Modeling Effects of U-tube Aeration in Recirculating Aquaculture Systems

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MODELING EFFECTS OF U-TUBE AERATION IN RECIRCULATING AQUACULTURE SYSTEMS

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(ABSTRACT)

Cost effective oxygenation equipment design and operation in high density recirculating aquaculture systems (RAS) requires consideration of the effects of RAS component design, water quality, and biological processes on oxygen transfer requirements. A computer model was developed to simulate multi-component gas transfer and biological respiration in a pilot scale recirculating aquaculture system employing an u-tube aerator.

Gas transfer and/or biological respiration are modeled for all system components; production tank, multi-tube clarifier, rotating biological contactor and the u-tube. The calibrated model predicted equilibrium oxygen concentrations in the production tank to within ± 1 mg/L in the validation tests. The model was then used to investigate techniques to reduce oxygen transfer costs associated with aquaculture in the RAS located at the Virginia Tech Aquaculture Research Facility.
Acknowledgements

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1. INTRODUCTION

The first known attempts at culturing aquatic organisms date as early as 500 B.C. in eastern Asia. Only within the past three decades have dedicated efforts been undertaken to domestically raise foodfish at production rates beyond what would generally be considered "subsistence levels." In contrast, agricultural research in the United States has resulted in reliable and economical production of land-based crops despite the fact that nearly three quarters of the earth's surface is covered by water. Man's alienation from the aquatic environment may have prompted this focus towards intensive but controlled cultivation of the and simultaneous unchecked exploitation of natural aquatic resources.

Aquaculture has the potential to become a major supplier of inexpensive protein (USDA, 1991). Some underdeveloped countries may continue to utilize low input aquaculture methods for subsistence purposes; however, in most developed nations, the bulk of current production and research is primarily motivated by commercial interest in producing high cash value aquaculture products, such as trout, salmon, and hybrid striped bass. Recent increases in production of these products, especially in the United States, may be attributed in part to the health conscious "baby boomer" generation. This health awareness has generated significant increased demand for high quality fish products. In 1989, seafood consumption was estimated at 6.8 kg per capita and is projected to increase to 9.1 kg by the year 2001, an increase of over 30%.
The aquaculture and seafood industries have responded by increasing both domestic production and imports of seafood products. In 1990, approximately 172 million kg of catfish were processed, an increase of 5% over 1989 and 19% over 1988 (USDA, 1991). Production of processed catfish has been projected to possibly increase an additional 20% by 1995. Farm-raised salmon in the U.S. totaled approximately 4.1 million kg in 1990 and is expected to increase (USDA, 1991). In 1989 approximately 0.45 million kg of farm-raised hybrid striped bass were produced and production increases were projected for 1990 and 1991. Sales of food size trout increased 7% from 1989 to 1990, totaling $64.5 million (USDA, 1991). However, despite increased domestic production for many species and recent relatively stable commercial harvests, over one half the nation's seafood was imported in 1990.

There are strong incentives to develop methods for increasing domestic production of aquaculture products. Extensive production techniques, such as traditional pond culture for finfish and utilization of natural water bodies for mollusk production, are becoming restricted by decreases in suitable land acreage with access to high quality water supplies. Productions is also increasingly restricted due to commercial development, governmental protection of natural coastal environments, and environmental pollution. These factors combined with the demand for high value seafood products have created a demand for the development of technology for intensive foodfish production coupled with minimal water use.

Intensive raceway culture has been utilized by the trout industry for many years. Such flow through systems require large flows of high quality water undergoing
serial use through several separate production units before discharge back to the environment. With the possible exception suspended solids removal, there is typically little removal of pollutants and nutrients prior to discharge.

Recirculating and partial reuse aquaculture systems attempt to minimize total water consumption by continuously recirculating water from the production unit through a reconditioning or waste treatment process and back to the production unit. This process typically involves three basic steps: suspended solids removal, biological treatment of dissolved waste products produced by feed and fish, and aeration to replace oxygen consumed by biological respiration. Energy requirements and associated monetary costs to maintain adequate water quality for acceptable fish growth and health are a direct function of the types of processes employed and the fraction of the recirculated flow that is discharged and replaced by clean, unused water.

