International Journal of Chemical and Process Engineering Research
2022 Vol. 9, No. 1, pp. 1-10.
ISSN(e): 2313-0776
ISSN(p): 2313-2558
DOI: 10.18488/65.v9i1.3123

EFFECTIVE COOLING OF CASSAVA STARCH TO ETHANOL BIO-REACTORS/FERMENTERS

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Article History
Received: 16 June 2022
Revised: 2 August 2022
Accepted: 19 August 2022
Published: 9 September 2022

Keywords
Cooling water economy
Shell-and-tube heat exchanger
Bio-ethanol production
Heat transfer
Mass transfer
Fermentation.

ABSTRACT

This paper presents an effective way to control the temperature of bio-reactors (fermenters) used in ethanol production and to reduce the volume of cooling water required per square meter of ethanol produced. This paper specifically focuses on the fermentation of cassava starch using *Saccharomyces cerevisiae* at 32°C. The flow across tube banks model is employed as the cooling mechanism of the bio-reactor. Cooling water at 28°C enters a shell containing five rows of fifteen (15) bio-reactors in a square in-line arrangement and exits the shell at 31.84°C. The total working volume of all fifteen (15) bio-reactors in the bank equals $180m^3$. Each bio-reactor in the bank is designed to have an effective heat transfer area to volume ratio of two (2) to enhance heat transfer. The total quantity of cooling water required per cubic meter of ethanol produced is found to be $97.833m^3$. A total amount of 1.303kW is required to power anchor impellers placed in each bio-reactor to provide mixing. The rotation speed of the impeller in each bio-reactor is $0.2rev \cdot s^{-1}$. A total of $3.453*10^{-6}$ W is required to move cooling water through the bio-reactor bank at a speed of $3.942*10^{-5}\ ms^{-1}$. An overall heat transfer coefficient of $11.345W \cdot m^{-2}°C^{-1}$ was found for the bio-reactor cooling system. Employing flow across tube banks model in cooling ethanol bio-reactors required significantly less amount of cooling water per cubic meter of ethanol produced compared to using internal cooling coils.

Contribution/Originality: This study shows how cooling water can be economized when flow across tubes in bank model is employed instead of using internal cooling coils as a cooling mechanism.

1. INTRODUCTION

This paper presents an effective way to cool industrial-scale bio-reactors (fermenters) to maintain optimum operating temperature and to reduce the volume of cooling water required per square meter of ethanol produced.

The yeast *Saccharomyces cerevisiae* remains the preferred organism for ethanol production due to its high ethanol, inhibitor, and osmo-tolerance in industrial processes, but it lacks starch degrading enzymes required for the efficient utilization of starch [1], therefore starch has to be first reduced to simple sugar before fermentation can occur. This process involves the conversion of cooked starch to maltodextrin using an α-amylase enzyme (Liquification Process) and the conversion of maltodextrin to glucose and fructose using glucoamylase enzyme (Saccharification Process) [2]. The simple sugar formed is then fermented anaerobically with yeast to produce ethanol and carbon dioxide (fermentation). Equation 1 describes the fermentation of glucose to produce ethanol.

$$C_6H_{12}O_6 + \text{Microorganisms} \xrightarrow{\text{yields}} 2C_2H_5\text{OH} + 2CO_2 + \text{Heat} \text{ [3]} \tag{1}$$

Fermentation of sugar with yeast generates heat as shown in Equation 1. The amount of heat produced during fermentation is dependent on factors including the type of microorganism used in the fermentation process [4]. The majority of ethanologens that are in industrial use belong to the mesophilic group (performed best in the range temperature 28-35°C) [5]. *Saccharomyces Cerevisiae* which is a mesophilic yeast is reported by Theerarattananoon, et al. [6] to generate heat of 35.09$kJmol^{-1}$ of ethanol produced during fermentation at 30°C with 300gl^{-1} initial glucose concentration. Thermophilic or thermotolerant yeasts unlike mesophilic yeast belong to a group of microorganisms that grow well at a temperature range of 41-122°C [7]. Thermophilic yeast Kluyveromyces sp. was found to generate heat of 67.84kJ.mol^{-1} of ethanol produced during fermentation at 50°C [4].

