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Simulation and Optimization of a Cell Recycle System

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ABSTRACT

This article describes the simulation and subsequent optimization of a cell recycle fermentor and membrane filter in a wheat-to-ethanol process. The optimum conditions with respect to economic operating performance were found using a search method similar to steepest gradients. Optimum fermentor cell and inlet substrate concentrations were 54 and 127 g/L, respectively. The study also involved the effects of parameter variations on net cost, and the effects of raw material and product prices on profitability. The net cost was found to be most sensitive to the product yield and wheat costs.

Index Entries: Ethanol; cell recycle; membrane filtration; optimizaton.

NOMENCLATURE

a	Product inhibition constant for the growth rate (1.1)
A	Membrane area available to filtration (m^2)
C_{DIST}	Cost associated with distillation, evaporating, and drying ($/h)
C_{FERM}	Cost associated with the fermentor ($/h)
C_{MEMB}	Cost associated with the membrane ($/h)
C_{SUBS}	Cost associated with the wheat and its hydrolyzation ($/h)
F_{CR}	Crossflow flow rate in the filter (m^3/h)
F_F	Feed flow rate into the fermentor (m^3/h)
F_{PR}	Flow rate of ethanol out of the distillation column (m^3/h)
FLUX	Flux through the membrane ($m^3/m^2/h$)

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K_{M1}	Pressure dependence of the filter cake resistance (0.76)
K_{M2}	Shear dependence of the filter cake resistance (−0.1)
K_{M3}	Cell concentration dependence of the filter cake resistance (0.45)
K_{M4}	Flow rate dependence of the pressure drop along the filter (1.2)
K_{M5}	Cell concentration constant for the pressure drop along the filter (0.0046)
K_S	Monod constant (0.4 g/L)
p	Product concentration in the fermentor and the distillation inlet stream (g/L)
P_I	Inlet pressure to the filter (bar)
p_{m1}	Product inhibition constant for the growth rate (95 g/L)
p_{m2}	Product inhibition constant for cell yield (g/L)
p_{m3}	Product inhibition constant for the ratio of cell to product yield (g/L)
P_O	Outlet pressure from the filter (bar)
s	Utilizable sugar concentration in the fermentor (g/L)
s_F	Utilizable sugars concentration in the fermentor feed (g/L)
V_F	Fermentor working volume (m^3)
x	Cell concentration in the fermentor (g dry wt/L)
$Y_{P/S}$	Product yield (g/g)
$(Y_{P/S})_O$	Product yield at zero ethanol concentration (g/g)
$Y_{X/P}$	Ratio of cell to product yield (g/g)
$(Y_{X/P})_O$	Ratio of cell to product yield at zero ethanol concentration (g/g)
$Y_{X/S}$	Cell yield (g/g)
$(Y_{X/S})_O$	Cell yield at zero ethanol concentration (0.11 g/g)

GREEK SYMBOLS

μ	Growth rate (1/h)
μ_m	Maximum growth rate (0.357/h)

INTRODUCTION

The full potential of ethanol as a "renewable" automotive fuel has so far been limited by its economics compared to gasoline. Although in the long term, as gasoline reserves continue to decline, these economics should improve, the key to allowing ethanol to compete in the short term without government subsidies may well be better processing technology. One area in which this has occurred in some of the newer plants is the use of continuous fermentation. This has allowed significantly lower fermentor volumes to be used compared to the previous batch technology, as well as allowing a smoother overall operation. Most of the continuous systems presently in operation are cascade systems (*1*), where a series of

continuously stirred tank fermentors, each with a progressively higher ethanol concentration, are used. Although these systems reduce the effects of product inhibition, they still only allow low fermentor cell concentrations. It follows that the next development is to use higher cell concentrations, thereby allowing the fermentor volume to be further reduced.

