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Technoeconomics of Butanol Extractive Fermentation in a Multiphase Fluidized Bed Bioreactor

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ABSTRACT

The design, performance, and projected economics of a new conceptual process for the *in situ* solvent extraction of butanol within a multiphase fluidized bed bioreactor are discussed. The process appears to have the potential to increase the effective concentration of product over tenfold while maintaining actual concentrations in contact with the organism at 10 g/L, i.e., below the threshold of inhibition. If this performance can be realized in commercial operation, the selling price to yield a 30% pretax return on investment could be reduced from over $1.25 per pound of butanol for the present process to $0.59 per pound for the new process.

Index Entries: Butanol; fermentation; economics; acetone; fluidized bed.

INTRODUCTION

Since the Mideast oil crisis of 1973, many people in government and industry have been concerned about the strategic implications of a loss of a major source of American crude oil and petrochemical feedstocks. Accordingly, over the past decade a large number of research programs have been directed toward reexploring the potential for abundant renewable raw materials, such as corn or cellulosics, as basic feedstocks for fuels and chemicals.

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HISTORICAL PERSPECTIVE

The Weizmann process for fermenting starch-containing grains to butanol and acetone was developed under the stimulus of the World War I demand for acetone for manufacturing "cordite," a double-based smokeless powder used for British naval guns (*1*). The process made a successful transition to civilian products but was finally surplanted in the 1950s by cheaper petrochemical processes.

Presently, butanol is produced primarily from propylene and synthesis gas by the oxo hydroformylation process over a rhodium catalyst, which favors the formation of *n*-butanol, the preferred isomer.

The US currently leads among world producers of *n*-butanol. Production has risen steadily at 5% annually since 1964 to a 1984 production level of 918 million pounds (*2*). European production is less and stagnant; Japanese production has declined to the point where Japan is now a net importer.

Interest in the Weizmann process revived as a result of the oil crisis of 1973, but the subsequent softening in crude oil prices in the early 1980s removed the competitive edge of renewable feedstocks compared with fossil materials. At the present time, the only ABE plant operating in the free world is at Germiston in the Union of South Africa at National Chemical Products, Ltd., a division of Sentrachem Ltd. The plant was started up in 1936. Its operation was described by Spivey (*3*).

The fermentation process has two serious problems: The yield is inherently poor, resulting from the production of byproducts, including high levels of carbon dioxide and hydrogen, to maintain the electronic balance of the metabolism of the organism; and the concentration of butanol product in the beer has been limited to about 1 wt% as a result of product feedback inhibition of the organism.

RAW MATERIAL DEMAND

Raw material economics has always been one of the most important parameters in the choice between synthetic and fermentation processes. Until 1938, the ABE process operated solely on corn using *Clostridium acetobutylicum* with a total solvents yield (i.e., butanol, acetone, and ethanol) of about 26.5% based on dry corn. After 1938, new organisms were developed that allowed the use of cheaper molasses.

Early in this study it became apparent that the stoichiometry of the fermentation was very important to the economics as a result of the large loss of glucose substrate to the production of byproducts. Even though the organism operates near its biological maximum, carbon yields are poor because of the large losses to carbon dioxide. The solvents are formed roughly in the ratio: 60% butanol, 30% acetone, and 10% ethanol, with large attendant release of carbon dioxide and hydrogen (*4–6*). Literature data

Table 1
Fermentation Stoichiometry, per 100 Moles of Butanol

Base Case

$$\underset{100.0}{\text{Glucose}} \rightarrow \underset{100.0}{\text{Butanol}} + \underset{200.0}{2CO_2} + \underset{100.0}{H_2O}$$

$$\underset{40.0}{\text{Glucose}} \rightarrow \underset{40.0}{\text{Acetone}} + \underset{120.0}{3CO_2} + \underset{159.8}{4H_2}$$

$$\underset{12.5}{\text{Glucose}} + \underset{12.5}{2H_2O} \rightarrow \underset{25.0}{2\text{Acetic Acid}} + \underset{25.0}{2CO_2} + \underset{50.0}{4H_2O}$$

$$\underset{8.3}{\text{Glucose}} \rightarrow \underset{16.6}{2\text{Ethanol}} + \underset{16.6}{2CO_2}$$

$$\underset{6.7}{\text{Glucose}} \rightarrow \underset{6.7}{\text{Butyric Acid}} + \underset{13.5}{2CO_2} + \underset{13.5}{2H_2O}$$

$$\begin{aligned} 167.5\ \text{Glucose} \rightarrow\ & 100.0\ \text{Butanol} + 40.0\ \text{Acetone} \\ & + 25.0\ \text{Acetic Acid} + 16.6\ \text{Ethanol} \\ & + 6.7\ \text{Butyric Acid} + 375.0\ CO_2 \\ & + 35.0\ H_2O + 223.4\ H_2 \end{aligned}$$

Goal Case

$$100.0\ \text{Glucose} \rightarrow 100\ \text{Butanol} + 200.0\ CO_2 + 100\ H_2O$$

for the usual Weizmann fermentation on *Clostridium acetobutylicum*, as shown in Table 1, were used to develop the process material balances used in the base case of this study.

Alternatively, Weimer has postulated that an organism could be developed that would produce only butanol according to the "goal" case of Table 1 (*7*). Theoretically, this could be accomplished, although probably with great difficulty, by blocking the acetyl CoA to acetaldehyde (to ethanol) pathway and the acetoacetyl CoA to acetone pathway. Small amounts of acetaldehyde might be added to the fermentation to inhibit the ethanol route, but since acetaldehyde is toxic in general, it might completely stop the fermentation.