Biological or enzyme reactions are temperature sensitive, careful temperature control is, therefore, essential [8]. The shell and tube heat exchanger model (flow over tubes bank) is employed in removing excess heat from the bio-rector during fermentation. This design is based on the fermentation of cassava starch using *Saccharomyces cerevisiae*.

2. METHOD

Flow across tube banks is a model of the shell-and-tube exchanger where heat exchange occurs between a fluid flowing through a bank of tubes and another fluid moving through a tube in a perpendicular direction. The tubes are usually placed in a shell, especially when the fluid flowing through the bank is a liquid. Fluid flows through the space between the tubes and the shell for heat exchange to occur. The tubes are either arranged in-line or in a staggered manner [9].

Excess heat generated within the bio-reactor is removed by water flowing through reactor banks in a square in-line arrangement.

3. THEORY/ CALCULATIONS

3.1. Thermal Properties of Cassava Starch Solution

The thermal properties and density of cassava starch solutions studied as a function of temperature (30-50°C) and concentration (20-50%w/w) is reported by Cansee, et al. [10] and presented in Table 1.

Table 1. Thermal properties of Cassava starch [10].

Property of Cassava starch	Value
Density(ρ), $K_g \cdot m^{-3}$	1044 - 1120
Specific Heat Capacity (C_P), $kJ \cdot K_g{}^{-1}$°C^{-1}	3.354 - 4.004
Thermal Conductivity(k), $W \cdot m^{-1}$°C^{-1}	0.307- 0.333

The intrinsic viscosity of Cassava Starch that has undergone hydrolysis is reported by Rocha, et al. [11] as 2.20. Equation 2 is used to determine the dynamic viscosity from intrinsic viscosity.

$$[\eta] = \lim_{c \to 0} \frac{\eta - \eta_o}{\eta_o \cdot c} \text{ [12]} \tag{2}$$

Where $[\eta]$ is intrinsic viscosity, η_o is solvent viscosity, c is mass concentration and η is the viscosity of the solution. Using water as the solvent the dynamic viscosity of a 30% w/v hydrolyzed cassava starch is 59.8cP.

Table 2. Dimensions and properties of proposed fermenter/bio-reactor

Parameter	Value
The material of bioreactor construction	Type 314 Stainless Steel
Thermal conductivity type 316 stainless steel at 31°C	$14.34 Wm^{-1}°C^{-1}$ [13]
Heat production rate (*Saccharomyces cerevisiae*)	$35.09 kJ \cdot mol^{-1}$ of ethanol produced [4]
Shape of bioreactor	Cylindrical
Type of reactor	Batch reactor
Reactor residence time	66hr [14]
Reactor cooling mechanism	Flow across tube banks
Total working volume of reactors in bank	$180m^3$
Working volume of each reactor in bank	$12m^3$
Number of reactors in bank	15
Inside radius of each reactor in bank	1m
The outside radius of each reactor in the bank	1.02m
Height of each reactor in bank	4.5m
Reactor impeller type	Anchor Impeller
Length, width, and height of the shell	23.4m, 14.5m, 5m

Table 2 above contains dimension and physical properties of proposed new fermenter or bio-reactor to enable effective heat transfer.

3.2. Rate of Heat Generation Within Bio-Reactor

Choi et al reported an ethanol yield of $2.1 g \cdot l^{-1} \cdot hr^{-1}$ ($45.652 mol \cdot m^{-3} \cdot hr^{-1}$) with *Saccharomyces Cerevisiae* at a fermentation temperature of 32°C. [14] [15]

3.2.1. Heat Generation in Each Reactor (E_G)

$$E_G = 35.09 \text{kJ}. mol^{-1} * 45.652 mol \cdot m^{-3} \cdot hr^{-1} = 0.445\, kWm^{-3} \quad (3)$$