Two methods have been suggested to obtain higher cell concentrations: either immobilization or cell recycle. A good review of bench-top scale immobilization systems is given by Godia et al. (*2*). Semicommercial production of ethanol using cells entrapped in calcium alginate has also been reported (*3*), but no major immobilized ethanol-producing facilities exist in North America.

To achieve cell recycle, the cells have to be separated from the exit stream. Three main techniques have been suggested for this separation:

1. Sedimentation;
2. Centrifugation; and
3. Filtration.

Initial cell recycle experiments used sedimentation (*4–8*). By its nature, sedimentation produces an imperfect separation, especially as the liquid velocity in the settler is increased. Cooling may also be required in the settler to help the cells sediment. Because of these drawbacks, this method has not been used on a large scale. Commercial cell recycle systems, for example the Biostill process (*9*), presently use centrifugation. However, because continuous centrifuges are very expensive, most of the recent experimental work on cell recycle has been with membrane filters (*10–14*). These devices allow sterile separation with very high recovery efficiencies. The system is also a lot simpler to operate than a centrifuge and has the potential to be considerably cheaper. This article presents the economic optimization of a cell recycle simulation using crossflow membrane filtration for a new continuous ethanol plant design.

PLANT DESIGN

The plant was designed with a capacity to produce 100 million L of 99.2% wt/vol ethanol/yr. The site is based near an urban center in western Canada and uses a feed-grade wheat grown nearby, with the byproducts being sold to local farmers. A flow sheet of the plant is shown in Figure 1. Although the principal change in the design is the use of a cell recycle fermentor, this caused a number of changes elsewhere in the plant. These include the separation of suspended solids before saccharification and the use of an immobilized enzyme saccharification column. The cost of the process was based on the report by R. Katzen Associates for the US Department of Energy (*15*), except for the saccharification column (*16*), the fermentor (*17*), the membrane filters (*18*), and the distillation columns,

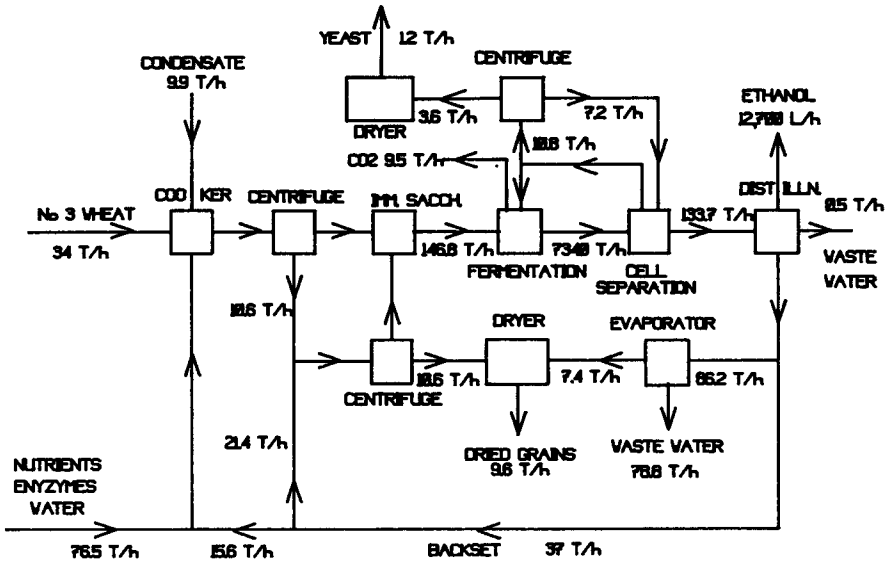

Fig. 1. Schematic diagram of a 10^8 L/yr ethanol plant using cell recycle showing typical flow rates from the simulation.

