



SEMESTER 7
C15 APPLIED BIOTECHNOLOGY
Paper Code (BBT 7001 [A])
(Credits: Theory-4, Practicals-2)
THEORY Lectures: 20

Course Objective(s)

- To study microbial growth and its maintenance.
- To develop understanding different bioreactors.
- To provide knowledge biochemistry of different industrial chemicals

Course Outcome(s):

- Learn basics of fermentation technology, bioreactors, production optimization, and downstream processing.
- On successful completion of the course the students should have understood the basics of fermentation technology and learnt the concept of different metabolite production by microbes in industrial setup.

THEORY

Unit-I Microbial Cell Growth and Death Kinetics: (10 Lectures)

Screening and improvement of industrially important microorganisms, Microbial Growth and Death Kinetics, Media for Industrial Fermentation, Air and Media Sterilization. IKS: Contribution of Indian Scientists.

Unit-II Operation and Control of Bioreactors: (10 Lectures)

Types of Fermentation Processes: Analysis of batch, fed-batch and continuous bioreactors, stability of microbial bioreactors, analysis of mixed populations, specialized bioreactors-pulsed,

fluidized, photo bioreactors, etc. Measurement and Control of bioprocess parameters. Downstream processing, Whole cell immobilization and their industrial applications.

Unit-III Fermentation Technology: (10 Lectures)

Industrial production of chemicals: Ethanol, Acids (citric, acetic and gluconic acid), Solvents (glycerol, acetone, butanol), Antibiotics (penicillin, streptomycin, tetracyclin), Semi-synthetic antibiotics, Amino acids (lysine, glutamic acid), Single cell protein.

Unit-IV Applications of Bioprocess Engineering: (10 Lectures)

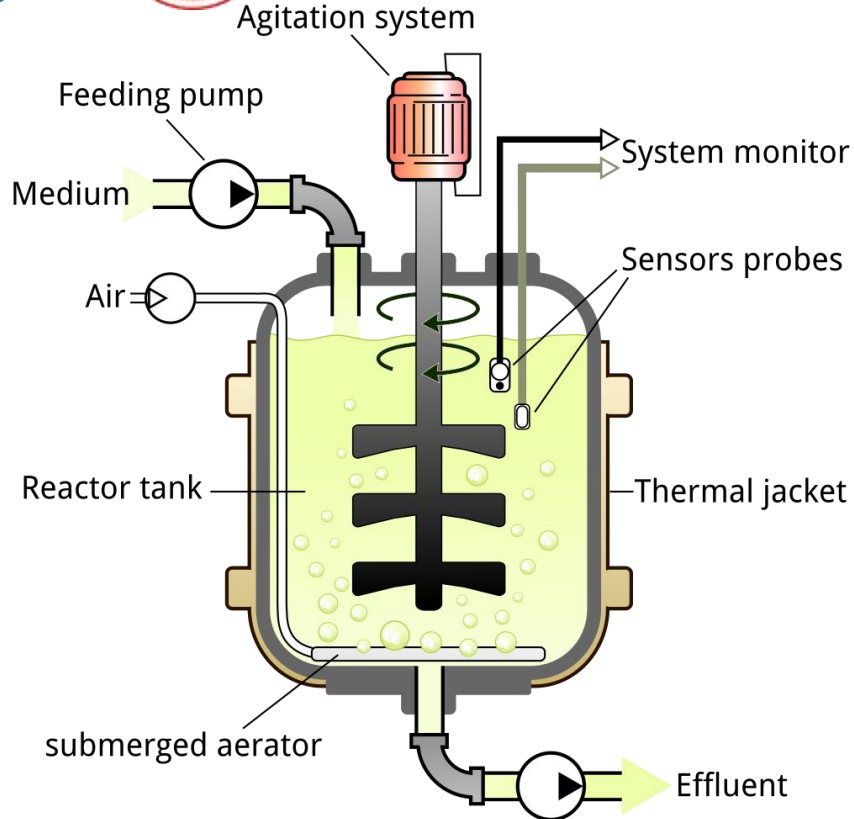
Agitation and aeration: requirement in industrial processes, concept of volumetric oxygen transfer coefficient and its determination (K_La), Factors affecting K_La values; Uses of microbes in mineral beneficiation and oil recovery. Introduction to food technology; Elementary idea of canning and packaging, Sterilization and pasteurization of food products.

PRACTICALS

1. To plot Microbial growth curve for shake flask culturing using turbidity method.
2. Prepare a standard curve of reducing sugar by 3,5-Dinitrosalicylic acid method
3. To produce invertase enzyme and find its activity from Baker's Yeast
4. Preparation of standard curve of Ethanol
5. Quantitative estimation of ethanol produced during Yeast fermentation
6. Production of Penicillin and assaying its activity.
7. To get familiarized with the lab scale fermenter (bench top fermenter)
8. To determine dissolved oxygen concentration in tap and aerated water.
9. To determine the volumetric transfer coefficient (K_La)
10. Estimation of BOD in a given waste water sample.
11. Centrifugation studies during settling of yeast cells.
12. Yeast cell disruption by mechanical methods.

SUGGESTED BOOKS

1. Bioprocess Engineering, Shuler M & Kargi F, Prentice Hall
2. Biochemical Engineering Fundamentals, Bailey JE & Ollis DF
3. Bioprocess Engineering Principles, Doran, PM, Academic Press, California



A **bioreactor** is any manufactured device or system that supports a biologically active environment. In one case, a bioreactor is a vessel in which a chemical process is carried out which involves organisms or biochemically active substances derived from such organisms. This process can either be aerobic or anaerobic.

These bioreactors are commonly cylindrical, ranging in size from litres to cubic metres, and are often made of stainless steel. It may also refer to a device or system designed to grow cells or tissues in the context of cell culture. These devices are being developed for use in tissue engineering or biochemical/bioprocess engineering.