In the past, attempts at commercial production of aquaculture products in the United States in recirculating systems have been uneconomical due primarily to the required energy inputs. Recirculating aquaculture systems are defined as those systems in which less than 1 to 2% of the recirculation flow is continually discharged and replaced by new water. Partial reuse systems, those systems in which more than 5 to 10% of the recirculated flow is discharged and replaced by clean water, have proven economical depending upon their location and water source. Attempts at production of coldwater species such as trout from acceptable underground spring sources have proven successful in partial reuse systems. Systems employing year around sources of inexpensive warmwater, such as geothermal aquifers or possibly waste heat from industrial sources, have resulted
in successful production of some warmwater species. Recent research has improved process water treatment techniques suitable for aquaculture and understanding of water quality effects on growth and development of potential aquaculture species. These factors, combined with increasing demand for high value fish products, suggest that economical production in commercial scale recirculating systems may be a realistic objective for the aquaculture industry in the near future.

The Aquaculture Research Facility at Virginia Polytechnic Institute and State University was established in 1988. A fundamental objective of the Virginia Tech Aquaculture Facility (VTAF) is to develop and disseminate information pertaining to the design, operation, and management of recirculating aquaculture systems (RAS). Nine identical pilot scale systems were constructed. Each consists of a production tank, a multi-tube clarification unit for solids settling, a rotating biological contactor (RBC) for nutrient oxidation, and a u-tube aerator coupled with pure oxygen injection to meet the biological oxygen demands of the system (Fig. 1.1).

To determine the economic viability of these systems, the performance and efficiencies of the process components are being quantified. Individual contributions to total production costs can be quantified to see where further improvements in performance may be most beneficial. Furthermore, component performance can compared with performance of alternative components.
Rotating Biological Contactor  
(total surface area= 331 m³)

Multi-tube Clarifier  
(overflow rate= 127 L/min/m²)

Production Tank  
(9000 L)

U-tube Aerator  
(depth= 12.3 m)

Figure 1.1 Pilot Scale Recirculating Aquaculture System
Costs associated with oxygen transfer required to meet biological demands in high density systems can be excessive if aeration equipment is improperly designed. Conventional aeration techniques utilized in pond culture, (surface agitation and shallow depth air diffusion) are inefficient and extremely cost prohibitive in intensive culture systems due to low oxygen transfer rates. The dissolved oxygen (DO) deficit, the difference between saturation concentration and actual concentration influences the rate of oxygen transfer into or out of the water. Oxygen deficits established by differences in minimum DO concentrations required by fish and atmospheric saturation are inadequate for energy efficient oxygen transfer at high transfer rates. Over the last decade, aeration processes, or more properly termed, oxygenation processes, employing pure oxygen have been developed to increase the rate of oxygen transfer to the water and reduce oxygenation costs. U-tube oxygenation, the method chosen to be used at the VTAF is just one of these methods.
2. **OBJECTIVES**

The focus of this thesis research is to study the performance of the u-tube aerators at the VTAF under production conditions and to investigate the effects of u-tube aeration within a recirculating aquaculture system. A computer model capable of performing a mass balance analysis on dissolved oxygen, nitrogen, and carbon dioxide in a recirculating aquaculture system under production conditions was developed to predict oxygenation performance.

Assessment of the 1989-90 production trial data, from the nine pilot scale recirculating systems, indicate that oxygen costs per unit mass of fish produced were higher than expected. Total oxygen costs, including both oxygen usage and rent for the liquid oxygen containment system, were approximately $0.31 per kg of fish produced (Libey, pers. comm.). Initial evaluations of u-tube performance early in the 1991 production trials indicated that these costs could be reduced by capture and reutilization of oxygen-rich off-gases wasted from the u-tube effluent or by continuously controlling u-tube operation in response to diurnal oxygen demand cycles exhibited by the RAS. The 1991 production trials were initiated in January, 1991, and continued for six months to investigate effects of RAS fish stocking density on water quality and growth rate.

