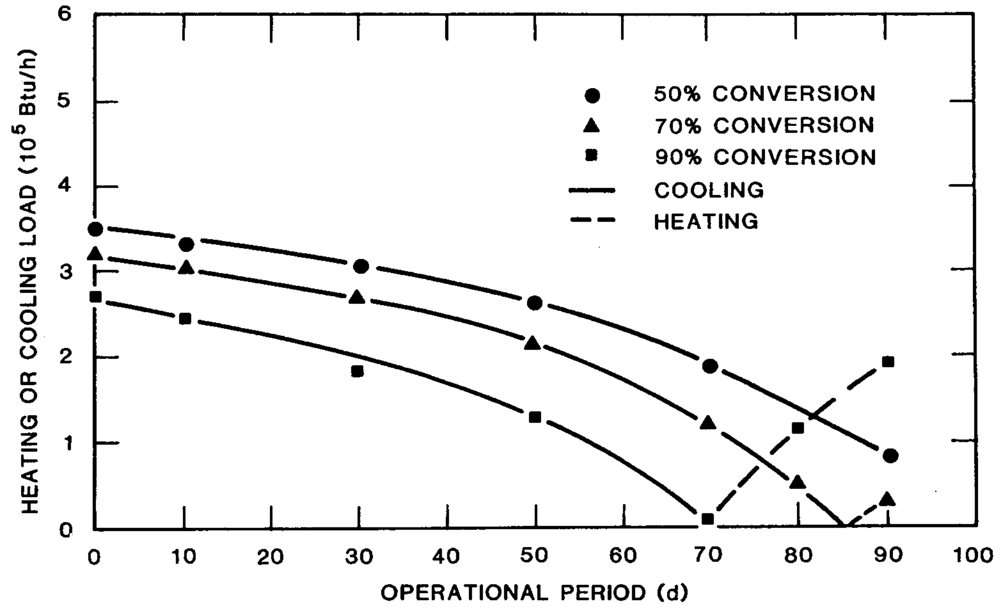

Fig. 2. Heating and cooling load on E1 as a function of days in operation.

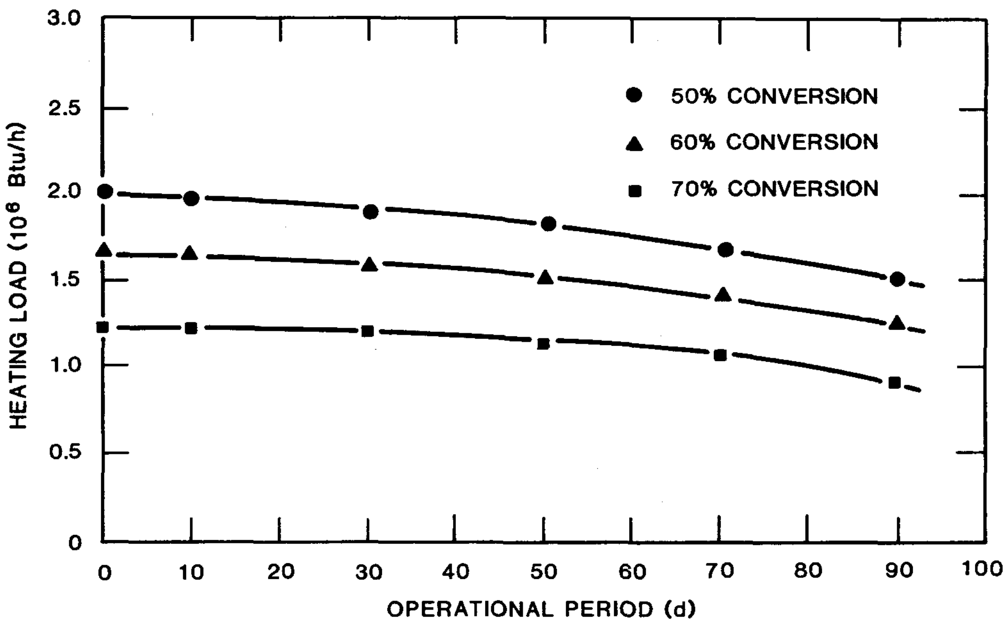

Fig. 3. Heating load in E2 for the sterlization of the hydrolysate stream.

operation for the cases of 50, 60, and 80% conversion. The profiles are substantially parallel because the temperature profile for the immobilized lactase reactor is essentially the same for each conversion level for the first 80 d in the cycle (*3*). The heating load is greater for the cases of lower conversion because the flow rates are larger.

Steam is used to provide the heat for the process. It is assumed that 250 psig steam was available at \$5.50/1000 lbs. of steam (*6*). The cost of using steam was assumed to be 0.67¢/1000 Btus of required heat. Cooling costs were determined assuming that an electric refrigeration system is used. With an electricity cost of 4.5 ¢/kw h (*6*), the cost of cooling the permeate is 1.32 ¢/1000 Btus of cooling (*8*). Table 7 summarizes the costs for the energy requirements for the reactor feed stream (E1), and for the sterilization of the hydrolysate stream (E2). The total energy cost decreases as the conversion increases from 50 to 90%. From these cost estimates, it is evident that sterilization of the hydrolysate stream is one of the most expensive operations in the process.

