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Aeration costs in stirred-tank and bubble column bioreactors

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ABSTRACT

To overcome knowledge gaps in the economics of large-scale aeration for production of commodity products, Aspen Plus is used to simulate steady-state oxygen delivery in both stirred-tank and bubble column bioreactors, using published engineering correlations for oxygen mass transfer as a function of aeration rate and power input, coupled with new equipment cost estimates developed in Aspen Capital Cost Estimator and validated against vendor quotations. These simulations describe the cost efficiency of oxygen delivery as a function of oxygen uptake rate and vessel size, and show that capital and operating costs for oxygen delivery drop considerably moving from standard-size (200 m^3) to world-class size (500 m^3) reactors, but only marginally in further scaling up to hypothetically large (1000 m^3) reactors. This analysis suggests bubble-column reactor systems can reduce overall costs for oxygen delivery by 10–20% relative to stirred tanks at low to moderate oxygen transfer rates up to 150 mmol/L-h.

1. Introduction

Since the development of penicillin production by submerged aerobic cultivation of *Penicillium chrysogenum*, aerobic biological production ("aerobic fermentation") has been used to produce an increasing variety of chemical products [1]. The range of products being produced or considered for biological production has grown rapidly with recent advances in metabolic engineering, synthetic biology and bio-based production technologies [2,3]. Owing largely to the relatively high cost of supplying molecular oxygen (O_2) to a submerged culture, aerobic fermentation has historically primarily been applied to produce lower volume, higher value (higher margin) compounds like pharmaceuticals and specialty chemicals. The challenges to achieving economical aerobic production become greater for larger volume, lower margin products where material and utility costs generally dominate fermentation economics [4]. The higher capital and operating costs for aerobic production are well recognized [5,6] and are also stimulating research and development on anaerobic routes for biological production. [7,8]. Cost constraints become the most acute for economic aerobic production of extremely low-margin, high-volume commodity products like biofuels, and this motivated us to assess aeration costs for large-scale aerobic production in the context of such products.

Aerobic fermentation is a critical unit operation in the process of making fuel-range hydrocarbons from sugars, when the hydrocarbon or its precursor, e.g., a lipid or free fatty acid, is directly produced in submerged culture by a microorganism. However, little public domain information exists about state-of-the-art designs and economics of large-scale aerobic bioprocesses, especially for those producing low-margin, commodity products like biofuels where extreme cost minimization is required. Previous techno-economic analysis (TEA) reports from the National Renewable Energy Laboratory (NREL) [9–11] demonstrated that, in the case of cellulosic ethanol, the fermentation area (comprising mechanically simple but extremely large anaerobic fermentation vessels up to 10^6 gallons) was not a primary cost contributor, generally falling behind larger cost drivers including biomass pretreatment, cellulase enzyme production/purchase, and wastewater treatment. A more recent TEA report [12] examined the aerobic conversion of lignocellulosic sugars to hydrocarbons by way of a fatty acid intermediate. In contrast to the earlier ethanol analyses [12], concluded that the aerobic fermentation area *was* a primary cost contributor for integrated cellulosic biofuel production; in fact, it was the largest contributor of all process areas, with fermentation compressors and agitators also representing the largest power demand in the biorefinery process.

As the first publicly available TEA for such a technology pathway, the [12] analysis carried a higher degree of uncertainty in its underlying process design and capital cost assumptions than more established pathway concepts. Key among such uncertainties were the operating conditions, performance parameters, and cost con-

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tributions of the aerobic fermentation step. Parts of the process design were supported by a partner engineering company, which provided initial input on the design of stirred-tank aerobic bioreactors, and associated capital cost estimates. Since publication, this design has been reviewed by several engineering companies and consultants, including Harris Group, Katzen International, Benz Technologies, and Genomatica. Following these critical reviews, it was concluded that several assumptions pertaining to the operational performance and capital costs used in [12] erred on the optimistic side for commercial-scale aerobic fermentation.

This article documents efforts to reduce uncertainty in such key process and cost parameters through (1) development of an independent framework for bioreactor cost estimation that is validated against high-quality vendor quotes and (2) modeling of low-viscosity, aerobic fermentation using simple design equations and a steady-state process simulator, to understand achievable oxygen transfer rates as a function of vessel configuration, power input, and aeration rates. Together, these developments will be used to guide future refinements in conceptual design and TEA of biochemical conversion processes.

2. Calculations

2.1. Bioreactor capital cost estimation

Ideally, well validated TEA studies should favor direct capital equipment inputs/quotations from equipment vendors, especially for critical and costly items like fermentors. However, external estimates are not always readily obtained for conceptual studies and can be scattered and sometimes conflicting. To facilitate rapid comparative analysis across multiple technology options, methods for consistent estimation of bioreactor capital costs were developed using Aspen Capital Cost Estimator (ACCE). ACCE estimates costs for individual equipment items using volumetric models (as opposed to factored models), which compute total estimated materials and labor involved in building a piece of equipment based on its specified size. For example, if a specified vessel is too large or heavy to be transported in one piece (and all the vessels considered in this article are), ACCE will compute costs for shop fabrication of transportable pieces and costs for final field fabrication, all as part of a total bare equipment cost. A variety of high-quality bioreactor quotes obtained over recent years from vendors and engineering firms were used to create general guidelines for specifying fermentation vessels in ACCE, resulting in reasonably accurate, absolute capital cost predictions for both stirred-tank reactors (STRs) and bubble-column reactors (BCRs).