PRODUCT INHIBITION

As with most fermentations, *Clostridium acetobutylicum* is inhibited by its own substrate and products. Previous studies have shown that butanol is the most toxic of the products (*8,9*). The fermentation is totally inhibited by butanol concentrations of 10–15 g/L (*1,10,11*). Thus, in normal batch operation, productivity is limited to about 0.25 g/L/h, i.e., 13 g/L butanol after 40 batch h plus 12 h of fermenter turnaround. Productivities in the range 0.16–0.58 g/L/h have been reported by various investigators (*12–17*). *This experience has suggested to many that the fermentation rate and product concentration could be raised to commercial levels by removing the butanol from the fermentation medium as fast as it forms.*

EXTRACTIVE FERMENTATION

Various methods for removing inhibitory products have been proposed. These include: extraction, adsorption, or membrane separation. The use of extraction by an organic solvent is the subject of this study.

As a preamble to extraction studies, a number of laboratories have tested the toxicity of various solvents toward the organism. In many cases it was found that good solvents for the product were also toxic.

In the early 1980s Leung of MIT (*18*) studied at small scale the extraction of butanol with corn oil to minimize inhibition.

Gill and Ratledge (*19*) suggested that the toxicity of hydrocarbons toward specific organisms might be related to the aqueous solubility of these compounds. They showed that toxicity was reduced by adding a nontoxic compound such as hexadecane to the organic phase.

Evans and Wang at Michigan (*20*) studied extractive fermentation at a 10 mL test tube scale with the aim of optimizing distribution coefficient vs toxicity by using mixtures of toxic solvents having good distribution coefficients with nontoxic hydrocarbons having low solubility in the aqueous fermentation medium. Oleyl alcohol was found to be a good solvent with low toxicity.

Japanese investigators also demonstrated the use of oleyl alcohol to extract butanol in batch experiments (*13*).

Blanch and coworkers at the University of California at Berkeley also showed in small-scale tests using oleyl alcohol as a solvent that the effect of butanol inhibition could be reduced and volumetric productivity increased by removing butanol in either batch (*14*) or fed-batch culture (*15*). The feasibility of using continuous processing in extractive fermentation was also demonstrated at bench scale in experiments in which the fermenter broth was continuously recycled to an external extraction column (*16*). In the fed-batch experiments using oleyl alcohol as a solvent, they were able to increase productivity to 1.5 g/L/h. However, in these experiments productivity was also inhibited by the high concentrations of the sugar substrate required in batch culture to drive the fermentation to high butanol productivities. True continuous culture would circumvent this problem. Blanch's continuous runs were operated for over 55 h at double the productivity of batch or fed-batch culture.

The scale of the extractive fermentation program at Battelle Memorial Institute picked up where the preceding studies left off. As one approach to alleviating the product inhibition problem, Battelle recently completed the first 2-y phase of a study for the US Department of Energy's Energy Conversion and Utilization Technology (ECUT) program to demonstrate the technical feasibility of its Multi-Phase Fluidized Bed (MPFB) bioreactor, in which a solvent is added to the fermentative environment to extract the product from the aqueous phase as fast as it forms (*21*). The MPFB reactor was shown to have the potential for dramatically improving fermentations that are strongly product inhibited. If successful with the ABE system,

this approach could be applied to other product-inhibited systems, such as acetic acid via *Clostridium thermoaceticum*.

MPFB BIOREACTOR CONCEPT

The MPFB bioreactor is a patented concept that extends the gas-phase Multi-solid Fluidized Bed process, developed and commercialized by Battelle for solid fuel combustion application, to the production of chemicals by fermentation. The bioreactor concept, illustrated in Fig. 1, is an advanced circulating fluidized bed reactor in which a dispersed phase of fine droplets of solvent (designated as the entrained phase) are circulated through coarse particles of immobilized cells (designated as the dense bed) that are in a fluidized state.

The entrained solvent is circulated through the dense-bed biocatalyst to continuously remove and concentrate the fermentation product(s). The solvent in the entrained extract phase is regenerated and the product recovered in an external system. For volatile fermentation products like butanol, the external system would consist of a conventional distillation train. The recovered solvent is continuously recycled to the fermenter to recover additional product. A photograph of the pilot-scale MPFB bioreactor that was built by Battelle is shown in Fig. 2.

The primary advantage of the MPFB bioreactor concept is in increasing the *effective* concentration of product in a broth that would be otherwise restricted to unacceptably low actual concentrations as a result of product inhibition. This increase in apparent concentration and, hence, productivity, reduces investment in fermenters and the recovery system by reducing process volume at a given production level and effects a reduction in both fermentation and recovery costs.

In the previous ECUT-funded program (*21*), the technical feasibility of the MPFB bioreactor was investigated using the classical acetone/butanol fermentation process as the model system. This feasibility demonstration work included

1. Development of an immobilized-cell biocatalyst for butanol production;
2. Characterization of product extractants;
3. Design and construction of a pilot-scale MPFB bioreactor;
4. Measurement of axial and microscale mixing in a conventional, liquid fluidized bed of gel biocatalyst beads; and
5. Operational testing of the pilot MPFB bioreactor to characterize permissible operating ranges and mixing characteristics in the absence of an actual fermentation.