A computer model capable of simulating gas transfer and consumption within all components of the recirculating system would be beneficial for preliminary assessment of techniques to improve oxygenation performance and to predict conditions of potential oxygen depletion or supersaturation within the production
tank. The research presented herein was initiated with the following specific objectives:

1. Develop a computer model capable of simulating gas transfer, carbon dioxide production, and oxygen consumption within a recirculating aquaculture system under operating conditions and to estimate equilibrium conditions by mass balance analyses.

2. Calibrate and validate the model with data from a full scale system with controlled oxygen demand similar to those observed during the 1991 Production trials.

3. Quantify the effects of water quality on oxygenation performance as opposed to the clean water conditions under which model was calibrated.

4. Assess the potential of several proposed methods to reduce costs associated with oxygen transfer in the VTAF RAS.
   a. Changing RAS recirculation flow rate
   b. Optimizing supplemental oxygen transfer across the RBC
   c. Reutilization of off-gases (off-gas recycle).
   d. Control of u-tube operation in response to diurnal oxygen demand cycles exhibited in the RAS.
3. LITERATURE REVIEW

Much effort was devoted during the last decade to improving waste treatment processes due to increasingly stringent discharge regulations. Likewise, a heightened interest in intensive aquaculture has prompted research focused on optimizing performance of treatment processes used in aquaculture systems. Often it is desirable to model such treatment processes to predict the effects on process performance when changes are proposed or made one or more of the system's components. Several researchers have modeled aeration and oxygenation process effects on biological waste treatment processes to predict BOD reduction, oxygen depletion, and the effects of contaminant gases on the oxygen transfer process (Mueller, Famularo, and Mulligan, 1979; Mueller, Mulligan, and Di Toro. 1973).

A similar model for recirculating aquaculture systems would be useful in predicting comparable effects. Assuming individual component contributions to changes in dissolved gas concentrations can be accurately modeled under a given set of conditions, system response to the combined effects of biological respiration, component performance, and continual recirculation can be predicted. Similar to waste treatment models, the model would predict required operational criteria to maintain dissolved gas concentrations oxygen and carbon dioxide necessary for optimum fish production and provide preliminary evaluations of changes in component operation or design. Therefore, this literature review attempts to discuss those topics relevant to modeling gas transfer in a recirculating
system employing u-tube aeration. The topics include gas transfer theory and factors influencing gas transfer, a review of existing u-tube models and their applications, and a brief overview of pure O₂ aeration processes to provide a comparative basis for evaluating u-tube performance. Before addressing these topics, the biological issues associated with aeration in fish culture are briefly reviewed. These topics include predicting fish respiration, criteria for DO, dissolved CO₂, and the effects of total dissolved gas pressure.

3.1. Biological Considerations for Modeling Oxygen Transfer in Fish Culture Systems

3.1.1. Predicting Fish Respiration

Intensive production in pond and raceway culture has often been limited by the inability to continuously meet DO requirements of the biological system and maintain adequate DO concentrations for optimum fish growth. Any attempt to optimize design of intensive fish culture systems or model their performance requires a reasonably valid estimate of total O₂ utilization, both macrophytic and microphytic, under culture conditions. While there has been much research to estimate O₂ consumption rates of fish, much of the data may be of limited value for recirculating systems because metabolic rates were established for small numbers of fish in respirometers (Andrews and Matsuda, 1975).

Oxygen consumption is influenced by physical activity, fish weight, feeding rations, temperature and to some degree other water quality parameters such as CO₂ level, hyperbaric total dissolved gas pressures, and ambient DO concentrations (Watten and Boyd, 1989). Perhaps the most useful parameters for
predicting \( \text{O}_2 \) consumption under culture conditions are fish weight, (average and total), water temperature, and feeding ration. Andrews and Matsuda (1975) investigated these effects on \( \text{O}_2 \) consumption by channel catfish. Oxygen consumption per kg of fish weight decreased with increasing fish size and decreased with decreasing ambient DO concentrations. Increasing temperature increased metabolic rates in accordance with Van't Hoff’s Law for chemical reactions. A common rule of thumb for aerobic biological organisms is a 10° C change in temperature will double the \( \text{O}_2 \) consumption rate.

