0273-2289/92/3435—0441$02.00

Application of Novel Technology to the ABE Fermentation Process

An Economic Analysis

N. QURESHI AND I. S. MADDOX*

Biotechnology Department, Massey University, Palmerston North, New Zealand

ABSTRACT

Traditional technology for ABE production and recovery does not allow an economic process. However, use of a fluidized-bed reactor using immobilized cells in a continuous fermentation process, coupled with product removal and concentration by a membrane process, is a viable option. The major factor affecting product price is the substrate cost. Membrane selectivity and flux have little effect.

Index Entries: ABE fermentation; immobilized cells; fluidized-bed reactor; pervaporation; economic analysis.

INTRODUCTION

The fluctuations in oil prices during the last two decades have stimulated considerable interest in fuel and chemical production from renewable resources. Much of the emphasis has been on ethanol production, but both the butanediol and the acetone-butanol-ethanol (ABE) production processes have also received attention. The ABE process utilizes the anaerobic bacterium *Clostridium acetobutylicum,* which can ferment a wide range of sugars (*1*). Unfortunately, the traditional batch fermentation process suffers from problems of low reactor productivity (e.g., 0.3 g/L·h) and low product concentration (maximum ABE concentration 20 g/L). The latter is because of severe product inhibition and compels the use of only dilute sugar solutions as the fermentation raw material, which in

*Author to whom all correspondence and reprint requests should be addressed.

turn leads to the use of large process volumes and increased costs of product recovery.

The two major approaches that can be utilized to solve these problems are genetic manipulation of the organism and improvement of process technology. Work in this laboratory has concentrated on the latter. Cells of *C. acetobutylicum* have been immobilized by adsorption onto bonechar, and used in packed-bed and fluidized-bed reactors for continuous ABE production. Reactor productivities of up to 6 g/L·h have been achieved (*2,3*). Unfortunately, the problem remains of low product concentration in the reactor effluent, and this is accompanied by poor utilization of sugar. One solution to this problem is to recycle the reactor effluent, but because of product inhibition, this is unlikely to succeed without prior removal of the toxic fermentation products. Hence, the purpose of the present work was to develop an integrated continuous fermentation/product recovery technique for the ABE fermentation process. The reactor was based on immobilized cells, whereas the product removal/recovery technique used was pervaporation (*4*). Whey permeate was chosen as the raw material because of its potential commercial application (*5*), but the data have also been applied to molasses.

METHODS

Reactor for ABE Production

Cells were immobilized onto bonechar and used in a fluidized-bed reactor as previously described (*3*). It is assumed that the reactor is fed with whey permeate, concentrated by reverse osmosis from a lactose concentration of 46 g/L to 150 g/L (*6*), and this is diluted prior to entry to the reactor using recycled effluent from which ABE has been removed (*7*). A reactor productivity of 4.5 g/L·h, with a product yield of 0.38, is achieved in this system (*3*).

Membrane for Recovery and Concentration of ABE

Recovery and concentration were achieved by the process of pervaporation. The membrane modules were made from silicone tubing purchased from Elastomer Products Ltd. (Auckland, NZ). The selectivity and flux were 10 and 3–30 $g/m^2 \cdot h$, respectively. Based on experience, the working life is assumed to be 2 yr. The retail price of the membrane is \$9.00/m^2, but, based on experience when scaling up, it is assumed to be \$2.20/m^2 when purchased in bulk. Details of the membrane and its operational characteristics are given in Table 1. It is assumed that the series of membranes can produce an ABE stream containing sufficiently low water content not to affect its use in combustion engines. However, if necessary, dewatering can be achieved using gasoline (*8*). Gas sweeping

Table 1
Details of the Membrane and Operational Conditions Used for ABE Concentration

Fiber internal diameter	3.92 mm
Wall thickness	0.40 mm
Effective filtration area	0.4204 m^2
Total length of the membrane tube	17,050 mm
Total internal volume of the membrane	206 mL
Working life of the membrane*	2 yr
Membrane inlet pressure	55 kPa
Working temperature	35°C
Rate of liquid circulation through the membrane	30.6 L/h
Cold bath (ethylene glycol 25%) temperature (to cool vapors)	−3.0°C

*Based on experience.