Thus, the critical elements of the MPFB system for butanol production have been developed and evaluated, but the actual integration of these elements to demonstrate practical feasibility of the MPFB bioreactor under

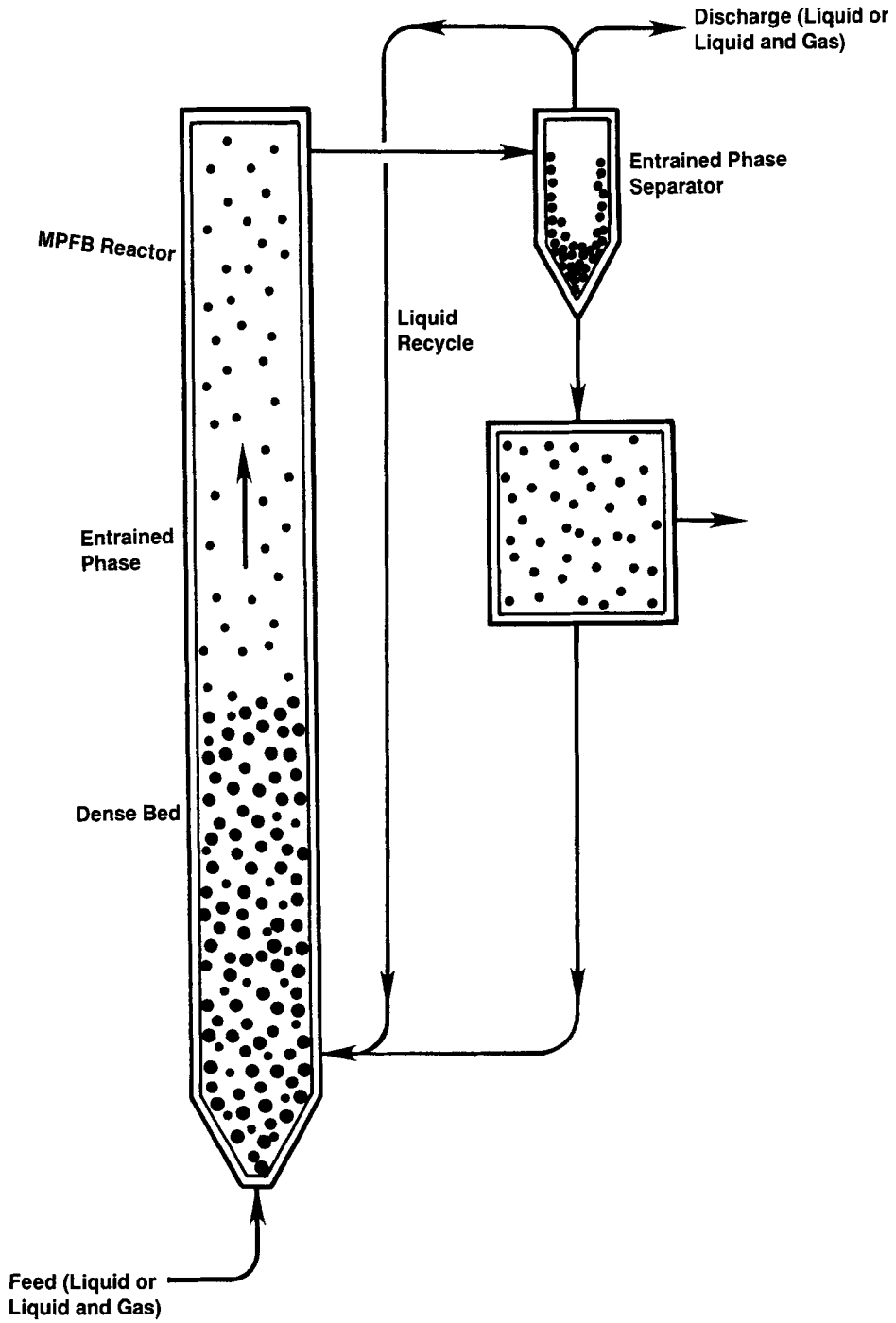

Fig. 1. Multiphase Fluidized Bed (MPFB) Concept

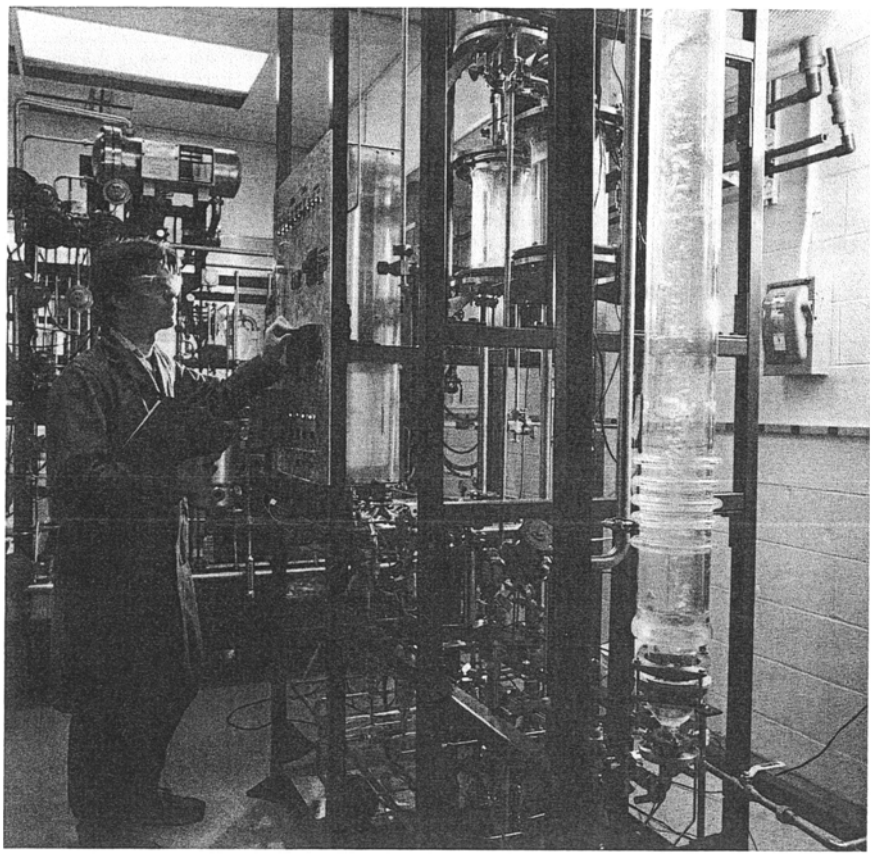

Fig. 2. Pilot-Scale MPFB Bioreactor

actual fermentation conditions has yet to be completed. Summarized in Table 2 are the key technical accomplishments, conclusions, and current state-of-the art for each of the six areas listed above.