Andrews and Matsuda (1975) found that under all conditions except extraordinarily low DO concentrations of 1 to 2 mg/L, well-fed catfish consumed more \( \text{O}_2 \) than fasted catfish. Well-fed catfish with an average weight of 25 g at 7 mg/L DO and 26° C, utilized \( \text{O}_2 \) at 900 mg-\( \text{O}_2 \)/kg\(_{\text{fish}} \)/hr while fasted catfish consumed \( \text{O}_2 \) at a rate of 600 mg-\( \text{O}_2 \)/kg\(_{\text{fish}} \)/hr. Well-fed and fasted catfish at an average weight of 1000 g and same water conditions exhibited \( \text{O}_2 \) consumption rates of approximately 480 and 260 mg-\( \text{O}_2 \)/kg\(_{\text{fish}} \)/hr, respectively. The majority of \( \text{O}_2 \) consumption relationships established under culture conditions utilize some combination of temperature, fish weight and feeding ration as predictors.

Boyd et al. (1978) utilized the data of Andrews and Matsuda (1975) to develop the following regression equation (correlation coefficient = 0.99) for predicting \( \text{O}_2 \) consumption of channel catfish:

\[
\text{DEM} = 10 \left( ^{0.999} - 0.000957 W + 0.0000006 W^2 + 0.0327 T_e - 0.0000087 T_e^2 + 0.0000003 W T_e \right) \quad [1]
\]

where
DEM  - average O\textsubscript{2} consumption, g O\textsubscript{2}/kg\textsubscript{fish}/hr

W  - average weight of fish, g

T\textsubscript{c}  - water temperature, °C. (Refer to Appendix C for definitions of all symbols and acronyms).

Oxygen consumption rates may also vary somewhat with species. Boyd and Watten (1989) cite a range of 205 to 888 mg-O\textsubscript{2}/kg\textsubscript{fish}/hr for freshwater fish. An equation for O\textsubscript{2} consumption in rainbow trout was developed by Muller-Feuga et al. (1978).

\[
DEM = A \cdot W^B \cdot 10^{C_c} \tag{2}
\]

where

DEM  - average O\textsubscript{2} consumption, mg O\textsubscript{2}/kg\textsubscript{fish}/hr

A  - Constant = 75 (4 to 10° C) or 249 (12 to 22° C)

W  - average weight, g

B  - constant = -0.196 (4 to 10° C) or -0.142 (12 to 22° C)

C  - constant = 0.055 (4 to 10° C) or 0.024 (12 to 22° C).

For 50 g fish and 1000 g fish at a temperature of 20° C, predicted consumption rates are 403 and 200 mg-O\textsubscript{2}/kg\textsubscript{fish}/hr for catfish (Eq [1]), and 431 and 282 mg-O\textsubscript{2}/kg\textsubscript{fish}/hr for trout, (Eq [2]), indicating trout may consume more O\textsubscript{2} at 20° C than catfish.

Feeding level exerts a pronounced effect on O\textsubscript{2} consumption and may be a more useful design parameter than weight, temperature, or species for estimating O\textsubscript{2} demands. Reviews of the literature and experience in raceway culture of trout...
indicate that approximately 0.2 to 0.22 kg O₂ are required per kg feed. (Colt and Watten, 1989; Colt and Orwicz, 1991). The ratio of required O₂ to feed is termed the oxygen/feed ratio (OFR). The value of OFR may vary considerably with species, temperature, and nutritional value of the feed, and little data on the effects of these parameters on OFR are available (Colt and Watten, 1989). Even though very little data exists for intensive warmwater culture, OFR for warmwater species is often assumed to be significantly higher than for coldwater species, and a value as high as 1.0 has been suggested (Colt and Watten, 1989). African catfish evaluated in respirometers under laboratory conditions consumed O₂ at a maximum OFR of 0.37 kg O₂/kg feed while decreased consumption, OFR = 0.26, was observed for lower feeding levels (Hogendoorn, 1983). Other researchers have suggested calculating the OFR based upon the energy content of the feed. However, in respirometer experiments, Hogendoorn (1983) found that 25 to 37% of the energy contained in feed consumed by catfish in controlled respirometer experiments was unaccounted for either in settleable faecal and feed particles or metabolic energy, ie. weight gain and heat release.