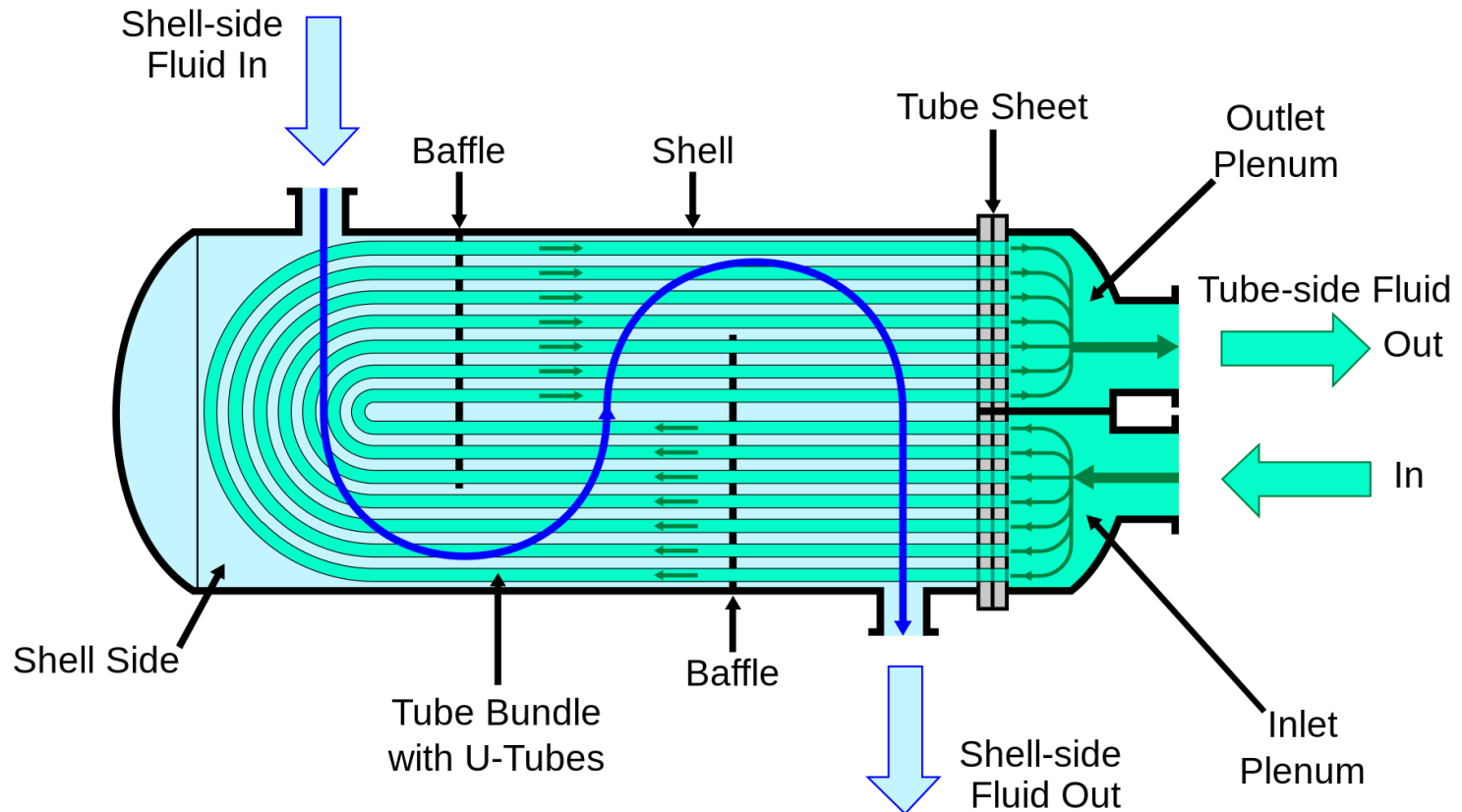
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U-Tube Heat Exchanger



Baffles are flow-directing or obstructing vanes or panels used to direct a flow of liquid or gas. It is used in some household stoves and in some industrial process vessels (tanks), such as shell and tube heat exchangers, chemical reactors, and static mixers.



Vessel or chamber

The vessel is the main container where cell culture or microbial fermentation takes place. It is typically made of stainless steel or glass and is designed to be sterile and airtight to prevent contamination and maintain controlled conditions.

Agitator or stirrer

The agitator is a mechanical component that mixes the contents of the vessel. It ensures uniform distribution of nutrients, oxygen, and temperature throughout the culture. Common types include impellers and paddles.

Heat exchanger

A heat exchanger is used to control the temperature inside the bioreactor. It can heat or cool the culture medium as needed to maintain the desired temperature for cell growth or fermentation.

pH control system

This system measures the pH level of the culture and adds acid or base as necessary to maintain the pH within a specified range. Consistent pH control is crucial for cell viability and product yield.

Aeration and gas supply

Bioreactors provide oxygen to the culture through aeration or sparging. Aeration systems can include spargers, diffusers, or bubblers to deliver oxygen and remove carbon dioxide produced during metabolism.

Nutrient inlet-outlet

Nutrients, such as sugars, salts, vitamins, and trace elements, are added to the bioreactor through an inlet. Waste products are removed through an outlet to maintain a balanced culture medium.

Sensors and analytical instruments

Various sensors are integrated into the bioreactor to monitor parameters like temperature, pH, dissolved oxygen, and biomass concentration. These sensors provide real-time data for process control and optimization.



Sampling ports

Sampling ports allow researchers to collect samples of the culture for analysis without disturbing the entire process. Sampling is essential for assessing cell density, product concentration, and other parameters.

Control system

The control system manages and regulates the bioreactor's operating conditions based on input from sensors. It controls temperature, pH, oxygen levels, agitation speed, and nutrient addition to maintain optimal culture conditions.

Sterilization system

To ensure aseptic conditions, bioreactors have a sterilization system that can use methods such as autoclaving, chemical sterilization, or in-place sterilization of components.

Sparger or air inlet

The sparger is a component that introduces air or oxygen into the culture medium. It may be a porous tube or diffuser that creates small bubbles for efficient oxygen transfer to the cells.

Foam control system

Agitation and aeration can produce foam on the surface of the culture. Foam control systems prevent excessive foam buildup, which can interfere with culture conditions and lead to contamination.



Classification of reactors

☐ Reactors are classified as follows:

1. **Mode of operation (e.g. batch, fed-batch, continuous)**
2. **Geometric configuration (e.g. tubular, agitated tank)**
3. **Contacting patterns between phases (e.g. packed bed, expanded bed, fluidized bed, trickle bed reactor, bubble column reactor, airlift reactor)**