Table 1
Important Raw Material, Energy, and Product Costs and Yields
Used in the Simulation and Optimization of a 10^8 L 99.2% w/v Ethanol Production Facility

Raw Material	Cost $CAN	Utility	Cost $CAN	Product	Cost $CAN
Wheat,*/T (62.6% starch)	105	Natural gas*/1000M^3	100	Ethanol[24]/1000 L	460
Gelatinization enzymes/T	5000	Electricity*/GWh	250	Dried grains*/T	180
Ammonia/T	250	Water*/1000M^3	4	Dried yeast[25]/T	350

Stage	Cooking	Saccharification	Overall
% Yield by weight	98	97	28
	Starch to dextrins	Dextrins to utilizable sugars	Wheat to dried grains

*Costs taken from values for Saskatoon.

driers, and boilers (*19*). These prices were then converted to 1990 values using the Chemical Engineering Price Index (*20*) and to the required size using size exponents from Garrett (*21*) and Kalk and Landglykke (*22*). The total capital costs for a plant assuming 7.4% wt/vol entering the distillation column were estimated at \$CAN 5.1×10^7. The prices of raw materials, energy sources, and products used in the optimization are given in Table 1. These prices were taken from values for a commercial ethanol plant currently operating in western Canada, except the yeast selling price, which was halved because laboratory kinetics have been shown to give twice the yield of commercial operations. Further information on the design and costing can be found in Warren (*25*).

FERMENTOR SIMULATION AND OPTIMIZATION

An accurate simulation of each stage, followed by an overall optimization of the plant, was out of the scope of this project. Instead, it was decided to concentrate on the fermentor and membrane filter. However, changes in these stages affect conditions in the rest of the plant, and these changes have to be taken into account in any costing equation.

For the fermentor simulation, the intrinsic kinetics for *Saccharomyces cerevisiae* previously presented by Warren et al. (*26*) were used. The values of the parameters obtained from Ghose and Tyagi's (*27*) data were used, except for $(Y_{X/P})_O$. These values are given in the nomenclature, except for the following, which vary depending on the model used:

1. Variable cell and product yield model—$(Y_{X/P})_O = 0.144$ g/g; $p_{m2} =$ 370 g/L; $p_{m3} = 660$ g/L.
2. Variable cell and constant product yield model—$(Y_{X/P})_O = 0.234$ g/g; $p_{m2} = p_{m3} = 370$ g/L.
3. Constant cell and product yield model—$(Y_{X/P})_O = 0.234$ g/g; $p_{m2} = p_{m3} = \infty$ g/L.

For the cell inhibition models, the growth rate is multiplied by $(1 - x/300)$. The initial equations for this model are shown below:

$$\mu = \mu_m (s / s + K_s) [1 - (p/p_{m1})^a] \quad (1)$$

$$Y_{X/S} = (Y_{X/S})_O (1 - p/p_{m2}) \quad (2)$$

$$Y_{X/P} = (Y_{X/P})_O (1 - p/p_{m3}) \quad (3)$$

Ghose and Tyagi's (*27*) data were obtained with a continuous stirred tank fermentor at low cell concentrations with exogenous ethanol. These conditions are different than would exist in a cell recycle fermentor. Previous cell recycle models have suggested that two extra terms may be needed to take into account the change in environment. The first is a cell inhibition term initially suggested by Lee et al. (*28*), which Lee and Chang (*14*) have since used to model experimental data. However, Lee and Chang's (*14*) work used ultrafilters, whose pore sizes are small enough to retain molecules that may be toxic in low concentrations. It is these molecules that probably caused the inhibitory effect. If microfilters are used, this inhibition is expected to be much less severe. However, since the addition of a cell inhibition term to the model is very simple, its effect has also been investigated here. The second extra complication suggested by Jarzebrski et al. (*29*) takes into account the fraction of the cells that are viable, since it has been suggested that the percentage viability decreases dramatically as the cell concentration increases. However, this model was based on previous data presented by Lafforgue et al. (*13*) obtained with a cell recycle system with no bleed. As has been previously stated by Cheryan and Mehaia (*30*), a cell bleed is necessary to remove dead or inactive cells and macromolecular products of metabolism that do not

permeate the membrane. When a sufficient bleed was used, Cysewski and Wilke (*4*) found that no change in cell viability occurred over 14 d.