A preliminary evaluation of the technical and economic status of the MPFB process was made with the objective of identifying strengths and weaknesses of the program and of assisting in developing an ongoing research strategy along economically relevant lines. These results are reported in the following sections.

ECONOMICS OF CONVENTIONAL PROCESS

Synthetic butanol currently sells for $0.34 per pound at a crude oil price of $15–20 per barrel. This price has been fairly stable since 1982.

Lenz and Moreira analyzed the economics of the conventional Weizmann process utilizing a molasses substrate and found that the process would have to operate at a loss because of the combined effect of high raw material costs and dilute concentrations (22).

Table 2
Summary of Current State of Development of MPFB Bioreactor

Task	Major Conclusions	Current State of Development
Biocatalyst development	Further optimization of the biocatalyst is needed for a commercial butanol process including developing a sporulation-deficient *Clostridium* strain and testing alternative cell immobilization methods and supports.	An active and sufficiently stable biocatalyst has been developed, and the preparation method scaled up to provide biocatalyst for demonstrating the feasibility of the MPFB bioreactor.
Characterization of product extractants	Extractants for fermentation products other than butanol are needed if the full potential of the MPFB bioreactor or other extractive fermentation are to be realized.	Two butanol extractants, oleyl alcohol and trioctyl phosphine oxide (TOPO), have been verified as having sufficiently high distribution coefficients for butanol and as being non-toxic for growth and product formation for *C. acetobutylicum* to be suitable for demonstration of the MPFB bioreactor.
Construction of pilot-scale MPFB bioreactor	Tests with bench-scale bioreactors demonstrated the need for a small pilot-scale MPFB unit as the minimum size for operational testing.	A fully operational pilot-scale demonstration unit has been constructed to test the MPFB bioreactor concept. Aseptic and anaerobic conditions can be maintained in the unit over a several day period.
Characterization of microscale and axial mixing	Axial mixing at low Reynolds numbers in a conventional non-aerated fluidized bed is more severe than predicted by the most widely accepted literature correlation. If a gas phase is present, axial mixing will be even more severe.	Back mixing has been documented as being an important scaleup issue for fluidized-bed bioreactors; more axial mixing tests (with and without a gas phase) are needed to establish a design basis for scaleup.
Operational testing of pilot-scale MPFB bioreactor	The MPFB bioreactor system can be operated at sufficiently high product extractant/aqueous flowrate ratios to achieve maximum benefits (i.e., cost savings) from the system. Extraction rates are much faster than the product formation rates to be encountered in an actual fermentation.	The basic principle of the MPFB bioreactor, that of *in situ* extraction to control product inhibition, has been demonstrated under nonfermentation conditions. Additional tests are need to demonstrate the system under actual fermentation conditions.
Preliminary technical and economic evaluation	There are no apparent technical barriers to scale up of the MPFB bioreactor system. The system can substantially reduce the cost of butanol and other similar fermentation products that are produced at low product concentrations in conventional batch processes once the fermentation processes have been fully optimized and scaleup.	An independent technical and preliminary economic evaluation of the MPFB bioreactor applied to the acetone/butanol fermentation has provided sufficient confidence to proceed with a staged technical development program.

This study reached the same conclusion. Butanol from a 167-million pound per year plant based on the conventional process and operating at 150 million annual pounds in 1987 would cost about $1.29 per pound for the fermentation operation alone (excluding the recovery operation). Thus, the cost would be considerable higher if recovery costs were included. This cost includes a 30% pretax return on total investment of $293 million.

The investment at a midpoint of construction of 1983 is comprised of: $230 million in direct process equipment; $33 million in allocated power, services and general facilities; and $30 million in working capital. Of the direct process investment, $200 million alone is for fermenters (16 million gallons) and their ancillaries priced installed at $12.50 per gross gallon. A contingency factor of 30% was applied to the investment estimate. Working capital is high because of the high value of product inventory and accounts receivables.

This estimate assumes a butanol concentration of 13 g/L at a batch time of 40 h and a turnaround of 12 h. Of the glucose converted, 5% is used for cell growth and 95% for products according to the stoichiometry of Table 1.

The product cost sheet is sumamrized as follows

Raw materials*	$0.40
Utilities	0.04
Labor related	0.03
Depreciation	0.14
Other capital related	0.05
Cost of manufacture	$0.66
SE, R&D, Adm.	0.16
Cost of sales	$0.82
Pretax earnings @30%	0.58
Byproduct Credit	(0.11)
Cost plus return price	$1.29

PROCESS SCENARIO FOR EXTRACTIVE FERMENTATION

The process model used in the economic study is shown on the flowsheet of Fig. 3. The extraction of the butanol product is effected by a single-stage contact in the back-mixed fermenter. Aqueous raffinate containing the cell debris and unextracted product and byproducts is sent to waste disposal. The extract is sent to a series of low boiler stills for recovery of butanol and acetone in purified form. It was assumed that ethanol, acetic acid, and butyric acid are produced in too small a quantity to justify recovery.

The higher boiling acids, acetic and butyric, are removed overhead from the solvent in the high boiler's still. Actually, it is assumed, rather than known, that the acids are, indeed, produced and that the pH of extraction is low enough for the acids to exist as acids rather than salts. In salt form they could not be extracted and would remain with the aqueous raffinate waste.

Production scale was sized to a 150 mm PPY butanol plant with a midpoint of construction in 1983 and operating in 1987.

EXTRACTION BASIC DATA

A distribution coefficient of 4.3 for butanol in an oleyl alcohol/water system has been determined experimentally (*21*). At a solvent to water in

*Sugar at $0.088/lb.