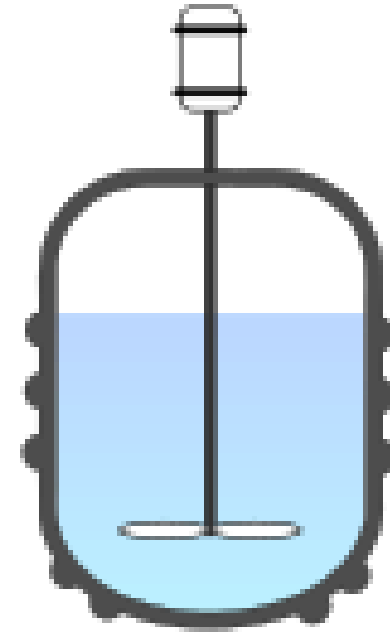


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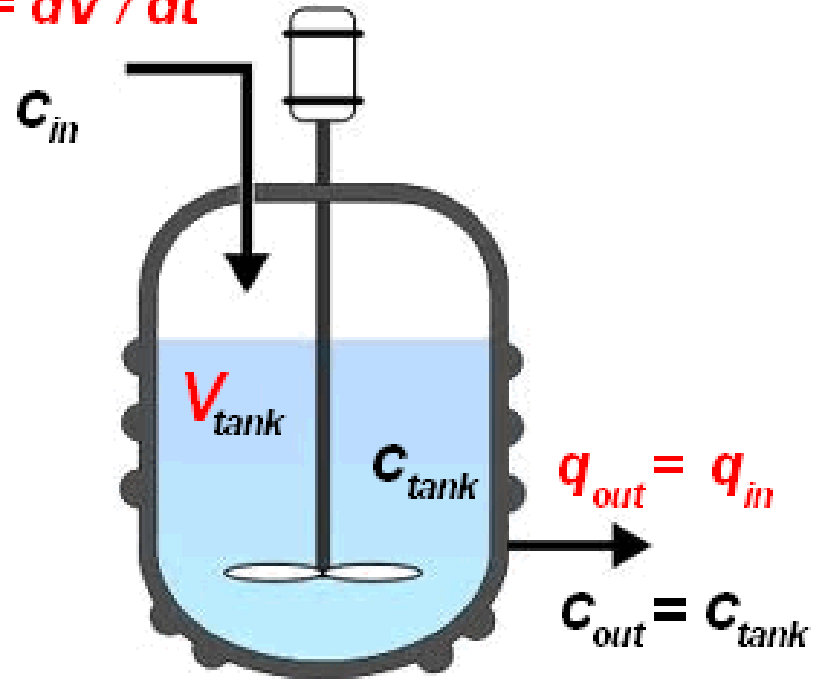


A **batch reactor** is a chemical reactor in which a non-continuous reaction is conducted, i.e., one where the reactants, products and solvent do not flow in or out of the vessel during the reaction until the target reaction conversion is achieved. By extension, the expression is somehow inappropriately used for other batch fluid processing operations that do not involve a chemical reaction, such as solids dissolution, product mixing, batch distillation, crystallization, and liquid/liquid extraction. In such cases, however, they may not be referred to as reactors but rather with a term specific to the function they perform



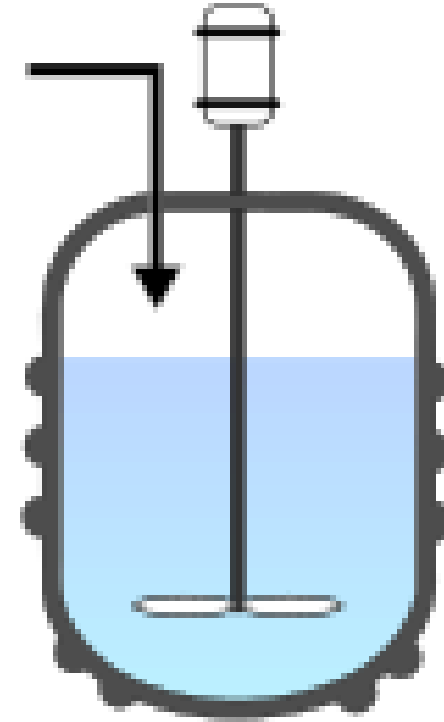
Continuous reactors (alternatively referred to as **flow reactors**) carry material as a flowing stream. Reactants are continuously fed into the reactor and emerge as continuous stream of product. Continuous reactors are used for a wide variety of chemical and biological processes within the food, chemical and pharmaceutical industries. A survey of the continuous reactor market will throw up a daunting variety of shapes and types of machine. Beneath this variation however lies a relatively small number of key design features which determine the capabilities of the reactor. When classifying continuous reactors, it can be more helpful to look at these design features rather than the whole system.

$$q_{in} = dV / dt$$





Fed-batch culture is, in the broadest sense, defined as an operational technique in biotechnological processes where one or more nutrients (substrates) are fed (supplied) to the bioreactor during cultivation and in which the product(s) remain in the bioreactor until the end of the run. An alternative description of the method is that of a culture in which "a base medium supports initial cell culture and a feed medium is added to prevent nutrient depletion". It is also a type of **semi-batch culture**. In some cases, all the nutrients are fed into the bioreactor. The advantage of the fed-batch culture is that one can control concentration of fed-substrate in the culture liquid at arbitrarily desired levels (in many cases, at low levels).





FED-BATCH CULTURE

Yoshida, Yamane, and Nakamoto (1973) introduced the term fed-batch culture to describe batch cultures which are fed continuously, or sequentially, with medium, without the removal of culture fluid. A fed-batch culture is established initially in batch mode and is then fed according to one of the following feed strategies:

1. The same medium used to establish the batch culture is added, resulting in an increase in volume.
2. A solution of the limiting substrate at the same concentration as that in the initial medium is added, resulting in an increase in volume.
3. A concentrated solution of the limiting substrate is added at a rate less than in (1) and (2), resulting in an increase in volume.
4. A very concentrated solution of the limiting substrate is added at a rate less than in (1), (2), and (3), resulting in an insignificant increase in volume.

Fed-batch systems employing strategies (1) and (2) are described as variable volume, whereas a system employing strategy (4) is described as fixed volume. The use of strategy (3) gives a culture intermediate between the two extremes of variable and fixed volume.



VARIABLE VOLUME FED-BATCH CULTURE

The kinetics of variable volume fed-batch culture have been developed by **Dunn and Mor (1975)** and **Pirt (1974, 1975, 1979)**. The following account is based on that of **Pirt (1975)**.

Consider a batch culture in which growth is limited by the concentration of one substrate; the biomass at any point in time will be described by the equation:

$$x_t = x_0 + Y(S_R - s) \quad (2.31)$$

where x_t is the biomass concentration after time, t hours, and x_0 is the inoculum concentration

The final biomass concentration produced when $s = 0$ may be described as x_{\max} and, provided that x_0 is small compared with x_{\max} :

$$x_{\max} \simeq Y \cdot S_R \quad (2.32)$$



If, at the time when $x = x_{\max}$, a medium feed is started such that the dilution rate is less than μ_{\max} , virtually all the substrate will be consumed as fast as it enters the culture, thus:

$$FS_R \simeq \mu \left(\frac{X}{Y} \right) \quad (2.33)$$

where F is the flow rate of the medium feed, and X is the total biomass in the culture, described by $X = xV$, where V is the volume of the culture medium in the vessel at time t .

From Eq. (2.33) it may be concluded that input of substrate is equalled by consumption of substrate by the cells. Thus, $(ds/dt) \simeq 0$. Although the total biomass in the culture (X) increases with time, cell concentration (x) remains virtually constant, that is, $(dx/dt) = 0$ and therefore $\mu = D$. This situation is termed a quasi steady state. As time progresses, the dilution rate will decrease as the volume increases and D will be given the expression:

$$D = \frac{F}{(V_0 + Ft)} \quad (2.34)$$



where V_0 is the original volume. Thus, according to Monod kinetics, residual substrate should decrease as D decreases resulting in an increase in the cell concentration.

However, over most of the range of μ that will operate in fed-batch culture, S_R will be much larger than K_s so that, for all practical purposes, the change in residual substrate concentration would be extremely small and may be considered as zero.