The model for the membrane filter shown below was taken from Warren et al. (*31*).

$$FLUX = 0.11\{[(P_I - P_O)/2]^{1-KM1}(F_{CR}/A)^{-KM2}X^{-KM3} \quad (4)$$

$$P_O = P_I - 0.108(F_{CR}/A)^{KM4}(1 + KM5\,x)^2 \quad (5)$$

The effects of scale-up were assumed to be negligible, since the membrane surface area can be increased by increasing the number of tubes. Because of the much shorter residence time, no reaction was assumed to occur in the filter or the tubing connecting it to the fermentor.

The cost functions/h for a 10^8 L/yr anhydrous ethanol plant shown below were calculated for:

1. The grain and the stages upstream of the fermentor;
2. The fermentor;
3. The membrane filter; and
4. The stages downstream of the filter.

The sum of these cost functions minus the credit for the byproducts then gives the production cost/h, from which a cost/1000 L can be obtained.

$$C_{SUBS} = 0.2\,s_F F_F \quad (6)$$

$$C_{FERM} = 0.83(V_F)^{0.7} \quad (7)$$

$$C_{MEMB} = 0.05A^{0.7} + 0.06A^{0.8} + 0.001\,P_I F_{CR} \quad (8)$$

$$C_{DIST} = (200 - 1.9p + 0.011p^2)F_{PR} \quad (9)$$

The cost/h for the grain and those stages upstream from the fermentor was calculated based on the capital and utility costs directly related to it and a third of the personnel costs. An assumption was made that the flow rate of backset going directly to the fermentor is limited by the concentration of unfermented solids rather than as a fraction of the new water added, with the extra backset being added after cooking. As such, the amount of new water added and the starch concentration entering the cooker do not vary when the inlet substrate concentration changes, so the variation in cost with substrate concentration is negligible. The cost/h of the fermentor was calculated using a size exponent of 0.7 as given by Kalk and Langlykke (*22*). The cost of the membrane itself was calculated based on a size exponent of 0.8, similar to that shown graphically by van Gassel and Ripperger (*32*) and a size exponent of 0.7 for the supporting equipment. Since costs for the size of hollow-fiber microfilter required here were not available, the pricing was based on scale-up of the largest presently available membrane produced by A/G Technology (*18*). It was also assumed that the membrane would need replacement once every 2 yr. The cost/h of the downstream stages was calculated based on the capital and utility

costs directly related to it and two-thirds of the personnel costs. Because the ethanol concentration fed to the distillation column varies, significant changes in the size of the first distillation column and in the energy used in this column occur. Figures for how these vary are given graphically by Zacchi and Axelsson (*33*). From these figures, a second-order correlation for cost with respect to the inlet ethanol concentration was derived.

The next step was to find the best operating conditions to minimize the overall production costs. Fortunately, this problem can be simplified since three of the variables, the inlet membrane filter pressure, the filter crossflow flow rate, and the filter area, are important only in the filter. So, if the other variables in the filter are held constant, the filter can be optimized analytically using the method of Lagrange multipliers. This gives a set of simultaneous nonlinear equations that can then be solved numerically using Newton's method. For the filter modeled here, the inlet pressure must be < 0.8 bar. When optimized, for all cases, this was found to be the optimum pressure. As such, in the overall optimization routine, this was taken as the inletpressure to the filter.

The remaining independent variables (cell concentration, inlet substrate concentration, and fermentor substrate concentration) were then optimized numerically using a variant of the steepest gradient method The method used differs from the steepest gradient in that the constant to determine the change in each variable from the partial derivative of the cost function depends on the value of each variable. This change was made because the steepest gradient method becomes very slow as the correct solution is approached. Changing these constants allowed a faster approach to the correct solution. The problem is also constrained, although the steepest gradient is an unconstrained optimization technique. By varying the constants, it is also possible to make sure the problem stays within the constraints.