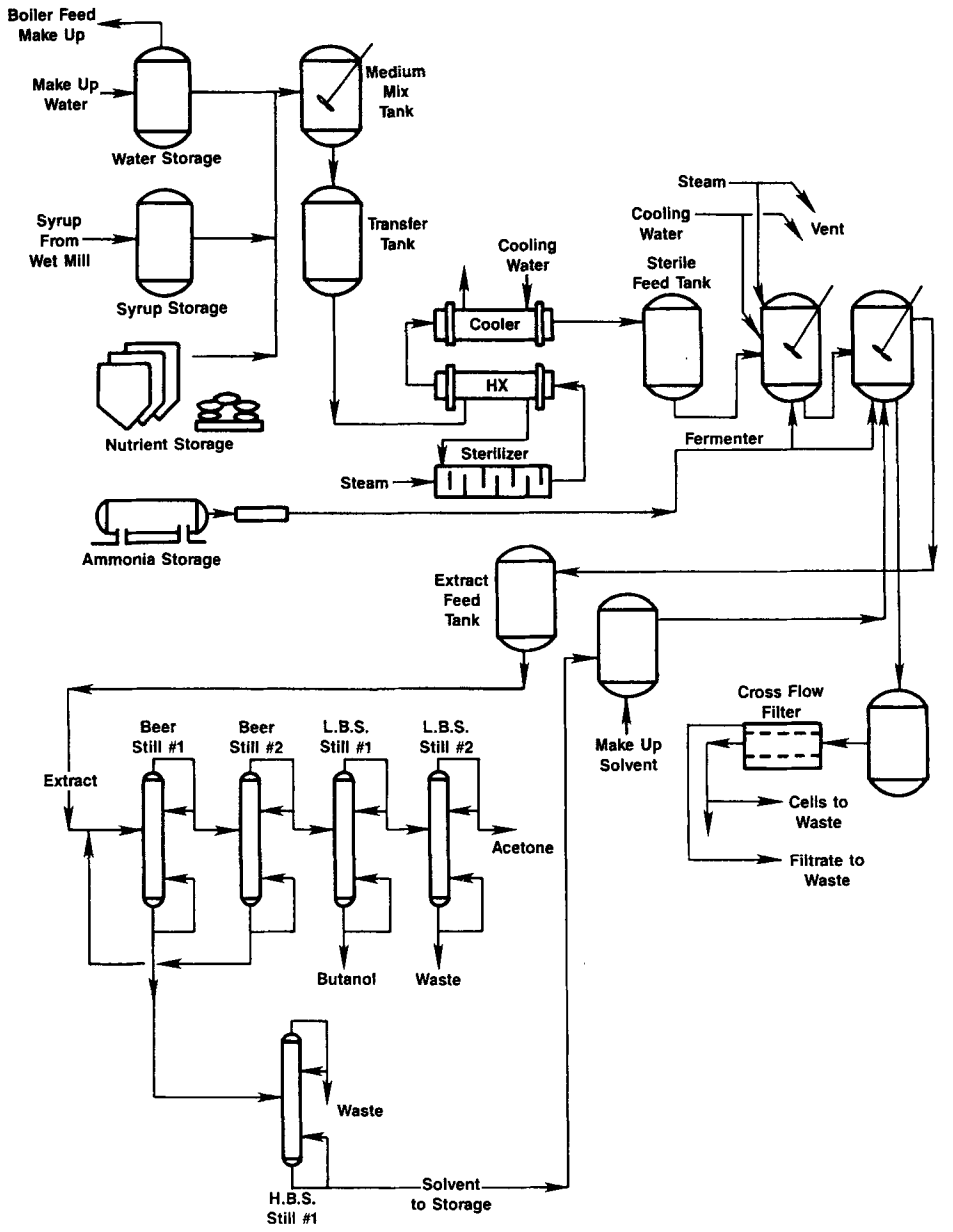

Fig. 3. Process Model Used in Economic Study

the beer ratio of 3.0, a distribution coefficient of this magnitude would lead to a partition of 92.8% of the butanol in the beer into the extract. The yield of solute to extract for other solvent/beer ratios and distribution coefficients is shown in Fig. 4. In Fig. 4 and subsequent plots the black dot denotes the basecase condition of the study.

It should be noted that although oleyl alcohol was used as a model solvent, it could not be used in commercial practice since it boils too high