The observations of the above researchers illustrate the effects of feed ration and temperature on average O₂ demand but give no indication of the range or cyclic nature of O₂ demands. In pond culture, diurnal DO concentrations cycle with daily lows generally occurring shortly before dawn and daily highs occurring mid to late afternoon due to photosynthesis and respiration of phytoplankton populations. Oxygen demands within recirculating and raceway intensive culture systems, however, cycle in relation to feeding schedules. Generally, O₂ demand (highest O₂ concentration) is at a minimum at early morning when fish are inactive and have not been fed for several hours. As light levels increase, fish activity
increases with the anticipation of feeding and the demand begins to rises (Brett and Groves, 1979). The daily maximum may be influenced by the timing, feeding size, and number of daily feedings and may peak at values 2 times the daily average. The peak demand may occur at the time of feeding (Brett and Groves, 1979) or 4 to 6 hours after feeding (Colt and Watten, 1989). A peaking factor of 1.44 has been recommended for establishing the maximum demand from the average demand (Wester, 1981). Clearly, oxygenation equipment should be designed to cost effectively operate over a wide range of anticipated demands.

3.1.2. Dissolved Gas Criteria

While limiting criteria for dissolved gases may not be directly required for modeling gas transfer, they establish allowable boundaries of operation for aeration systems and likewise a window within which model predictions should be evaluated as reasonable representations of desirable culture conditions. Little data exist which allow establishment of recommended criteria for allowable maxima and minima of various dissolved gas concentrations or partial pressures commonly associated with fish production. The partial pressure, rather than concentration, of most gases in solution determines the effects on fish. However, much of the data are reported as a concentration without water temperature stated and one should be aware of the relationships between partial pressure, temperature, and concentration when assessing gas criteria. One article (Colt and Orwicz, 1991) attempts to provide a generalized set of criteria from available data for many water quality parameters.
In addition to O₂, other gases of concern are nitrogen (N₂), and argon (Ar), and CO₂. Ammonia production in aquaculture systems is also a primary consideration, but due to its high solubility, direct removal by gas transfer is inappropriate. Methane or hydrogen sulfide production in clarifiers, where anaerobic conditions exist in settled solids, may occur to some extent but these gases are generally considered insignificant.

3.1.2.1. Oxygen

Minimum allowable levels of DO are difficult to define since many species of fish can survive at relatively low O₂ concentrations (2 to 3 mg/L) for indefinite periods of time. Suggested minimum levels are 6 to 7 mg/L for coldwater species and 4 to 5 mg/L for warmwater species (Colt and Orwicz, 1991). Continued exposure to low DO levels may lead to increased sensitivity to toxins or increased susceptibility to disease pathogens (Smart, 1981). Other researchers have observed significantly reduced growth rates in channel catfish at DO levels below 50 % of saturation (Andrews et al., 1973).

The recommended allowable maximum level for DO is approximately 300 mm Hg partial pressure. Oxygen toxicity may be an issue at high supersaturated DO levels, with toxic effects occurring at approximately 300 % air saturation for rainbow trout (50 mg/L at 15°C) (Boyd and Watten, 1989) and as low as 25 to 35 mg/L for striped bass (Colt and Watten, 1988). Researchers have found no benefits such as increased growth or feed conversion when DO levels are maintained above 70 to 90 % air saturation values if other water quality parameters are not growth limiting.
3.1.2.2. Carbon Dioxide and Hydrogen Sulfide

The recommended maximum level for CO$_2$ is 20 mg/L but in well-buffered high alkalinity waters, 80 - 150 mg/L as CaCO$_3$ with adequate DO, fish appear capable of tolerating high dissolved CO$_2$ levels, possibly in excess of 40 to 50 mg/L (Colt and Orwicz, 1991). Carbon dioxide is associated with formation of calcareous deposits in the kidneys of rainbow trout, a condition termed nephrocalcinosis (Smart et al., 1979). However, growth was only slightly affected at ambient dissolved CO$_2$ levels of 50 to 60 mg/L. A possible indirect effects of increasing CO$_2$ is a reduction in pH in low alkalinity waters. High CO$_2$ levels may also interfere with respiration by inhibiting O$_2$ transport in the blood. However, some evidence suggests that elevated DO levels may serve to negate that effect by increasing the partial pressure of O$_2$ in the blood supply.