Thus, provided that D is less than μ_{max} and K_s is much smaller than S_R a quasi steady state may be achieved. The quasi steady state is illustrated in [Fig. 2.15a](#).

The major difference between the steady state of a chemostat and the quasi steady state of a fed-batch culture is that μ is constant in the chemostat but decreases in the fed-batch.

[Pirt \(1979\)](#) has expressed the change in product concentration in variable volume fed-batch culture in the same way as for continuous culture:

$$dp/dt = q_p x - D_p.$$

Thus, product concentration changes according to the balance between production rate and dilution by the feed.

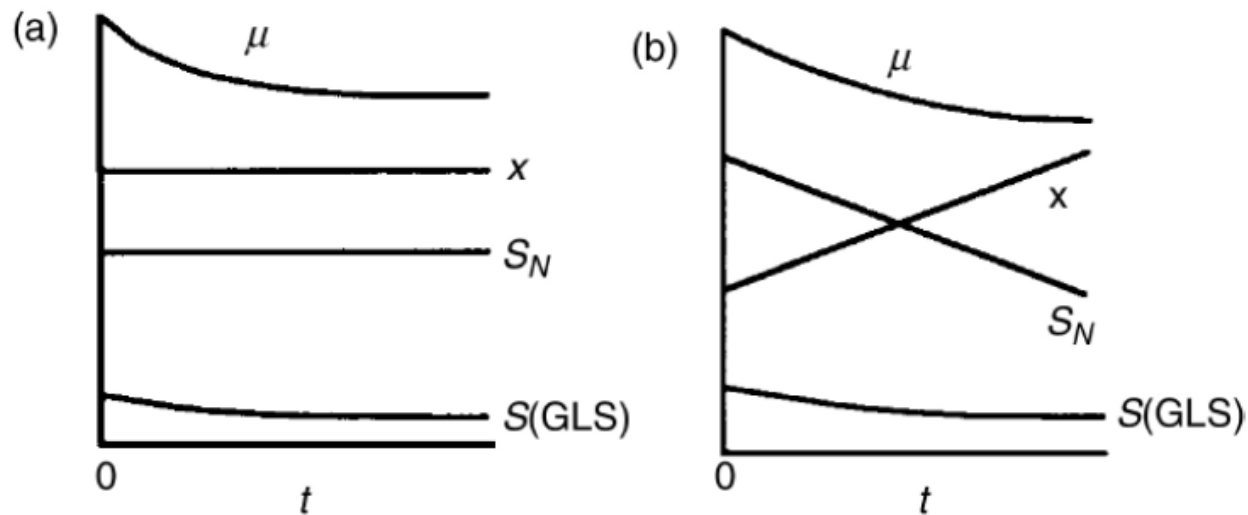


FIGURE 2.15 Time Profiles of Fed-Batch Cultures

μ , specific growth rate; x , biomass concentration; $S(\text{GLS})$, growth limiting substrate; S_N , any other substrate than $S(\text{GLS})$. (a) Variable volume fed-batch culture. (b) Fixed volume fed-batch culture (Pirt, 1979).



However, in the genuine steady state of a chemostat, dilution rate and growth rate are constant whereas in a fed-batch quasi steady state they change over the time of the fermentation.

Product concentration in the chemostat will reach a steady state, but in a fed-batch system the profile of the product concentration over the time of the fermentation will be dependent on the relationship between q_p and μ (hence D).

If q_p is strictly growth related then it will change as μ changes with D and, thus, the product concentration will remain constant.

If q_p is constant and independent of μ , then product concentration will decrease at the start of the cycle when D_p is greater than $q_p x$ but will rise with time as D decreases and $q_p x$ becomes greater than D_p .

These relationships are shown [in Fig. 2.16a](#). If q_p is related to μ in a complex manner, then the product concentration will vary according to that relationship. Thus, the feed strategy of a fed-batch system would be optimized according to the relationship between q_p and μ .

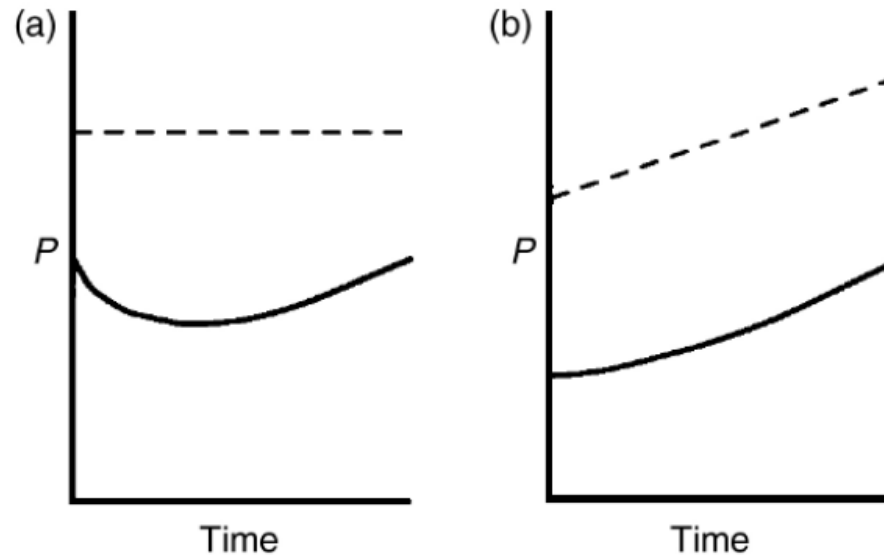


FIGURE 2.16

Product Concentration (p) in Fed-Batch Culture When q_p is Growth Related (-----) or Nongrowth Related, that is, q_p Constant (——). (a) Variable volume fed-batch culture. (b) Fixed volume fed-batch culture.

Modified from Pirt, 1979



FIXED VOLUME FED-BATCH CULTURE

Pirt (1979) described the kinetics of fixed volume fed-batch culture as follows. Consider a batch culture in which the growth of the process organism has depleted the limiting substrate to a limiting level. If the limiting substrate is then added in a concentrated feed such that the broth volume remains almost constant, then:

$$\frac{dx}{dt} = GY \quad (2.35)$$

where G is the substrate feed rate ($\text{g dm}^{-3} \text{h}^{-1}$) and Y is the yield factor.

But $dx/dt = \mu x$, thus substituting for dx/dt in Eq. (2.35) gives:

$\mu x = GY$, and thus:

$$\mu = \frac{GY}{x} \quad (2.36)$$



Provided that GY/x does not exceed μ_{\max} then the limiting substrate will be consumed as soon as it enters the fermenter and $ds/dt = 0$. However, dx/dt cannot be equated to zero, as in the case of variable volume fed-batch, because the biomass concentration, as well as the total amount of biomass in the fermenter, will increase with time. Biomass concentration is given by the equation:

$$x_t = x_{0a} + GYt \quad (2.37)$$

where x_t is the biomass after operating in fed batch for t hours and x_{0a} is the biomass concentration at the onset of fed-batch culture.