3.3. Temperature Profile Across the Bio-Reactor

$$q_{flux} = \frac{(T_C - T_S)}{\frac{1}{2\pi R_1 Lh} + \frac{In(\frac{R_2}{R_1})}{2\pi Lk}} = \frac{E_G R_1}{2} \text{ [9]} \quad (4)$$

Where q_{flux} is heat flux, Wm^{-2}; h is convective heat transfer coefficient, $Wm^{-2}°C^{-1}$W/m²· °C; T_s is the temperature at the bio-reactor outside surface, °C; T_c is the temperature at the core or center of the bio-reactor, °C; k is the thermal conductivity of reactor walls, L is the height of the reactor meters; R_1 and R_2 are the inside and outside radius of the bio-reactor respectively in meters. A function of Reynolds number (Re), Prandtl's number (Pr), and friction factor(f) correlates Nusselt number(Nu) to convective heat transfer coefficient(h). A modified form of the petukhov equation by Gnielinski [16] for determining the Nusselt number for a forced turbulent flow in a tube shown in Equation 5;

$$Nu = \frac{(\frac{f}{8})(Re - 1000) \cdot Pr}{1 + 12.7 \left(\frac{f}{8}\right)^{0.5} (Pr^{\frac{2}{3}} - 1)} = \frac{hD_1}{k_{hs}} \text{ [17]} \quad (5)$$

Equation 5 above is applicable when Re is greater than 3000 and less than $5*10^6$ and Pr is equal or greater than 0.5 and equal or less than 2000. Where k_{hs} is the thermal conductivity of the bio-reactor content.

3.3.1. Calculation of Reynolds Number, Prandtl's Number, and Friction Factor

Impeller Reynolds(Re_i) number is calculated due to mixing occurring in the reactor

$$Re_i = \frac{rotational\ speed(N) \cdot Density(\rho) \cdot squared\ impeller\ Diameter(D_i^2)}{Dynamic\ Viscosity(\mu)} \text{ [17]} \quad (6)$$

The standard impeller diameter of an Anchor impeller is 90% of the reactor inside diameter [17]

Substituting values into Equation 6;

$$Re_i = \frac{0.2rev \cdot s^{-1} \cdot 1120K_g \cdot m^{-3} \cdot (0.9 \cdot 2)^2 m}{59.8*10^{-3} Kg \cdot m^{-1} \cdot s^{-1}} = 12136.455 \quad (7)$$

$$\text{Prandtl's Number (Pr)} = \frac{specific\ heat\ capacity(C_p) \cdot Dynamic\ Viscosity(\mu)}{Thermal\ Conductivity(k)} \text{ [9]} \quad (8)$$

Substituting values into Equation 8;

$$\text{Prandtl's Number (Pr)} = \frac{3.354*10^3 J \cdot K_g{}^{-1} {}^\circ C^{-1} \cdot 59.8*10^{-3} Kg \cdot m^{-1} \cdot s^{-1}}{0.307 W \cdot m^{-1} {}^\circ C^{-1}} = 653.32 \quad (9)$$

Assuming a smooth reactor walls friction factor(f) in turbulent flow can be determined from the explicit *first Petukhov equation* given as;

$$\text{Friction Factor(f)} = (0.790 \text{In} Re_i - 1.64)^{-2} \text{ [9]} \quad (10)$$

Substituting values into Equation 10:

$$\text{Friction Factor (f)} = (0.790\text{In}(12136.455) - 1.64)^{-2} = 0.0298 \quad (11)$$

From Equation 11 above friction factor is 0.0298

Substituting values in Equation 5;

$$Nu = \frac{(\frac{0.0298}{8})(12136.455 - 1000)*653.32}{1+12.7(\frac{0.0298}{8})^{0.5}(653.32^{\frac{2}{3}} - 1)} = \frac{h \cdot 2}{0.307} \quad (12)$$

$$Nu = 462.92 = 6.515h \quad (13)$$

Convective heat transfer coefficient (h) = 71.058 $W \cdot m^{-2} {}^\circ C^{-1}$ (14)