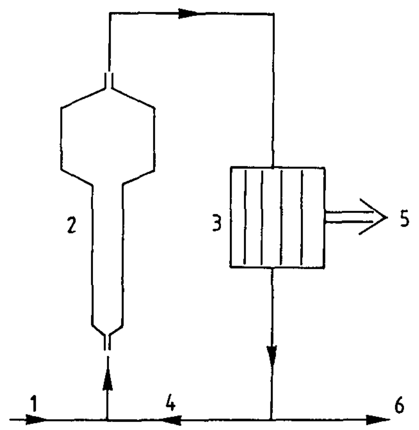

Fig. 1. Schematic diagram of integrated continuous fermentation/product removal process for ABE production. (1) Concentrated whey permeate feed, (2) fluidized-bed reactor, (3) membranes, (4) recycle of fermentation broth, (5) concentrated ABE stream, (6) bleed stream to reverse osmosis unit.

(oxygen-free nitrogen for ABE removal and air for subsequent concentration) is used as the means of removing ABE from the membrane surface. Condensation of ABE is achieved at −3°C. Figure 1 shows a schematic diagram of the integrated continuous fermentation/product recovery plant.

Substrate and Plant Size

Because of its availability in New Zealand, the major substrate considered is whey permeate. However, the use of molasses is also noted. Whey permeate, containing lactose at 46 g/L, would be supplied by a dairy factory to which the plant is annexed. A suitable plant size for New Zealand would be 300 m^3 of whey permeate/d, but sizes of up to 6000 m^3/d have been considered. In addition, a plant size of 20,000 m^3/d has been noted, since this scale of operation may be applicable in other countries. The permeate would be concentrated as noted above. For a flow of 300 m^3/d, the ABE production is taken as 1.94×10^6 L/y. For higher throughputs, the production is increased proportionately.

Financial Investments and Assumptions

All figures are given in US dollars, but all financial assumptions are based on those currently applicable in New Zealand. It is assumed that no capital is borrowed, no taxes are paid on the profit, and the rate of return on investment is 12% (ABE prices for higher rates of return can be calculated from the data provided). The working life of the plant is assumed to be 15 y with a straight-line depreciation of 10%. Design parameters and costs have been taken from standard texts (*9,10*), except those for the reverse osmosis plant, which were provided by Mawson (*6*). The fluidized-bed reactor and the alkali tank (for pH control) are made of stainless steel, whereas the remainder of the plant is made of carbon steel. A Lang factor of 1.5 has been used, since the plant is annexed to an existing dairy factory. Costs for the higher substrate throughputs were calculated using the six-tenth rule.

Whey permeate prices were quoted by the New Zealand Dairy Board (Wellington, NZ) as $116/t of solids. Molasses was assumed to be 50% fermentable sugars, and its price has been reported to be in the range of $25–75/per t (*11*).

It is assumed that all products sell at the same price, i.e., $0.68/L. No credits have been taken for byproducts. Operation of the plant is taken as 300 working days (typical New Zealand dairy season) and a throughput of 300 m^3 whey permeate/d, unless otherwise stated.

RESULTS AND DISCUSSION

Initially, an assessment was made of a production plant based on traditional batch fermentation technology using freely suspended cells, followed by product recovery by distillation. This was based on a previous assessment (*12*), with costs updated to 1990. The total capital investment was estimated as $\$3.4 \times 10^6$ with production expenses of $\$2.2 \times 10^6$, resulting in a product price of $1.43/L. Increasing the plant size to 6000 m^3 of whey permeate/d reduces the price to $0.89/L. At present, this is not an economically viable option.