Colt and Orwicz (1991) present no recommended criterion for hydrogen sulfide since under most culture conditions it is non-detectable. Hydrogen sulfide is moderately soluble in water and is extremely toxic to fish in minute amounts. Acute toxic effects on adult bluegills were observed at concentrations as low as 0.0448 mg/L and chronic effects marked by decreased growth were observed at levels ranging from 0.0031 to 0.0107 mg/L (Smith et al., 1976). Due to the combination of apparent toxicity at very low concentrations and solubility, toxic concentrations of hydrogen sulfide could occur in systems receiving makeup water from contaminated underground sources, especially in coastal regions. Small concentrations may also be produced in anaerobic treatment devices or in anaerobic sediments accumulating in clarifiers.
3.1.2.3. Total Gas Pressure

If total dissolved gas pressure exceeds barometric pressure, hyperbaric conditions exist, and the potential for development of Gas Bubble Disease, or more correctly Gas Bubble Trauma, GBT, in the fish population must be considered. Gas Bubble Trauma is a physically induced condition caused by an uncompensated, hyperbaric total dissolved gas pressure experienced by an aquatic organism resulting in formation of gas bubbles in blood and tissues. The uncompensated pressure difference, Eq [3], experienced by an organism is equal to the measured hyperbaric pressure difference minus the sum of the compensating pressures:

\[
\Delta P_{\text{uncomp}} = \Delta P_m - C \rho g Z - P_{\text{blood}} - P_{\text{surface}} \quad [3]
\]

where

- \( \Delta P_{\text{uncomp}} \) - uncompensated pressure change, mm Hg
- \( \Delta P_m \) - measured pressure difference between total dissolved gas pressure and atmospheric pressure (mm Hg)
- \( P_{\text{blood}} \) - blood or tissue pressure (mm Hg)
- \( P_{\text{surface}} \) - compensating pressure due to surface tension effects (mm Hg)
- \( \rho \) - density of water, kg / m\(^3\)
- \( g \) - acceleration due to gravity, m / s\(^2\)
- \( Z \) - depth, m
- \( C \) - conversion, = 0.00752 kPa/mm Hg

The compensating pressures include local hydrostatic pressure at a specific depth, blood and tissue pressure, and surface tension effects. (Bouck, 1980).
Surface tension, as well as tissue and blood pressure are often unknown and are difficult or impossible to measure. Blood and tissue pressure may vary between species, life-stage and physical condition of the organism. Therefore these pressures are generally omitted when determining $\Delta P_{uncomp}$. Under intensive culture conditions, water depths are generally shallow and the hydrostatic pressure is also often omitted. Total gas pressure is often expressed as a percentage relative to the local barometric pressure and is determined from $\Delta P_m$ (Colt, 1984):

$$TGP = \left( \frac{BP + \Delta P_m}{BP} \right) 100$$

[4]

where

- TGP - Total dissolved gas pressure (%)
- BP - Local barometric pressure (mm Hg)

Generally, $\Delta P_m$ is measured *in situ* by any one of several devices employing the membrane diffusion method. A length of gas permeable tubing connected to a low pressure measurement device is lowered into the water and the pressure inside the tubing, initially at equilibrium with the local barometric pressure is allowed to equilibrate to the total dissolved gas pressure in the liquid. A commonly used device, the Weis Saturometer, is commercially available or a device such as the "gasometer" (Bouck, 1980) can be constructed for use in many aquaculture applications.

The total gas pressure is undersaturated when TGP is less than 100 % ($\Delta P_m < 0$), saturated when TGP = 100 % ($\Delta P_m = 0$), and supersaturated when TGP is greater 100 % ($\Delta P_m > 0$). Since $\Delta P_m$ represents the difference in total dissolved gas
pressure and barometric pressure, $\Delta P_m$ is also the sum of the differences between the partial pressures of dissolved gases at their actual concentrations and their atmospheric partial pressures. Therefore, the total gas pressure may be supersaturated even if one or more individual gases are undersaturated. This is important when considering the roles of $O_2$ and $N_2$ partial pressures in causation of GBT, because GBT will not occur if $\Delta P_m < 0$ or $\Delta P_m = 0$ regardless of the degree of supersaturation of an individual gas. If the concentration of a gas is known, the partial pressure of a gas can be computed from the following formula (Colt, 1984):