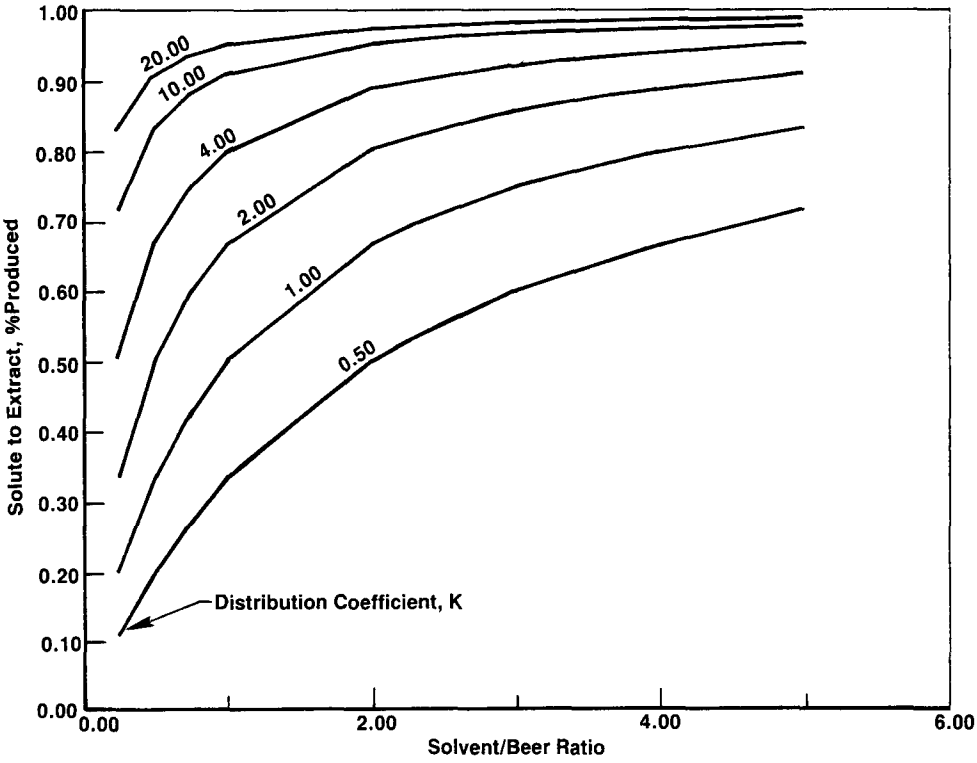

Fig. 4. Yield of solute to extract as percent of solute produced.

to allow a practical separation of product from the extract. It was assumed in this study that the oleyl alcohol was either in a carrier of nonyl alcohol or that a new solvent was identified that had the solvent characteristics of oleyl alcohol and the distillation characteristics of nonyl alcohol.

At a distribution coefficient of 4.3 and a solvent/beer ratio of 3.0, the final concentration of butanol in the beer is less than one-tenth the initial concentration (before extraction). Presumably, the fermentation can be operated at very high apparent concentrations without exceeding inhibitory levels; but this needs to be demonstrated. This effect is shown in Fig. 5. At a lower coefficient of 0.25 the extraction leverage is quite small.

BASECASE AND GOAL ECONOMICS

Wang compared the economics of extractive batch fermenter designs with conventional batch designs and concluded that the extraction design appeared to be considerably more profitable than the conventional design (23). Blanch made a similar conclusion with respect to a fed-batch extractive design (24).

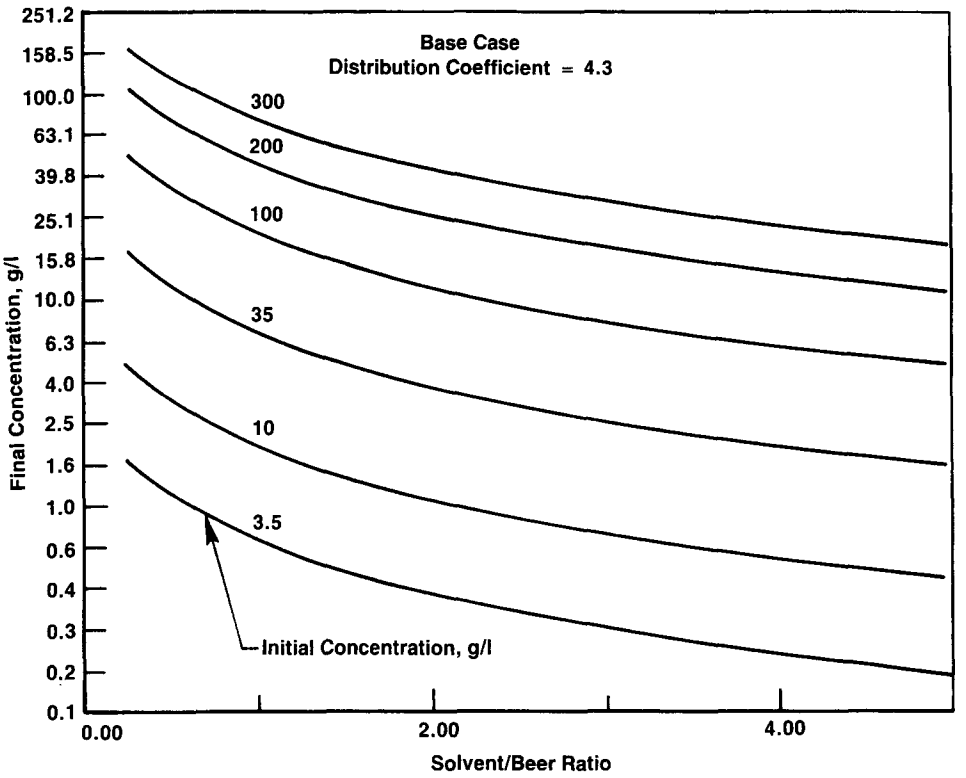

Fig. 5. Sensitivity of final concentration to solvent/beer ratio and initial concentration.

The innovations suggested by Battelle in the *in situ* single stage extraction of butanol from a fluidized fermentation broth would, if subsequently demonstrated in the next experimental phase, have a profound effect on increasing effective concentration of the product as much as tenfold while maintaining actual concentrations in the broth below the threshold of organism inhibition. To evaluate this possibility, an economic analysis was made based on the extrapolation of the current status of the process to plausible goal limits for the research program.

The results are summarized in Table 3, which compares the economics of a base case based on conventional product stoichiometry with the goal stoichiometry outlined in Table 1. Both cases assume that at a conventional dilution rate of 0.03 h^{-1}, a constant product exit concentration of 13 g/L (at the threshold of inhibition) is maintained in equilibrium with a solvent having a distribution coefficient of 4.3 (measured for oleyl alcohol) and at a solvent to aqueous beer ratio of 3:1. The resulting productivity is 4.L g/L/h.

This condition probably represents a limiting condition. To approach this productivity would probably require backing down on exit concentration to ensure that the fermentation remains at log phase production rates.