For a plant using a fluidized-bed reactor for ABE production, and a membrane process for product removal and concentration, total capital investment would be $\$1.2 \times 10^6$ and production expenditure $\$1.5 \times 10^6$ (Table 2). This results in a product price of $0.76/L. The single most important factor to influence the ABE price is the cost of whey permeate. The effect of this cost on product price is shown in Fig. 2. This figure also

Table 2
Total Capital Investment and Manufacturing Expenses for ABE Production from Whey Permeate (Plant Capacity of 300 m^3 Whey Permeate/d)

Fixed capital investment	$1,098,696
Working capital	109,870
Total capital investment	$1,208,566
Manufacturing expenses	
Direct costs	
Raw material	$ 480,000
Labor	192,000
Utilities	50,660
Maintenance	54,935
Operating supplies	7,691
Laboratory charges	57,600
Indirect costs	
Depreciation	72,513
Insurance and taxes	54,935
Overheads	115,200
General	
Administration	72,513
Distribution	108,771
Research and development	60,428
Financing	145,028
Total manufacturing expenses	$1,472,274
Total production, L	1,940,000
ABE price	$0.76/L

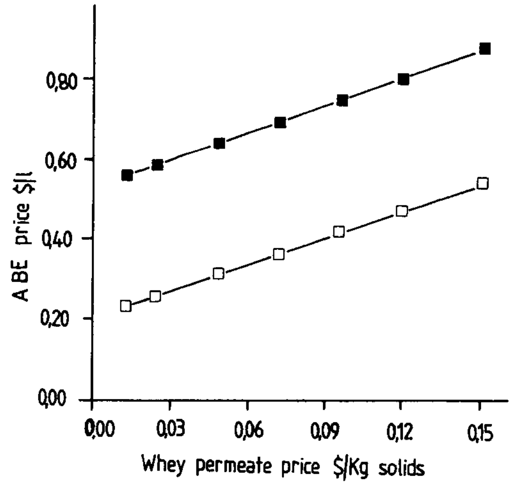

Fig. 2. Effect of whey permeate price and plant size on ABE price. ■ 300 m^3 whey permeate/d, □ 6000 m^3 whey permeate/d.

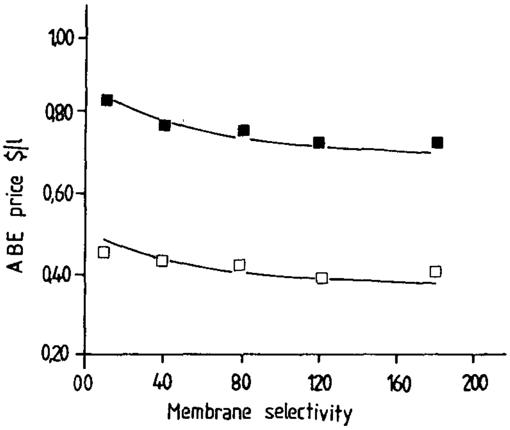

Fig. 3. Effect of membrane selectivity on ABE price. ■ 300 m^3 whey permeate/d, □ 6000 m^3 whey permeate/d. ABE flux 3–30 $g/m^2 \cdot h$.

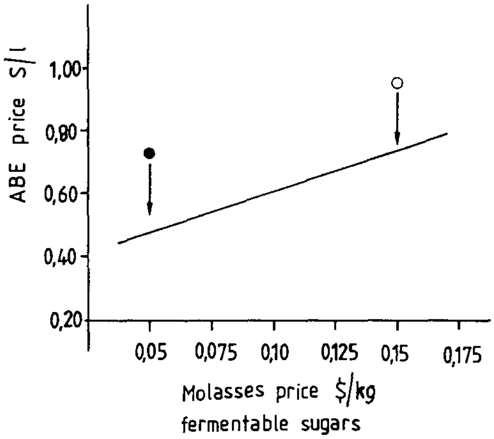

Fig. 4. Effect of molasses price on ABE price. ○ \$150/t; ● \$50/t. Plant size equivalent to 3000 m^3 whey permeate/d.

shows the ABE price at various whey permeate costs for a plant capacity of 6000 m^3 whey permeate/d. At the current substrate cost, the ABE price is \$0.44/Litre, whereas for a no-cost substrate, the price becomes \$0.19/L. For a plant of 20,000 m^3/d, at the current substrate cost, the price becomes \$0.37/L.