3.3.2. Heat Flux and Temperature of the Outside Surface of the Reactor

Substituting values into Equation 4 above and setting the core or center of the reactor at 32 °C

$$q_{flux} = \frac{(32 - T_S)}{\frac{1}{2\pi \cdot 1 \cdot 4.5 \cdot 71.058} + \frac{In(\frac{1.02}{1})}{2\pi \cdot 4.5(14.34)}} = \frac{0.445*10^3 \cdot 1}{2} \quad (15)$$

$$q_{flux} = \frac{(32 - T_S)}{5.4631*10^{-4}} = 222.5 W m^{-2} \quad (16)$$

$$T_s = 31.88 {}^\circ C \quad (17)$$

$$Heat\ production\ rate(\dot{Q}) = Heat\ flux(q_{flux}) \cdot Total\ Heat\ Transfer\ Area(A_s) \text{ [9]} \quad (18)$$

$$\dot{Q} = 222.5 W m^{-2} \cdot 2\pi \cdot 1m \cdot 4.5m * 15 = 94365.59 W \quad (19)$$

3.4. Mass Flowrate of Cooling Water

$$q_{flux} = hA_s(T_s - T_e) \text{ [9]} \quad (20)$$

Where T_e the exit temperature of the cooling is water from the shell and T_s is the external surface temperature of the reactor. A_s is total effective heat transfer area (curved surface area of the bio-reactor), h is the heat transfer coefficient.

$$A_s = 2\pi \cdot 1 \cdot 4.5 \cdot 15 = 424.115 m^2 \quad (21)$$

$$Reynolds\ Number(R_e) = \frac{V_{max} \cdot \rho \cdot D_C}{\mu} \quad (22)$$

Where ρ is density, D_C is outer diameter, μ is dynamic viscosity, V_{max} expresses in Equation 23.

$$V_{max} = \frac{S}{S - D_c} V \quad (23)$$

For square in-line tubes in a bank, V is the flow velocity of cooling water through the reactor bank. S is the distance between the center of any two tubes in a column or row. Setting S at 2.2m and V at $3.942 * 10^{-5}\ ms^{-1}$ Substituting values into Equation 23,

$$V_{max} = \frac{2.2}{2.2 - 2.04} * 3.942 * 10^{-5}\ ms^{-1} = 5.421 * 10^{-4} ms^{-1} \quad (24)$$

All properties of cooling water except unless specified are evaluated at the arithmetic mean temperature (T_m) of the fluid determined from;

$$T_m = \frac{T_i + T_e}{2} \quad (25)$$

Where T_i and T_e are the fluid temperatures at the inlet and the exit respectively of cooling water flowing through the shell. The fluid temperature at the inlet is set at 28°C and the exit temperature is assumed at 31°C.

Substituting values into Equation 25,

$$T_{m1} = \frac{28 + 31}{2} = 29.5 \approx 30°\mathrm{C} \quad (26)$$

Table 3. Properties of water at arithmetic mean temperature [18].

Property of Water	$T_{m1}(30°C)$
Density(ρ), $K_g \cdot m^{-3}$	995.7
Specific Heat Capacity(C_P), $kJ \cdot K_g{}^{-1}°\mathrm{C}^{-1}$	4.183
Thermal Conductivity(k), $W \cdot m^{-1}°\mathrm{C}^{-1}$	0.602
Prandtl's Number(Pr)	5.535
Dynamic Viscosity(μ), $10^{-6}\mathrm{Pa} \cdot \mathrm{s}$	797.7

Table 3 above presents properties of water at arithmetic mean temperature

Substituting values into Equation 22,

$$Reynolds\ Number(R_e) = \frac{(5.421*10^{-4})(995.7)(2.04)}{797.7*10^{-6}} = 1380.38 \quad (27)$$

Using the appropriate Zukauskas equation based on the Reynolds Number obtained;

$$Nu_{D,Nl} = 0.27Re^{0.63}Pr^{0.36}(\frac{Pr}{Pr_s})^{0.25} = \frac{hD_c}{k} \text{ [9]} \quad (28)$$