Table 3
Generalized Fermentation Economics
In Situ Extraction with Distillation Summary

Investment—$million	Base case	Goal case
Production Level—150 MM PPY; MPC = 1983		
Direct permanent investment	$26.7	$22.5
Allocated power, services and general	$ 3.8	$ 1.8
Working capital	$17.8	$11.6
Total investment	$48.3	$36.0
Cost-$/lb (1987)		
Raw materials	$0.42	$0.25
Utilities	$0.02	$0.01
Labor-related	$0.03	$0.03
Capital-related	$0.02	$0.02
Cost of manufacture	$0.49	$0.30
SE, D, R&D, Adm, and I.C.	$0.06	$0.05
Cost of sales	$0.55	$0.35
Pretax earnings based on: 30% pretax ROI	$0.10	$0.07
By-product credit	($0.05)	$0.00
Selling price	$0.59	$0.42
Financial Criteria		
Net ROI 3rd year (assumed)	16%	16%
Investors rate of return (20 operating years)	13%	18%
Year to break even—annual cash	1986	1986
—cumulative cash	1991	1989
—cum. disc. cash (NPV)	2004	1997
Net present value $MM (20 years @ 15%)	($3.9)	$4.6

For the base case stoichiometry a butanol selling price of $0.59 would be required to yield a 30% pretax return on $48 million in total investment. Similarly, as a result of reduced raw material requirements, the required price for the goal case drops to $0.42 compared with a current market price of $0.34 per pound for *n*-butanol. These values are very encouraging compared with prices over $1.30 for the state-of-the-art process.

Commercial acceptance of the enhanced fermentation process will ultimately depend on the direction taken by crude oil prices. This market is currently soft, at about $15–18 per barrel. However, James McNabb of Conoco (25) has pointed out that OPEC is presently operating at only 60% of capacity. By the early 1990s production is expected to reach 80%; and market power will shift from the buyer to the seller, with a correlary increase in oil prices. As a result, he forecasts that although oil prices will remain in the low $20s until 1990, they will reach the mid $30s by 1995,

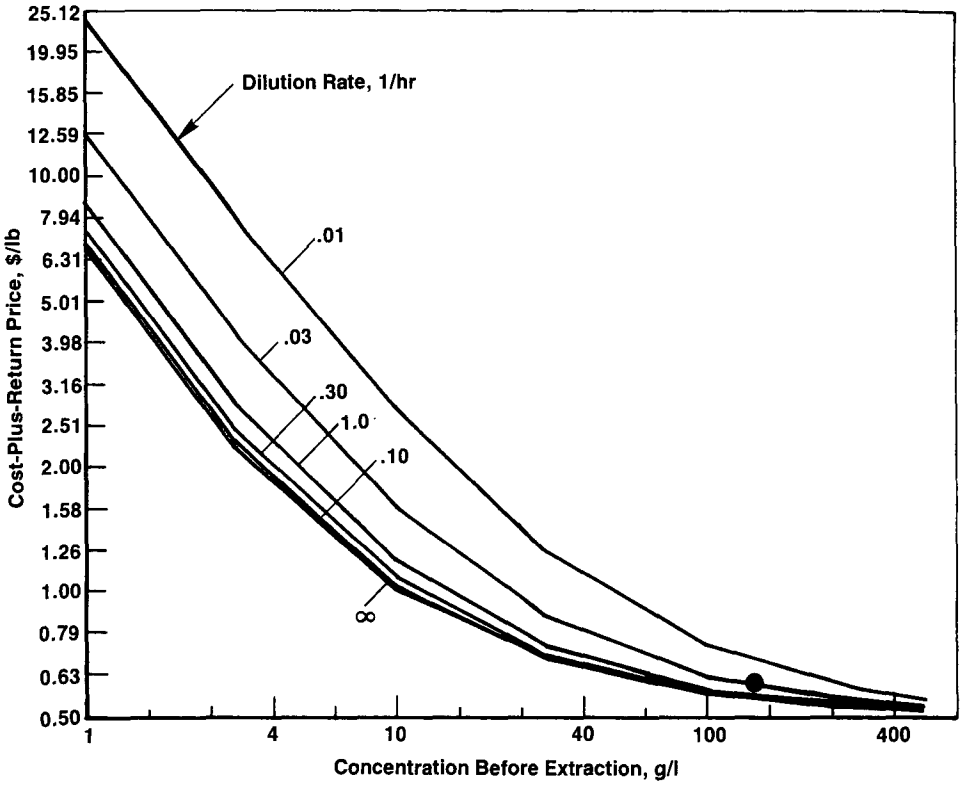

Fig. 6. Base case sensitivity price to dilution rate/concentration.

and $50 per barrel by the year 2000. Thus, a doubling in the $0.34 price for butanol over the next decade is not out of the question.

SENSITIVITY TO CHANGES IN CONDITIONS

As an aid to the evaluation, sensitivity plots were developed to explore the effects of changes in the more important parameters on cost. To simplify the calculations, these sensitivity analyses were made at a constant production level of butanol in the beer. Actual production levels for the recovered product could differ widely from this since it was assumed that unextracted product would be lost to the raffinate waste. Holding production level constant would have been more appropriate for comparisons but would have been too time-consuming to calculate within the time constraints of the study.

Sensitivity to changes in product concentration and dilution rate are shown in Fig. 6. Both cases appear to have gone as far as they can go in these parameters. The curve for an infinite dilution rate represents the economic limit to which fermentation rate can be pushed. Beyond this the investment in fermenters is insignificant compared with the investment