RESULTS AND DISCUSSION

Figure 2 demonstrates the effect of varying inlet fermentor substrate and fermentor cell concentrations on the total net operating costs. From this figure, it can be seen that, as the cell concentration increases, the minima shift to higher substrate concentrations. The minima at a given cell concentration can also be seen to be fairly flat, an important point for operating conditions since it means that small deviations from the "optimum" conditions are unlikely to cause major cost increases. Figure 3 shows a breakdown of the costs for varying inlet fermentor substrate concentrations. The increase in substrate costs with substrate concentration occurs as the product yield decreases, so more wheat is required per 1000 L, whereas the decrease in membrane costs occurs as a lower filtrate flow rate is required per 1000 L owing to the higher ethanol concentration.

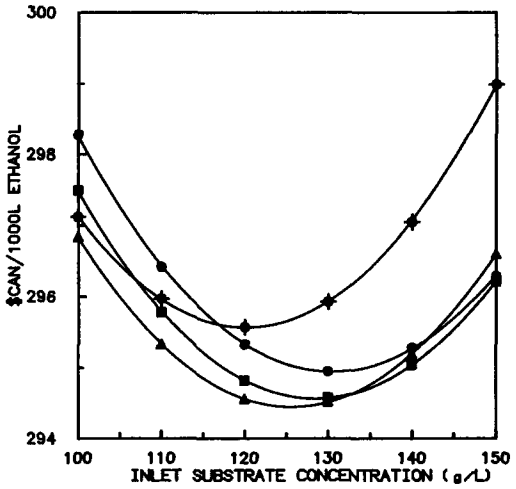

Fig. 2. Inlet fermentor substrate and fermentor biomass effects on the total operating cost of producing 1000 L of 99.2% wt/vol ethanol (variable cell and ethanol yields; fermentor substrate concentration of 0.5 g/L). Cell concentration gDW/L: ✦✦✦✦✦ 25; ▲▲▲▲▲ 50; ■■■■■ 75; ●●●●● 100.

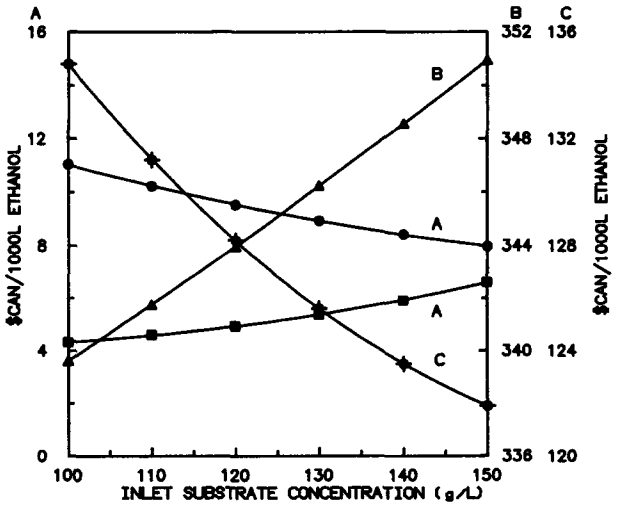

Fig. 3. Inlet fermentor substrate effects on the operating costs of different stages of producing 1000 L of 99.2% wt/vol ethanol (variable cell and product yields; fermentor substrate concentration 0.5 g/L; fermentor biomass concentration 50 g/L). Stage: ▲▲▲▲▲ wheat/hydrolysis; ■■■■■ fermentor; ●●●●● membrane; ✦✦✦✦✦ distillation/drying.