As biomass increases then the specific growth rate will decline according to Eq. (2.36). The behavior of a fixed volume fed-batch culture is illustrated in [Fig. 2.15b](#) from which it may be seen that μ declines (according to Eq. 2.36), the limiting substrate concentration remains virtually constant, biomass increases and the concentrations of the non limiting nutrients decline.



Pirt (1979) described the product balance in a fixed volume fed-batch system as:

$$\frac{dp}{dt} = q_p x$$

but substituting for x from Eq. (2.37) gives:

$$\frac{dp}{dt} = q_p (x_a + GYt)$$

If q_p is strictly growth-rate related then product concentration will rise linearly as for biomass. However, if q_p is constant then the rate of increase in product concentration will rise as growth rate declines, that is, as time progresses and x increases. These relationships are shown in [Fig. 2.16b](#).

If q_p is related to μ in a complex manner then the product concentration will vary according to that relationship.

As in the case of variable volume fed-batch the feed profile would be optimized according to the relationship between q_p and μ .



FED-BATCH CULTURE AT A CONSTANT SPECIFIC GROWTH RATE

It is possible to operate both a variable volume and a fixed volume fed-batch culture at a constant specific growth rate by increasing the feed rate exponentially over the period of the fermentation. Lee (1996) gave the following equation, which can be used to program the control system:

$$G_t = F_t \cdot S_t = \left(\frac{\mu}{Y_{x/s}} + m \right) x_t \cdot V_t$$

where G_t is the mass flow rate of the feed at time t (g h^{-1}) F_t is the volumetric flow rate at time t ($\text{dm}^3 \text{h}^{-1}$), S_t is the substrate concentration in the feed medium at time t (g dm^{-3}), μ is the specific growth rate (h^{-1}), $Y_{x/s}$ is the yield factor ($\text{g substrate g}^{-1}$ biomass), m is the maintenance coefficient ($\text{g substrate g}^{-1}$ biomass h^{-1}), x_t is the biomass concentration at time t (g dm^{-3}), and V_t is the volume of culture in the fermenter at time t (dm^3).

During growth at the constant specific growth rate (μ), the total amount of biomass in the fermenter ($x_t \cdot V_t$) increases exponentially with time according to the equation:

$$x_t \cdot V_t = x_0 \cdot V_0 \cdot e^{\mu t}$$



where x_0 is the initial concentration of the biomass, t is the time, and V_0 is the initial volume of the culture. Thus:

$$G_t = F_t \cdot S_t = \left(\frac{\mu}{Y_{x/s}} + m \right) x_0 \cdot V_0 \cdot e^{\mu t}$$

The application of this equation assumes that the yield factor and maintenance energy are constant, an assumption which does not always hold true, as discussed earlier. If the maintenance energy is very low then this has been disregarded in the equation by some workers

CYCLIC FED-BATCH CULTURE

The life of a variable volume fed-batch fermentation may be extended beyond the time it takes to fill the fermenter by withdrawing a portion of the culture and using the residual culture as the starting point for a further fed-batch process. The decrease in volume results in a significant increase in the dilution rate and thus, eventually, in an increase in the specific growth rate.



The design of a fermenter will involve cooperation between experts in microbiology, biochemistry, chemical engineering, mechanical engineering, and costing.

The most commonly used **fermenters** are based on a stirred upright cylinder with sparger aeration.

This type of vessel can be produced in a range of sizes from 1 dm³ to 1000s of dm³.



Table 7.2 Details of Geometrical Ratios of Fermenters With Single Multiblade Impellers (Fig. 7.1)

Dimension	Steel and Maxon (1961)	Wegrich and Shurter (1953)	Blakeborough (1967)
Operating volume	250 dm ³	12 dm ³	—
Liquid height (L)	55 cm	27 cm	—
L/D (tank diameter)	0.72	1.1	1.0–1.5
Impeller diameter (P/D)	0.4	0.5	0.33
Baffle width/D	0.10	0.08	0.08–0.10
Impeller height/D	—	—	0.33

Table 7.3 Details of Geometrical Ratios of Fermenters With Three Multibladed Impellers (Fig. 7.2)

Dimension	Jackson (1958)	Aiba, Humphrey, & Millis (1973)	Paca, Ettler, & Gregr (1976)
Operating volume	—	100,000 dm ³ (total)	170 dm ³
Liquid height (L)	—	—	150 cm
L/D (tank diameter)	—	—	1.7
Impeller diameter	0.34–0.5	0.4	0.33
Baffle width/D	0.08–10	0.095	0.098
Impeller height/D	0.5	0.24	0.37
P/V	0.5–1.0	—	0.74
P/W	0.5–1.0	0.85	0.77
P/Y	0.5–1.0	0.85	0.77
P/Z	—	2.1	0.91
H/D	1.0–1.6	2.2	2.95

1. [Tables 7.2 and 7.3](#) give geometrical ratios of various of the dimensions which have been quoted in the literature for a variety of sizes of vessel. [Moucha, Rejl, Kordac, and Labik \(2012\)](#) have described the design of multiple-impeller fermenters with particular emphasis on scale-up, oxygen mass transfer, and the limitations of experimental kLa data.



ASEPTIC OPERATION AND CONTAINMENT

Aseptic operation involves protection against contamination and it is a well established and understood concept in the fermentation industries, whereas containment involves prevention of escape of viable cells from a fermenter or downstream equipment and is much more recent in origin.