Where $Nu_{D,Nl}$ is Nusselt number for a bank with sixteen (16) or more rows, Pr_s is Prandtl's number of water at (31.88°C) the temperature of the outer surface of the reactor, Pr is Prandtl's number of water at arithmetic mean temperature, h is convective heat transfer coefficient, Re is Raynolds Number, D_c the outside diameter of the reactor, k is the thermal conductivity of water at arithmetic mean temperature.

$$Nu_D = FNu_{D,Nl} \text{ [9]} \quad (29)$$

Where $Nu_D$is corrected Nusselt number for a bank with less than sixteen(16) rows, F is a correction factor for the number of rows less than 16, for five (5) rows of square in-line tubes in a bank, F equals 0.93. [9]

Substituting value into Equation 28,

$$Nu_{D,Nl} = 0.27(1380.38)^{0.63}(5.535^{0.36})(\frac{5.535}{5.307})^{0.25} = 48.045 \quad (30)$$

Substituting value into Equation 29,

$$Nu_D = 48.045 * 0.93 \quad (31)$$

$$Nu_D = 44.682 = \frac{hD_c}{k} \quad (32)$$

$$Nu = 44.682 = \frac{2.04h}{0.602} \quad (33)$$

$$h = 13.186W \cdot m^{-2}°\mathrm{C}^{-1} \quad (34)$$

Therefore the average heat transfer coefficient and Nusselt number of cooling water flowing through the bank is $13.186W \cdot m^{-2}°\mathrm{C}^{-1}$ and 44.682 respectively.

3.4.1 Actual Exit Temperature, T_e of Cooling Water from the Shell

Substituting values into Equation 20,

$$(T_s - T_e) = \frac{q_{flux}}{hA_s} = \frac{222.5Wm^{-2}}{13.186 * 424.115} \quad (35)$$

$$(31.88 - T_e) = \frac{222.5Wm^{-2}}{13.186 * 424.115} \quad (36)$$

$$T_e = 31.84°C \quad (37)$$

$Rate\ of\ heat\ production(\dot{Q}) = Heat\ flux(q_{flux}) \cdot Total\ Heat\ Transfer\ Area(A_s)$ [9] (38) Substituting values into Equation 38,

$$Rate\ of\ heat\ production(\dot{Q}) = 222.5Wm^{-2} * 424.115m^2 \quad (39)$$

$$Rate\ of\ heat\ production(\dot{Q}) = 94.366kW \quad (40)$$

$$\dot{Q} = \dot{m}C_p(T_e - T_i) \text{ [9]} \quad (41)$$

Where T_i and T_e is the cooling water temperatures at the inlet and the exit of the tube bank, respectively. $\dot{m}$ is the mass flow rate of cooling water and C_p is the specific heat capacity of the water at arithmetic mean temperature (T_m).

Substituting values into Equation 41,

$$94.366kW = \dot{m} \cdot 4183J \cdot K_g{}^{-1}°C^{-1}(31.84°C - 28°C) \quad (42)$$

$$\dot{m} = 5.875K_g \cdot s^{-1} = 21149.4K_g \cdot hr^{-1} \quad (43)$$

Average ethanol production rate $= 180m^3 \cdot 45.652mol \cdot m^{-3} \cdot hr^{-1} = 8217.36\ mol \cdot hr^{-1}$ (44)

The total volume of ethanol produced after 66 hours of fermentation

$$\frac{66hr \cdot 8.21736kmol \cdot hr^{-1} \cdot 46.07k_g kmol^{-1}}{789k_g m^{-3}} = 31.67m^3 \quad (45)$$

The volume of cooling water needed to fill the shell before the bio-reactor operation

The volume of shell = volume of cooling water = $1696.5m^3$ (46)

The volume of cooling water needed during operation of bio-reactor

$$\frac{21149.4K_g \cdot hr^{-1} \cdot 66hr}{995.7k_g m^{-3}} = 1401.86m^3 \quad (47)$$

The volume of cooling water (V_{CE}) per cubic meter of ethanol produced

$$V_{CE} = \frac{Total\ Volume\ of\ cooling\ water}{Total\ Volume\ Ethanol\ Produced} \quad (48)$$