With these specification guidelines, a set of capital costs for STRs and BCRs were developed at different standard vessel volumes. Economies of scale naturally dictate use of the largest reaction vessel possible; however, while million-gallon (3800 m^3) *anaerobic* fermentors are in use at industrial fuel ethanol plants, the maximum practical *aerobic* reactor volume is less clear. Our industry contacts have intimated that the largest STRs in operation are in the hundreds of cubic meters ($\leq$500 m^3) and their ultimate maximum size must be on the order of 1000 m^3, owing to diminishing returns on oxygen transfer relative to volumetric power input, as well as practical limitations regarding the fabrication and maintenance of very large impellers, shafts, bearings, and motors. Bubble columns are not limited by moving parts and BCRs up to 1000 m^3 are known to be in operation [6]. Costs were therefore estimated for BCRs and STRs at three standard vessel sizes, with the understanding that uncertainty in cost increases with vessel size:

- 200 m^3, representing an "off-the-shelf," readily purchasable reactor.
- 500 m^3, representing a "world's-largest" class of reactor that exists in relatively small numbers.
- 1000 m^3, representing a "hypothetically large" reactor that may or may not exist today, representing a ceiling for what is likely viable from a design and operational standpoint.

2.2. Flowsheet simulation of aerobic fermentation

To accompany the new standard bioreactor capital costs, we carried out steady-state flowsheet simulations in Aspen Plus to investigate the operating costs associated with aeration power demand. Fig. 1 depicts the general bioreactor schematic considered here, equally applicable to STR or BCR vessels aside from the depicted agitator; a bioreactor is part of a complex of interacting energized systems, including an agitator (for STRs, eliminated for BCRs), an air compressor with discharge cooler, and a chilled-water system for temperature control, itself connected to a larger cooling water system. The system shown in Fig. 1 uses forced-circulation heat removal, but jackets or coils may be favored instead, depending on sterility and shear stress concerns. Aspen Plus simulations of STRs and similarly-equipped BCRs of 200, 500, and 1000 m^3 were performed to determine the total system power demand for most of the users shown in Fig. 1: air compressor, air cooler, agitator for STRs, circulation pump, and chiller (scaled by cooling duty). The cooling tower was not included because its power contribution is insignificant compared to the chiller.

The independent variable determining total system power was taken to be the oxygen uptake rate (OUR). In an operating bioreactor, the submerged culture provides some OUR, which, at steady-state, is equal to an oxygen transfer rate (OTR); the product of a mass transfer coefficient, k_L, a mean bubble specific interfacial surface area, a, and an oxygen concentration driving force, (C^*-C_L):

$$\mathrm{OUR} = \mathrm{OTR} = k_L\mathrm{a}(\mathrm{C}^* - \mathrm{C_L}) \quad (1)$$

where k_La is usually lumped together and C^* and C_L are respectively the equilibrium and the actual dissolved oxygen concentrations (mmol/L). For reactor design purposes, several literature correlations are available to relate k_La to fundamental operating parameters. In STRs, the non-viscous mass transfer correlation of [13] is frequently used as a design equation. This correlation describes k_La as a function of bioreactor gassed power input (P) per unit volume (V) and the gas superficial velocity within the reactor, u_S:

$$k_L\mathrm{a}[\mathrm{s}^{-1}] = 0.002(P/V[\mathrm{W/m^3}])^{0.7}(u_S[\mathrm{m/s}])^{0.2} \quad (2)$$

where the pre-factor and exponents are adjustable for specific systems. This correlation is often relied on for its simplicity, as it does not depend on specific reactor geometry, or impeller speed, number, and type (though the accessible range of P/V is an implicit function of these [14]). The review of [15] lists other, more complex, correlations for k_La, but notes that the original Van't Riet correlation (Eq. (2)) is the most frequently used for basic design in non-viscous systems. Reactor sizes >100 m^3 are out of the fit space for the original correlation, but [16] developed a zoned model where the correlation was applied independently to stirred and unstirred zones within a larger reactor, and concluded that in the limit of good mixing ($1/k_La > t_{mix}$), the correlation could be simply applied to the entire volume. In any event, well proven correlations like Eq. (2) can be used to determine idealized scaled-up aerated bioreactor scenarios and make cost predictions.

For BCRs, [17] proposed a similar correlation, with k_La a primary function of the gas superficial velocity only. [13] further described how this correlation can be corrected for temperature, T, and effective broth viscosity, m_{eff}, resulting in:

$$k_L\mathrm{a}[\mathrm{s}^{-1}] = 0.32(u_S[\mathrm{m/s}])^{0.7}(m_{eff}[\mathrm{cP}])^{-0.84} \times 1.025^{(T[^\circ\mathrm{C}]-20)} \quad (3)$$