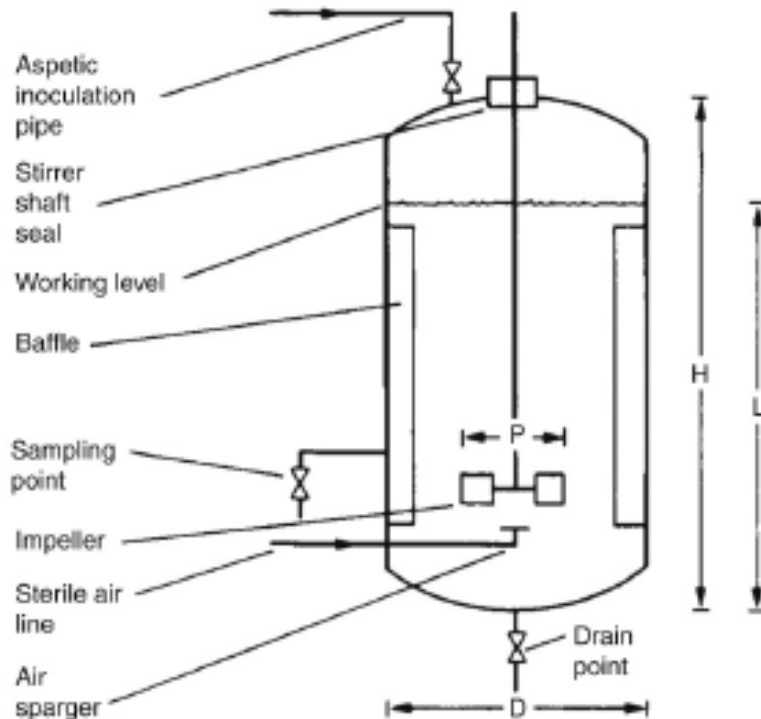


Diagram of a Fermenter With One Multibladed Impeller

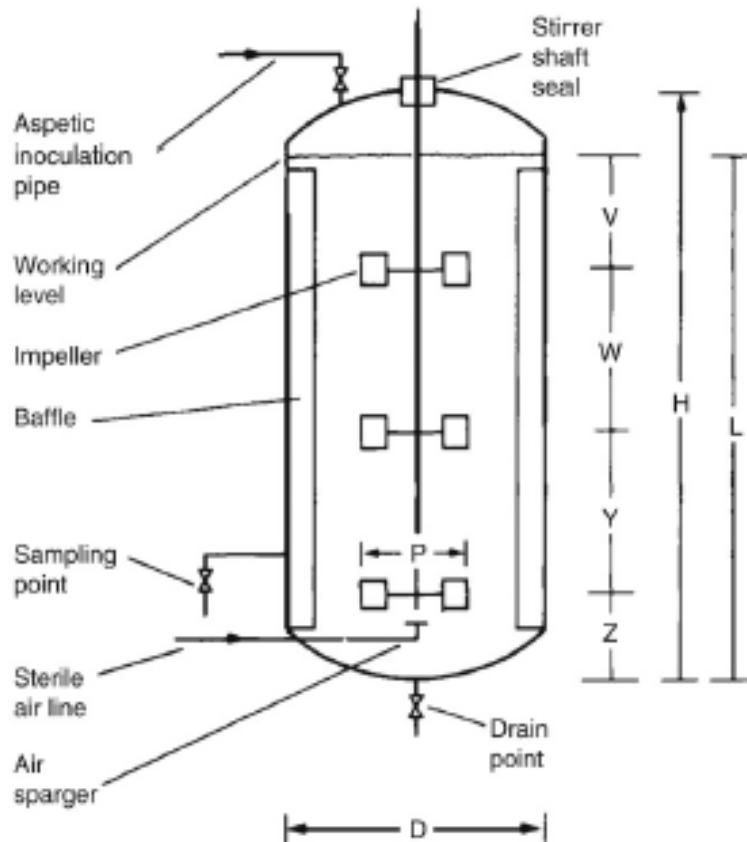


Diagram of a Fermenter With Three Multibladed Impellers



Basic function of bioreactors:

The main function of fermenter is to provide a controlled environment for growth of a micro organism or a defined mixture of micro organisms, to obtain a desired product while bio reactor referred to production unit of mammalian cell cultures.

The vessel should be capable of being operated aseptically for a number of days and should be reliable for long term operation. The adequate aeration and agitation should be provided to meet the metabolic requirements of the microbes. However the mixing should not damage the micro organisms. The power consumption should be low and temperature and pH control system should be provided.

The evaporation losses from the fermenter should not be excessive. The vessel should be designed to require the minimal use of labor in operation, harvesting, cleansing and maintenance. It should have proper sampling facility. The vessel should be constructed to infuse instead of flange joints. The cheapest and the best material should be used and there should be adequate service provision for individual plants.

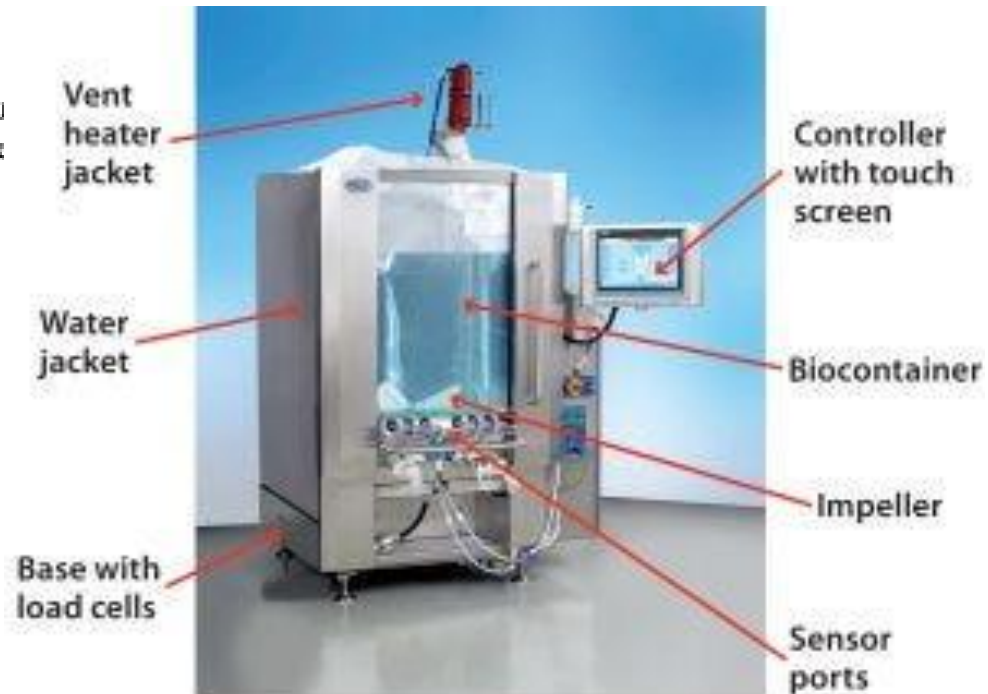
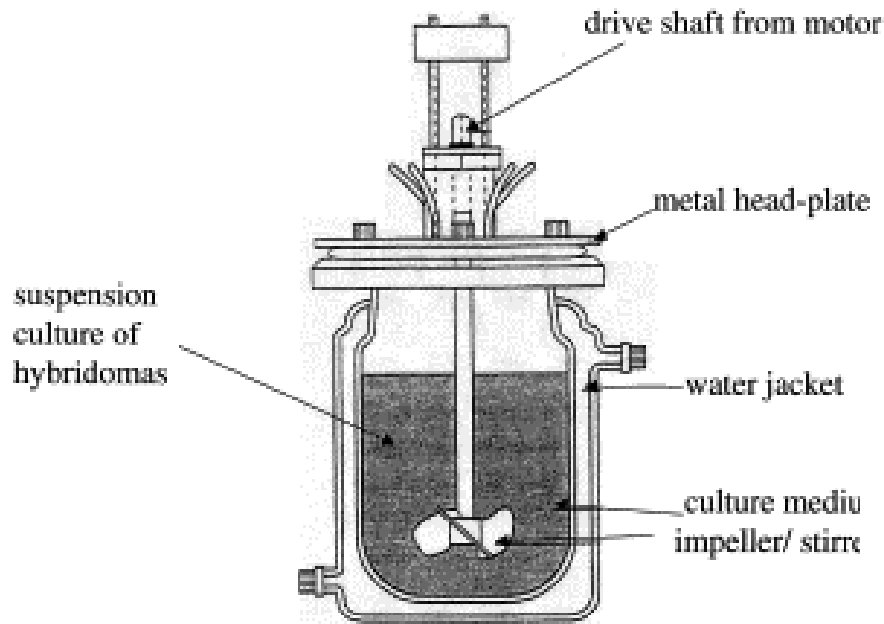


Types of Bioreactors.