Table 8 summarizes the operating costs that were estimated in this section. The tabular entries indicate that costs associated with operation of the immobilized enzyme reactor and with process energy requirements constitute the major portion of the total processing costs. The total operating costs appear to be minimized when operating the reactor at approximately 70% conversion. These estimates do not include any costs which pertain to the yeast fermentation process, nor do they include any capital costs.

Table 7
Energy Costs for the Reactor Feed Stream (E1) and for the Sterilization of the Hydrolysate Stream (E2) as Functions of the Level of Conversion in the Immobilized Lactase Reactor

	E1		E2		
% Conversion in the ILR	Cooling cost, ¢/lb lactose	Heating cost, ¢/lb lactose	Cooling cost, ¢/lb lactose	Heating cost, ¢/lb lactose	Total energy cost, ¢/lb lactose
50	.32	.03	5.07	2.38	7.80
60	.41	.01	4.22	1.98	6.62
70	.49	.00	3.62	1.71	5.82
80	.57	.00	3.17	1.51	5.25
90	.62	.00	2.82	1.34	4.78

Table 8
Estimated Operating Costs for the Production of Baker's Yeast from the Permeate from the Ultrafiltration of Cottage Cheese Whey

Level of lactose conversion	Operating costs in cents per pound of lactose that is processed					
	ILR	UF	RO	IE	Energy	Total
50	1.25	1.44	.83	.26	7.80	11.58
60	1.75	1.12	.78	.25	6.62	10.52
70	2.38	.85	.69	.22	5.82	9.96
80	3.87	.55	.53	.16	5.25	10.35
90	7.58	.00	.00	.00	4.78	12.36

DISCUSSION OF RESULTS

The estimates in Table 8 indicate that the two major cost elements are those associated with operation of the immobilized enzyme reactor and with the energy requirements of the process. The cost of operating the reactor is 6.33 ¢/lb. of lactose greater at 90% conversion than at 50% conversion. This difference in the cost of reactor operation occurs as a consequence of the product inhibition characteristic of the enzymatic hydrolysis reaction (*1–3*). The energy costs are greater for the lower levels of conversion because the heating and cooling requirements are larger in these cases.

The tabular entries indicate the economic tradeoff established between the additional enzyme cost required at higher levels of conversion and the cost of processing a lactose recycle stream. For example, at 90% conversion the energy costs are relatively low and the recycle of lactose is not necessary. The cost of operating the reactor is extremely high, as is the overall cost of operation for the process. At 70% conversion the energy costs are greater than those at 90%, and a lactose recycle stream must be incorporated. In spite of these two expenses, the overall operating cost at 70% conversion is lower than at 90% conversion. The savings realized by operation of the reactor at 70% conversion are greater than the costs associated with the increased energy requirements and the necessity for processing a lactose recycle stream. Figure 4 is a plot of the operating costs as a function of the fractional conversion of lactose in the immobilized lactase reactor. From the cost profile it appears that the optimum level of lactose conversion is approximately 72%.

The total costs for the process must also include the capital expenditures necessary to build the system. Inspection of Table 8 reveals that additional capital costs of about 2.4 ¢/lb. of lactose can be tolerated for the 72% conversion system before the total cost of production approaches that of the 90% conversion system. This cost increment corresponds to