Membrane selectivity and flux are two factors that may influence ABE price. In the present work, a selectivity of 10 has been used, but values of up to 180 have been reported in the literature (*13–15*). Figure 3 indicates that an improvement in selectivity has no major effect on the product price. Similarly, the ABE flux through the membrane has little effect (approx \$0.06/L price reduction for a fivefold improvement in flux).

In addition to whey permeate, molasses has been considered as the fermentation substrate. This eliminates the need for a reverse osmosis

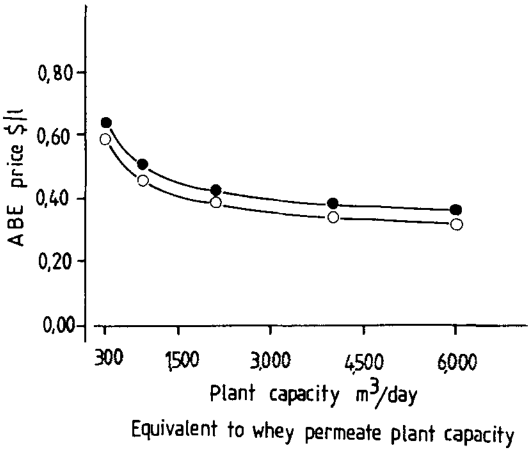

Fig. 5. Effect of plant size on ABE price, using molasses as substrate. ● Molasses at $100/t of fermentable sugars, ○ molasses at $77/t of fermentable sugars.

plant. For a production plant of size equivalent to 300 m^3 whey permeate/d, the ABE price is shown in Fig. 4. This varies from $0.48 to $0.74/L for molasses at $50 and $150/t fermentable sugars, respectively. The effect of plant capacity on ABE price is shown in Fig. 5. At a capacity equivalent to 6000 m^3 whey permeate/d, ABE price is in the range of $0.32 to $0.37/L.

In conclusion, it is evident that the necessary process technology exists for an economically viable ABE production process. However, much still depends on the cost of the fermentation substrate and the plant capacity.

REFERENCES

1. Jones, D. T. and Woods, D. R. (1986), *Microbiol. Rev.* **50,** 484–524.
2. Qureshi, N. and Maddox, I. S. (1987), *Enz. Microb. Technol.* **9,** 668–671.
3. Qureshi, N. and Maddox, I. S. (1988), *Bioproc. Eng.* **3,** 69–72.
4. Ennis, B. M., Gutierrez, N. A., and Maddox, I. S. (1986), *Process Biochem.* **21,** 131–147.
5. Maddox, I. S. (1980), *Biotechnol. Letts.* **2,** 493–498.
6. Mawson, A. J. (1990), *Fermentation Technologies: Industrial Applications,* Yu, P. L., eds., Elsevier Applied Science, London, pp. 264–270.
7. Qureshi, N. and Maddox, I. S. (1991), *Bioproc. Eng.* **6,** 63–69.
8. Schoutens, G. H. and Groot, W. J. (1985), *Process Biochem.* **20,** 117–121.
9. Ulrich, G. D. (1984), *A Guide to Chemical Engineering Process Design and Economics,* John Wiley & Sons, New York.
10. Baasel, W. D. (1990), *Preliminary Chemical Engineering Plant Design,* 2nd ed., Van Nostrand Reinhold, New York.
11. Paturau, J. M. (1989), *By-Products of the Cane Sugar Industry: An Introduction to Their Industrialized Utilization,* 3rd ed., Elsevier, New York, pp. 265, 268, 285.

12. Maddox, I. S., Gapes, J. R., and Larsen, V. F. (1981), *Proc. Chemeca 81: 9th Australasian Conference on Chemical Engineering,* Aug. 30–Sept, 4, pp. 535–542.
13. Groot, W. J., van der Oever, C. E., and Kossen, N. W. F. (1984), *Biotechnol. Letts.* **11,** 709–714.
14. Larrayoz, M. A. and Puigjar, L. (1987), *Biotechnol. Bioeng.* **30,** 692–696.
15. Matsumura, M. and Kataoka, H. (1987), *Biotechnol. Bioeng.* **30,** 887–893.