Based on design of the bioreactors they can be grouped into the following types:

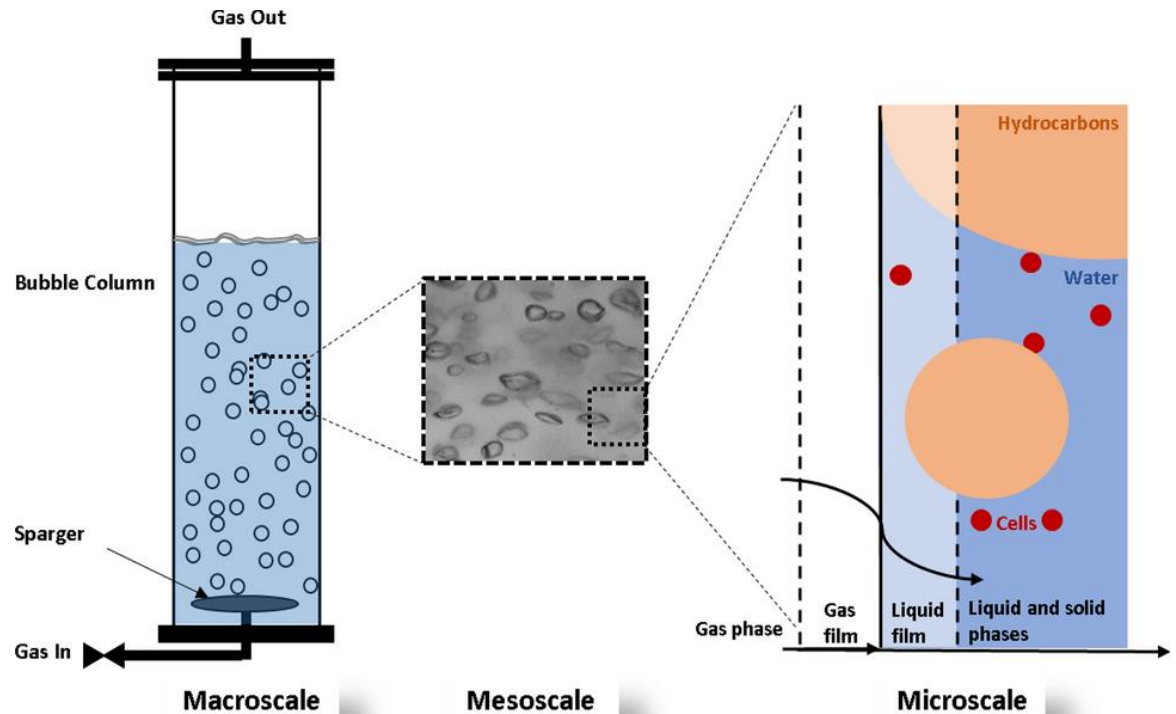
Continuous stirred tank bioreactor: A continuous stirred tank bioreactor consists of a cylindrical vessel with motor driver central shaft that supports one or more agitator(impeller).The shaft is fitted at the bottom of the bioreactor. The no. of impeller is variable and depends on the size of the bioreactor. Different types of impellers (concave bladed, marine propeller etc.,) are in use.

In stirred tank bioreactor ,the air is added to the culture medium under pressure through a device called sparger. The sparger may be the ring with many holes or a tube with single orifice. The sparger along with impeller(agitators) enable better gas distribution system throughout the vessel. The bubbles generated by sparger are broken down to small ones by impellers and dispersed throughout the medium. This enables the creation of a uniform and homogenous environment throughout the bioreactor.





Bubble column bioreactor: In the bubble column bioreactor the air or gas is introduced at the base of the column through perforated pipes or plates, or metal microporous spargers. The flow rate of the air/gas influences the performance factors- O_2 transfer, mixing. The bubble column bioreactor may be fitted with perforated plates to improve performance. The vessel used for bubble column bioreactors is usually cylindrical with an aspect ratio of 4-6 (i.e., high to diameter ratio)





Bubble columns are extensively used in many industrial applications.

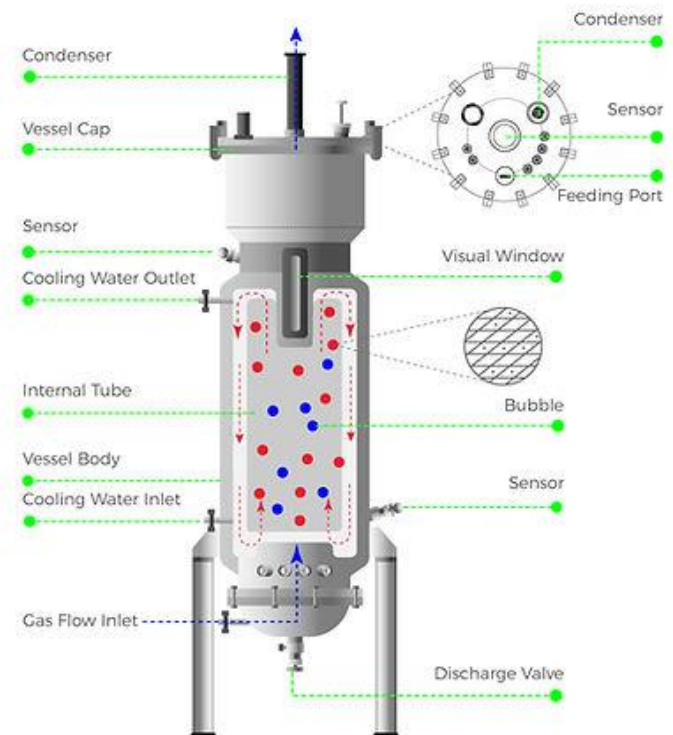
They are of considerable interest in chemical processes involving reactions like oxidation, chlorination, alkylation, polymerization, and hydrogenation, as well as in the production of synthetic fuels via a gas conversion process (Fischer-Tropsch process) in biochemical processes such as fermentation and biological wastewater treatment



Airlift bioreactors: Also known as tower reactor, an airlift bioreactor can be described as a bubble column containing a draft tube. Many types of airlift bioreactors are currently in use today. Air is typically fed through a sparger ring into the bottom of a central draft tube that controls the circulation of air and medium. Air flows up the tube forming bubbles and exhaust gas leaves at top of the column. The degassed liquid then flows downward and the product is drained from the tank. There are two types

a.) *Internal loop airlift bioreactor:* Has a single container with a central draft tube that creates interior liquid circulation channels.

b.) *External-loop airlift bioreactor:* This system consists of a riser and an external down-comer, which are connected at the bottom and the top respectively. As the injected air at the bottom of the riser creates gas bubbles that begin to rise through the main tank, exhausted gas leaves at the top and the resulting heavier solution descends through the down comer.