Substituting values into Equation 48,

$$V_{CE} = \frac{1696.5m^3 + 1401.86m^3}{31.67m^3} = 97.833m^3 \quad (49)$$

3.4.2. Pressure Drop

Pressure drop(ΔP) of cooling water flowing across the bank reactors is expressed as;

$$\Delta P = N_L \cdot f \cdot x \frac{\rho \cdot V_{max}^2}{2} \text{ [9]} \quad (50)$$

Where; N_L is the number of rows, f is friction factor, x is a correction factor, V_{max} maximum velocity as calculated in Equation 24 above, x is equal to 1 for a square arrangement of tubes in a bank [9]. Friction factor(f), a function of Reynolds number and the ratio between the distances between any two tubes in a column to the outer diameter of the tube. The value of f is read from a graph [9].

Substituting value into Equation 50,

$$\Delta P = 5 \cdot 0.8 \cdot 1 \cdot \frac{995.7 \cdot (5.421 * 10^{-4})^2}{2} = 5.852 * 10^{-4}Pa \quad (51)$$

3.4.3. Power Required to Move Cooling Water Through the Reactor Bank

$$\dot{W}_{pump} = \frac{\dot{m}\Delta P}{\rho} \quad [9] \tag{52}$$

Where, $\dot{W}_{pump}$ is pumping power, $\dot{m}$ is the mass flow rate, ΔP is the pressure drop and ρ is density.

Substituting value into Equation 52,

$$\dot{W}_{pump} = \frac{5.875 \cdot 5.8522 * 10^{-4}}{995.7} = 3.453 * 10^{-6} W \tag{53}$$

3.5 Agitation Power Consumption

Agitation power consumption estimates the theoretical power needed to stir the content within the bioreactors. Power Number (N_p) for a gate impeller is defined below.

$$N_p = 215({Re_i}^{-0.955}) \quad [17] \tag{54}$$

Substituting values into Equation 54,

$$N_p = 215(12136.455)^{-0.955} = 0.027 \tag{55}$$

$$\text{Power consumed per Blade, (P)} = N_p \cdot \rho \cdot N^3 \cdot D_i^5 \quad [17] \tag{56}$$

Where $N_p$is power number of the impeller, ρ is density, N is impeller rotation speed, D_i is impeller diameter.

Substituting values into Equation 56

$$\text{Power (P)} = (0.027) \cdot (1120 K_g \cdot m^{-3}) \cdot (0.2 rev \cdot s^{-1})^3 \cdot (0.9 \cdot 2)^5 m \tag{57}$$

$$\text{Power (P)} = 4.571\text{W} \tag{58}$$

For a standard Anchor impeller, the width of the blade equals 10% of reactor Diameter [17] [19]

$$\text{Number of required Blades} = \frac{working\ Hieght\ of\ reactor}{width\ of\ impeller\ blade} \quad [17] \tag{59}$$

$$\text{Number of required Blades} = \frac{3.82m}{0.1 \cdot 2} = 19.1 \tag{60}$$

The number of required blades on a shaft is approximated at 19

$$\text{Total stirring power per reactor(P)} = 19 \cdot 4.571\ \text{W} = 86.846\text{W} \tag{61}$$

$$\text{Total Agitation Power for all 15 bio-reactors in bank} = 1.303\text{kW} \tag{62}$$

3.6. Overall Heat Transfer Coefficient (U)

$$U = \frac{1}{\frac{1}{h_1} + \frac{R_1}{k_A}\ln\left(\frac{R_2}{R_1}\right) + \left(\frac{R_1}{R_2}\right)\frac{1}{h_2}} \quad [9] \tag{63}$$

Where h_1 is h_2 are convective heat transfer coefficients within the bio-reactor and the cooling water respectively. R_1 and R_2 are inside and outside radius of bio-reactor. k_A is the thermal conductivity of bio-reactor walls.