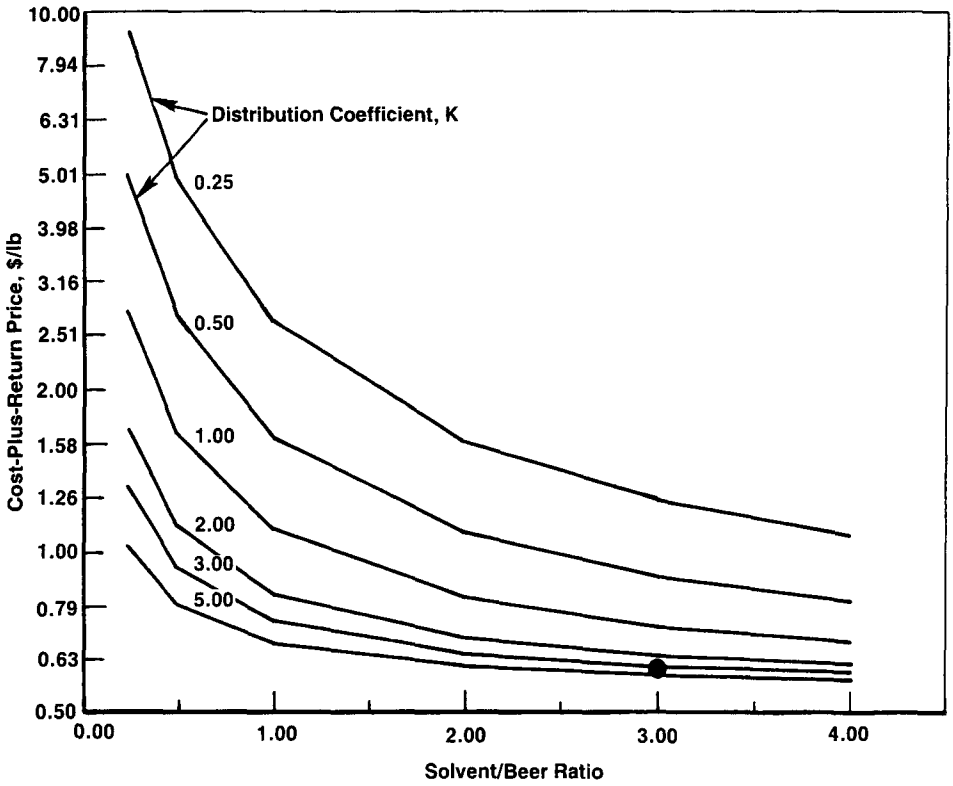

Fig. 7. Sensitivity of base case price to solvent/beer ratio and K.

in the rest of the plant. Note that the basecase at 138 g/L of butanol before extraction is equivalent to a concentration in contact with the organism of 12 g/L, i.e., at the Weizmann limit.

Sensitivity to changes in the solvent/beer ratio and the distribution coefficient is shown in Fig. 7. For high distribution coefficients, raising the solvent/beer ratio much above 1.0 does not appear warranted. There is a considerable adverse affect to not attaining a high coefficient.

As might be expected, price is very sensitive to the cost of sugar. For the base case, the cost of raw materials comprises 86% of cost of manufacture and 71% of cost-plus-return. Sensitivity to sugar price is shown in Fig. 8. The basecase would be competitive with current petroleum prices at a sugar price of $0.04 per pound. However, a price this low seems unreasonable.

CONCLUSIONS

Based on the economic analysis, it appears that the MPFB bioreactor system could substantially reduce the cost of butanol and other similar fermentation products that are now produced at low product concentra-

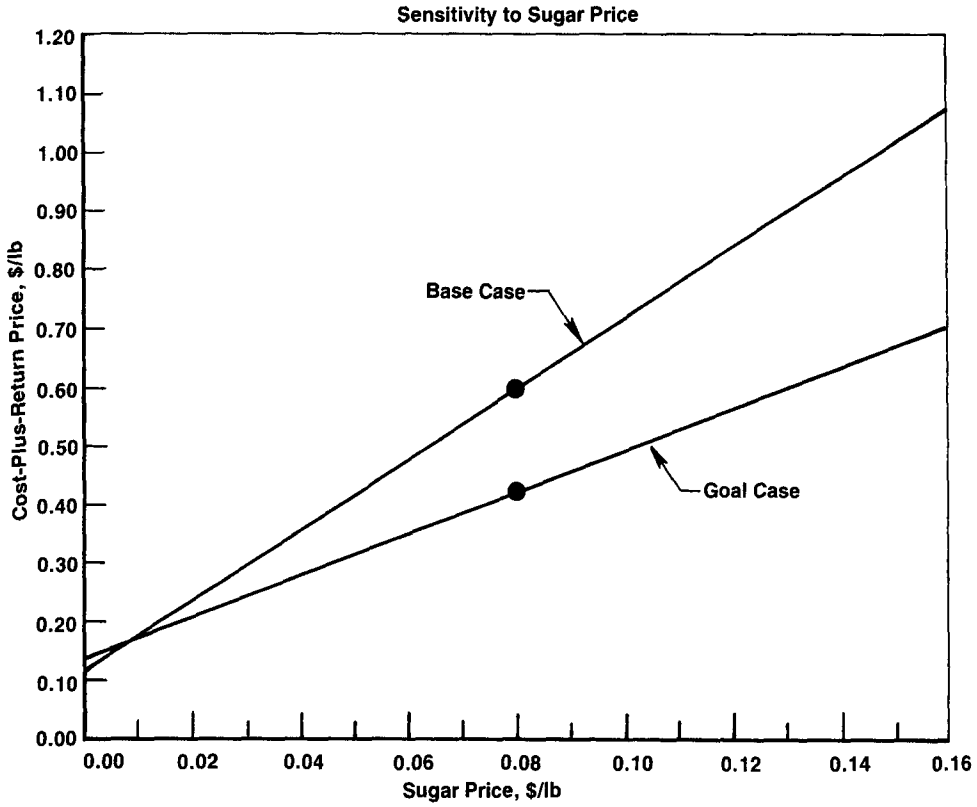

Fig. 8. Sensitivity to sugar price.

tions as a result of product inhibition. Such an economic breakthrough cannot be realized until the system has been fully optimized and scaled up for the specific fermentation process of interest. However, there appears to be no inherent design limitation in the MPFB bioreactor concept that would prevent a successful scaleup if sufficient time and funds are committed to the project.

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