$$P_i = \frac{A_i C_i}{B_i}$$

[6]

where

- $A_i$ - constant $= 760/(1000 K_i)$, for $i^{th}$ gas
- $B_i$ - Bunsen coefficient of $i^{th}$ gas, $L_{gas}/(L_{water, atm})$
- $K_i$ - ratio of molecular weight to molecular volume of $i^{th}$ gas, mg/mL
- $C_i$ - dissolved concentration of $i^{th}$ gas, mg/L.

In most situations, $N_2$, $Ar$, and $O_2$ are the only gases contributing significant partial pressures to total gas pressure. Unlike $O_2$, which can be measured by either wet chemical titration, (the Winkler procedure APHA, 1989), or by a commercially available DO meter, the concentration of $N_2$ or $Ar$ is not easily measured in water. Therefore, being biologically inert, $N_2$ and $Ar$ are often considered as a single gas, $N_2+Ar$. Once the DO concentration and $\Delta P_m$ are
known, their combined partial pressures are determined from the following equation (Colt, 1986):

\[ P_{N_2+Ar} = BP + \Delta P_a - \left( \frac{C_{O_2}}{B_{O_2}} \right) - P_w \]  

Historically, the N\textsubscript{2} and Ar partial pressures were considered the causative gases for GBT when TGP was supersaturated because they are biologically inert and may be held in simple solution in the blood stream. However, research now indicates oxygen partial pressure is a significant contributing factor although its role is not completely understood (Weitcamp and Katz, 1980). Unlike N\textsubscript{2}, the O\textsubscript{2} partial pressure is influenced by biological processes such as metabolism, binding with hemoglobin during blood transport, and swim bladder hyperinflation in fish. Much of the recent research indicates that GBT must be considered under two conditions, acute and chronic. The boundary is not well-defined. The risk of a given \(\Delta P\) to an aquatic organism is influenced by species, life-stage of the organism, temperature, total excess gas pressure, and the ratio of the N\textsubscript{2} and O\textsubscript{2} partial pressures (Colt, 1986).

Gas Bubble Trauma can be recognized by an array of symptoms or behavior. External signs associated with GBT are emphysema, recognizable as small bubbles, blisters, or lesions on the skin, fins, or along the lateral line; exophthalmia or "popeye" in which the eyeball is driven out of the socket due to gas pressure behind the eye; or emboli in blood vessels in the gill membranes due to bubble formation in the blood stream. Behavioral symptoms include refusal of food, listless swimming at the water surface, disorientation and loss of equilibrium, and writhing movements. Internal signs include, but are not limited
to: bubble coalescence in the auricle of the heart; emphysema and hemorrhages in muscular and organ tissues; and possibly gas pocket formation in the intestinal tract and body cavity (Weitcamp and Katz, 1980).

The N₂/O₂ partial pressure ratio is strongly correlated with acute and chronic effects. Increasing the N₂/O₂ ratio for a given supersaturated TGP value decreases the average time to death and decreasing the ratio exhibits the opposite effect (Nebeker et al., 1979; Nebeker et al., 1976; U.S. Nat. Marine Fisheries Service, 1975). Nebeker et al. (1976) reported ratios as percent saturation of O₂ to N₂ rather than partial pressure ratios of N₂/O₂. At TGP = 120 %, the mean time to death for 50 % of the sample population was 71 hours at an N₂/O₂ ratio of 117.2:121.3 %. However, at TGP = 120 % and a N₂/O₂ ratio = 170:107 %, only 7 % of the sample died after 167 hours. At high TGP and high N₂/O₂ ratios (low N₂/O₂) fish did not develop external signs characteristic of GBT but died quickly due to hemostasis from bubble formation in the blood and heart. At lower N₂/O₂ ratios fish lived longer but began to exhibit gross external signs of GBT, i.e. emphysema on skin and membrane surfaces. Carbon dioxide concentration showed no correlation to mean time to death for excess TGP over CO₂ ranges of 2 to 22 mg/L (Nebeker et al., 1976). GBT was induced in rainbow trout at TGP = 120 % (ΔPₐ = 138) at O₂ concentrations in excess of 200 % and N₂ concentrations essentially 100 % indicating that O₂ alone could induce GBT (Edsall and Smith, 1991).