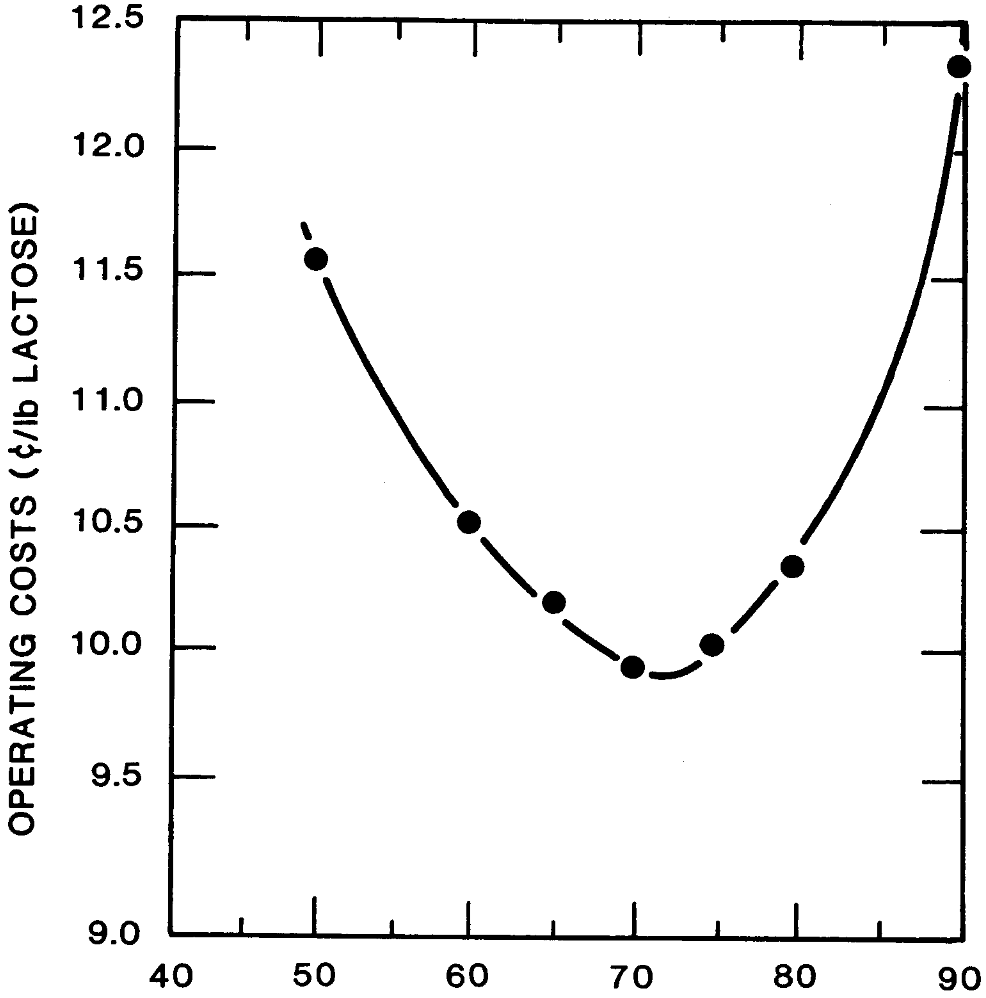

Fig. 4. Operating costs versus percent conversion of lactose in the immobilized lactase reactor.

an allowable extra capital cost of approximately $105,120/yr for a system in which the reactor operates at 72% conversion vs a system that contains a reactor that operates at 90% conversion. The total capital costs can be estimated by assuming a rate of depreciation of 10%/yr over a 10-yr period. This translates into an extra allowance of over 1 million dollars in capital costs for a reactor operating at 72% conversion rather than 90% conversion. The operating and capital costs for the yeast fermentation process were assumed to be constant in each case and were therefore not included in this analysis. Table 9 lists approximate capital costs for the RO, UF, and IE systems needed for this process. Subtracting the approximated capital costs from the allowable extra capital costs yields a surplus of about $500,000. The capital cost of the immobilized lactase reactor sys-

Table 9
Approximate Capital Costs for the Reverse Osmosis, Ultrafiltration, and Ion Exchange Systems

Component	Capital Cost, $[a]	Source
RO	270,000	Ref. *6*
UF	100,000	Ref. *8*
IE	130,000	Ref. *7*
Total	500,000	

[a]All numbers are based on the case of 72% conversion.

tem is greater for higher conversions because of the extra volume required to contain the catalyst. The lower cost of the reactor tends to make the surplus even more favorable for the cases in which the lower conversion levels are used.

There appears to be a significant incentive for operation of the immobilized lactase reactor at 72% rather than 90% conversion. However, there are several possible disadvantages that may affect the economic analysis of the system. Items of major concern include the assumption of no glucose, galactose, or yeast metabolites in the lactose recycle stream, the assumption that cost of operating the yeast fermentation system is equivalent for each case and the possible control problems that may arise as a consequence of the more complex nature of the system. Precluding the possibility of the presence of glucose, galactose, and yeast metabolites in the lactose recycle stream favors the economic analysis of the process proposed in this section. If galactose is present in the recycle stream, the reaction rate will slow down as a result of the effects of competitive product inhibition. Consequently, the reactor operating costs will increase. The presence of glucose will not affect the reaction rate but may promote the growth of undesirable microorganisms within the system. The effect of the yeast metabolites (mostly large hydrocarbons) on the reaction rate is unknown. At low metabolite concentrations, however, it is not expected that the reaction rate will be diminished to any significant extent.

Assuming that the operating costs for the yeast fermentation step are constant gives a slight economic advantage to the cases where the lactose recycle stream is required. In each case the amount of yeast that is produced is equivalent; however, the size of the flow streams increases as the level of lactose conversion decreases. Processing of the increased quantities of material would involve higher pumping costs and require larger fermentation vessels. The additional pumping and capital costs are not expected to make a major difference in the overall cost estimates. The presence of the lactose recycle stream necessitates the incorporation of three additional unit operations (ultrafiltration, reverse osmosis, and ion exchange) into the process. This process is not complex in comparison to others in the petroleum industry (for example); however, the presence of the additional equipment will complicate both operating and control procedures.

The results obtained from the simplified analysis of the yeast manufacturing process suggest that a more in-depth study of process operating conditions is needed in order to ascertain whether or not significant cost benefits can be realized by varying the level of conversion of lactose in the immobilized lactase reactor. Whether or not the proposed changes lead to an increase in the productivity of this particular system is not as important as stressing the general engineering concept of optimizing operating strategies to enhance the performance of systems in the area of biotechnology. This type of analysis could be particularly important when dealing with some of the advanced bioreactor concepts where use of a clever reactor operating strategy could be the deciding factor in designing a successful system.

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