Substituting values into Equation 63;

$$U = \frac{1}{\frac{1}{71.058} + \frac{1}{14.34}\ln\left(\frac{1.02}{1}\right) + \left(\frac{1}{1.02}\right)\frac{1}{13.185}} = 11.345 W \cdot m^{-2} {}^{\circ}C^{-1} \tag{64}$$

4. RESULTS

Table 4 presents a summary of mechanical and chemical engineering design parameters of newly designed fermenter/bio-reactor.

Table 4. Summary of chemical/mechanical engineering design parameters

Parameter	Value
Bio-reactor	
Shape of bioreactor	Cylindrical
Material of construction	Stainless Steel
Inside Radius	1m
Outside Radius	1.02m
Height	4.5m
Type of bio-reactor	Batch reactor
Operation temperature	32°C
Working volume of each reactor in bank	$12m^3$
Total working volume of reactors in bank	$180m^3$
Cooling mechanism	Flow across tube banks
Impeller type	Anchor impeller
Impeller rotation speed	$0.2 rev \cdot s^{-1}$
Total bio-reactor stirring power requirement	1.303kW
Convective heat transfer coefficient within bio-reactor	$71.058 W \cdot m^{-2} °C^{-1}$
Overall Heat Transfer Coefficient	$11.345\ W \cdot m^{-2} °C^{-1}$
Shell	
Volume of shell	$1696.5m^3$
Number of bio-reactors in bank	15
Number of rows in the bank	5
Arrangement of bio-reactors in bank	Square in-line
Spacing between bio-reactors in bank	2.2m
Cooling water	
The volume of cooling water per cubic meter of ethanol produced	$97.833m^3$
Power required to move cooling water through the shell	$3.453*10^{-6}W$
Inlet temperature	28°C
Exit temperature	31.84°C
Velocity through shell	$3.942* 10^{-5}\ ms^{-1}$
Convective heat transfer coefficient	$13.185\ W \cdot m^{-2} °C^{-1}$

5. DISCUSSIONS

Khatiwada and Silveira in a case study of an Indian distillery reported that, fermentation of $400m^3$ wort using *S. cerevisiae* completes in 32h with 8% (v/v) ethanol concentration, which requires $550m^3$cooling water per cubic meter of ethanol produced to maintain the bio-reactor temperature at 32°C using internal cooling coils as a heat exchange medium [20] $97.833m^3$ cooling water per cubic meter of ethanol produced presented by this paper is five times lower and can significantly reduce the unit cost of ethanol production.

A benefit of using a flow across tube banks to using internal cooling coils to control the temperature of ethanol bio-reactors is a lower risk of fouling. When internal cooling coils are used in ethanol bio-reactor typically in the fermentation of viscous fluids such as cassava starch, the content of the reactor can easily coat the outside surface of the cooling coils thereby decreasing effective heat transfer. In this design, the whole curved surface area of the cylindrical bio-reactor is used as a heat transfer area given an effective heat transfer surface area to volume ratio of 2. A sudden temperature change within the bio-reactor system is unlikely due to the large volume of water ($1696.5m^3$) flowing through the shell. This is because an amount of about $7 * 10^6 J$ of energy is required to cause a one-degree temperature change in the mass of cooling water in the shell of the bio-reactor system compared to a total heat generation rate of $94.366kW$. This design assumes no heat is lost to the environment from the bio-reactor but all heat is transferred to cooling water flowing through the bio-reactor bank.

The Anchor impeller type used is designed to sweep through the whole volume of the bio-reactor to enhance effective heat transfer [19].

The close temperature difference between the cooling water inlet and outlet of 3.84°C saves energy required in the cooling tower to restore the cooling water to 28°C for reuse to cool the bio-reactor.

One major disadvantage of this design is, it required high initial capital investment to build many small bio-reactors instead of building a single large bio-reactor.

6. CONCLUSION

Employing flow across tube banks model in cooling ethanol bio-reactors can save the amount of cooling water required to produce a cubic meter of ethanol by about a fifth compared to using internal cooling coils.

Funding: This study received no specific financial support.
Competing Interests: The authors declare that they have no competing interests.
Authors' Contributions: All authors contributed equally to the conception and design of the study.

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