Table 2
Values of Fermentor Conditions from the Optimization Routine

Model	Inlet substrate concentration, g/L	Fermentor substrate concentration, g/L	Fermentor cell concentration, g/L	Projected net cost, $CAN/1000 L ethanol
Variable $Y_{X/S}$ and $Y_{P/S}$	127	0.5	54	294
Variable $Y_{X/S}$ and $Y_{P/S}$ with cell inhibition	125	0.6	48	295
Variable $Y_{X/S}$, constant $Y_{P/S}$	155	0.6	88	285
Variable $Y_{X/S}$, constant $Y_{P/S}$ with cell inhibition	152	0.7	68	286
Constant $Y_{X/S}$ and $Y_{P/S}$	162	0.7	125	272
Constant $Y_{X/S}$ and $Y_{P/S}$ with cell inhibition	157	0.8	82	274
Variable $Y_{X/S}$ and $Y_{P/S}$ with twice the membrane flux	125	0.4	96	290

The optimization routine was then used to calculate the best conditions for a variety of biokinetic models. The results from this work are given in Table 2, from which it can be seen that significant differences occur among the different models. A variable product yield favors a lower substrate concentration. This lower substrate concentration, in turn, favors a lower cell concentration because (i) the same ethanol productivity can be obtained at a lower cell concentration because of lower product inhibition, and (ii) a lower cell concentration will give higher fluxes that will compensate for the higher filtrate flow rate required through the membrane. The significant difference in price as the yield model is changed occur because of changes in the value of cell and product yields. The effect of adding a cell inhibition term can be seen to be insignificant for the variable product yield model, whereas if the product yield is constant, the fermentor cell concentration and to a lesser extent the inlet fermentor substrate concentration decrease. It can also be seen from Table 2 that, if the flux can be doubled in the membrane, the optimum fermentation cell concentration also approximately doubles. The inlet substrate concentrations found from the models presented here are all lower than the values given by Maiorella et al. (*17*) using sugar as a feedstock and with constant cell and product yields (*17*). This is because Maiorella used no backset, so sterilization and evaporation costs decreased as substrate concentration increased, which then favored higher inlet substrate concentrations.

Table 3
Sensitivity of the Optimized Net Operating Cost to Changes in the Fermentor Parameters and the Maximized Profit to Changes in Raw Material, Energy, and Product Costs

Fermentor parameter	μ_m	K_S	p_{m1}	$(Y_{X/S})_O$	$(Y_{P/S})_O$
% Change in operating cost per % change in parameter	−0.012	0.004	−0.073	−0.18	−1.0
Raw material/energy/product	Wheat		Natural gas	Dried grains	Dried yeast
% Change in pretax profit per % change in cost	4.2		0.50	2.1	0.85

Table 3 shows the effect of altering fermentor parameters on the optimum values for the varying cell and product yield model. The greatest sensitivity is to the product yield, followed by the cell yield. This is because the most significant cost in producing ethanol is the grain, whereas an increase in cell yield allows a higher byproduct credit. The other parameters affect only the cost of the fermentor, and from Fig. 3, it can be seen that changes in the substrate cost are much more significant.

Using the prices given in Table 1, it was estimated that the recycle fermentation plant studied here would generate an annual pretax profit of \$CAN 6×10^6. The effect of changing the raw material, energy, and byproduct selling prices on the profitability of the future plant for the variable cell and product model is shown in Table 3. Not surprisingly, the parameter with the greatest sensitivity is the wheat cost, where a 10% cost change can dramatically alter the profitability of the plant.

CONCLUSIONS

A continuous gasohol process has been simulated in order to determine the sensitivity of optimum costs and profitability on the values of operating conditions and biokinetic models. The simulation used realistic capital and operating costs typical of those in western Canada.

Variations in the fermentor model used were found to alter the optimum concentrations in a cell recycle system dramatically and exerted a significant influence on the net costs and profitability. For instance, fermentor cell concentrations varied from 48 to 125 g/L.

The most important parameters in the fermentation model are those involving the yields, especially the product yield. Further study of the dependence of yields on fermentor conditions is required in cell recycle systems, since significant savings may be possible. The cost of the mem-

brane in this system is greater than the fermentor. In any future system, a significant increase in flux over the experimentally measured values (*31*) of 10–20 L/h/m^2 would be required.

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