Applications of Airlift Bioreactor

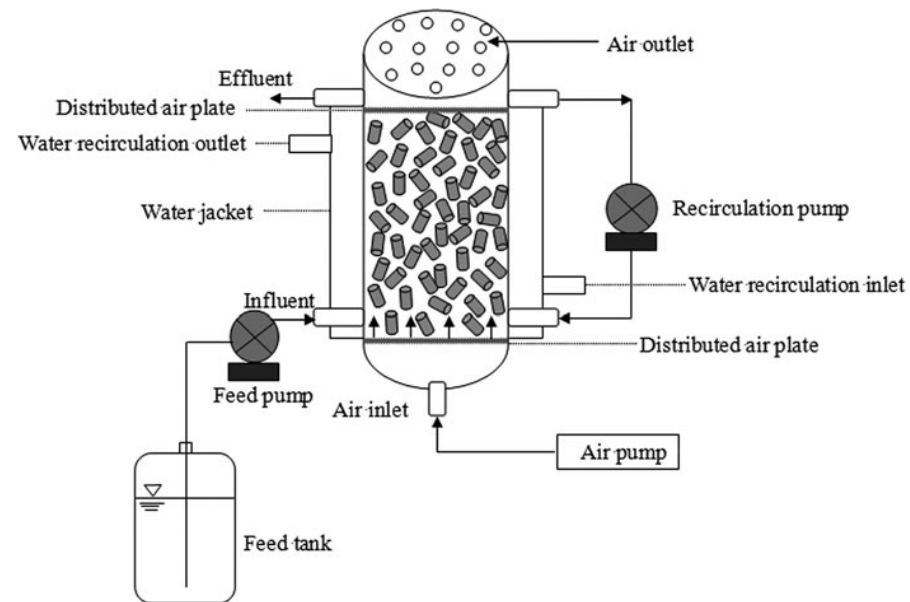
- Ideal for culturing cells that are too sensitive to the shear stress and heat generation caused by the mechanical agitation.
- Pallet form fermentation; immobilized enzyme reactions.
- SCP production from methanol as carbon substrate.
- Typically, airlift bioreactors are used when the desired reactants and/or final products are in a gaseous state and for aerobic cell cultures.





4. Packed bed bioreactors: A bed of solid particles with biocatalysts on or within the matrix of solids, packed in a column constitute a packed bed bioreactor. The solid may be porous or non porous gels, and may be compressible or rigid in nature. A nutrient broth flows continuously over the immobilized biocatalyst. The product obtained in the packed bed bioreactor are released into the fluid and removed .While the flow of the fluid can be upward or downwards, downflow under gravity is preferred.

The advantages of packed bed bioreactor over other bioreactors are reuse of enzymes, continuous mode of operations, low substrate and product inhibition, high yield of desired product. These bioreactors are widely applied for valorization of food, beverage, nutraceutical synthesis, as well as waste treatment.





5. Photobioreactors: These are the reactor specialized for fermentation and can be carried out either by exposing to sunlight or artificial illumination. Since artificial illumination is expensive, only the outdoor photo-bioreactor are preferred. Certain important compounds are produced by employing photo-bioreactors e.g., Beta carotene, asthaxanthin.

The photoreactors are made up of glass or more commonly transparent plastic. The array of tubes or flat panels constitute light receiving system (solar receiver). The culture can be circulated through the solar receiver by methods such as using centrifugal pumps or airlift pump. It is essential that the cells are continuous circulation without forming sediments. Further adequate penetration of sunlight should be maintained. The tubes should also be cooled to prevent rise in temp. Photobioreactors are usually operated in a continuous mode at a temp in the range 25-40°C. Microalgae and cyanobacteria are normally used. The organisms grow during day light while the products are produced during night.



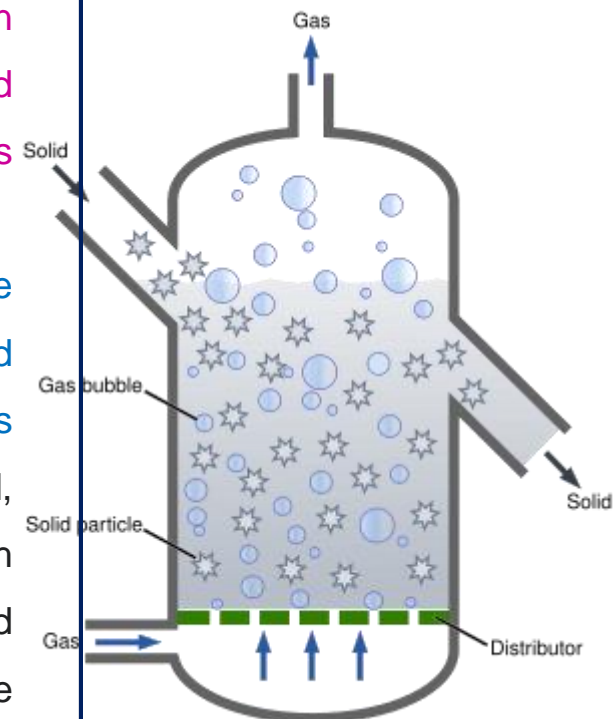
The advantages of tubular photobioreactors at production scale are also transferred to laboratory scale. A combination of the mentioned glass vessel with a thin tube coil allows relevant biomass production rates at laboratory research scale. Being controlled by a complex process control system the regulation of the environmental conditions reaches a high level



A **fluidized bed reactor (FBR)** is a type of reactor device that can be used to carry out a variety of multiphase chemical reactions. In this type of reactor, a fluid (gas or liquid) is passed through a solid granular material (usually a catalyst) at high enough speeds to suspend the solid and cause it to behave as though it were a fluid. This process, known as fluidization, imparts many important advantages to an FBR. As a result, FBRs are used for many industrial applications.

The solid substrate material in the fluidized bed reactor is typically supported by a porous plate, known as a distributor. The fluid is then forced through the distributor up through the solid material. At lower fluid velocities, the solids remain in place as the fluid passes through the voids in the material. This is known as a packed bed reactor.

As the fluid velocity is increased, the reactor will reach a stage where the force of the fluid on the solids is enough to balance the weight of the solid material. This stage is known as incipient fluidization and occurs at this minimum fluidization velocity. Once this minimum velocity is surpassed, the contents of the reactor bed begin to expand and swirl around much like an agitated tank or boiling pot of water. The reactor is now a fluidized bed. Depending on the operating conditions and properties of solid phase various flow regimes can be observed in this reactor.





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