Colt (1986) suggests that the potential for acute GBT increases dramatically when ΔPₐ > 76 mm Hg (TGP = 110 at BP = 760 mm Hg) and chronic effects may be observed at ΔPₐ as low as 38 mm Hg (TGP = 105 at BP = 760 mm Hg). These
values may correspond to atmospheric supersaturation when the $N_2/O_2$ partial pressure ratio = 3.77. Lower ratios would decrease effects and higher ratios may increase effects of GBT at TGB in excess of 100%. Larval fish may be more sensitive than adults. Larval striped bass exhibited GBT signs at total gas pressures as low as 103% (Cornacchia and Colt, 1984). More research on the role of the $N_2/O_2$ ratio producing GBT and effects of chronic levels of TGP are needed but elimination of potential for GBT can only be assured when $\Delta P_n < 0$ even when pure $O_2$ applications are utilized and $N_2/O_2$ ratios are low.

3.2. Gas Transfer Theory

The "two film theory" (Lewis and Whitman, 1924) states that interphase mass transfer of a chemically non-reactive gas into or out of a liquid is rate controlled by the combined resistances and driving forces established across a thin gas film and a thin liquid film extending from the interphase boundary into the respective medium. The film thickness may be a function of turbulence and mixing in the bulk mediums, but quiescent or laminar flow conditions exist within the films. Within these films gas movement is limited to molecular diffusion and thus represents the primary resistance to transfer. The driving force is the concentration gradient established across the film. While the thin film theory may be formally correct and adequately describes gas transfer behavior, other authors contend that in practice, establishment of true films seldom occurs and a more correct terminology maybe the "two resistance" theory (Bilstad and Lightfoot, 1985; Treybal, 1980).
The driving force and net direction of gas transfer is established by the magnitude and sign of the concentration gradient across the respective film. The concentration gradient are determined by the partial pressure difference of the gas between the gas and liquid phases. Generally, either the liquid or gas phase film resistance tends to be rate limiting to the extent that the other is considered insignificant. The liquid phase is rate limiting during transfer of slightly soluble gases such as O$_2$, N$_2$ and Ar, while the gas phase combination is rate limiting for highly soluble gases such as ammonia. For slightly soluble gases, the dissolved gas concentration at the gas/liquid interface is assumed to be the equilibrium or saturation concentration, $C_S$, in the liquid which is proportional to its partial pressure in the gas phase in accordance with Henry’s Law. Thus, the rate of gas transfer between the aqueous and gas phase is most conveniently described by the following equation.

$$\frac{dC}{dt} = k_L \frac{a}{V} (C_S - C)$$  \[7\]

where

- $\frac{dC}{dt}$ - time rate of change in the dissolved gas concentration, mg/(L-min)
- $k_L$ - liquid film resistance coefficient, m/min
- $a$ - interfacial area between the gas and liquid phases, m$^2$
- $V$ - liquid volume, m$^3$
- $C_S$ - equilibrium or saturation concentration of dissolved gas, mg/L
- $C$ - concentration of dissolved gas in the bulk liquid, mg/L
Due to the difficulty of obtaining accurate measurements of the interfacial area, $a$, the product of $k_L$ and $(a/V)$ are usually experimentally determined as a single parameter $K_L\cdot a$, and termed the overall transfer coefficient with unit of $\text{min}^{-1}$.

The value of $K_L\cdot a$ will vary considerably among different aeration process designs due to differences in the interfacial surface area generated and the rate of surface film renewal, both of which are direct functions of energy transfer from the aeration device to the liquid/gas system. For any single aeration device, the value of $K_L\cdot a$ is also strongly dependant upon test conditions and may vary considerably if test conditions change or are not representative of field operating conditions. These conditions include but are not limited to: water quality characteristics, water temperature, basin geometry and water circulation patterns. Thus, many authors maintain that "apparent" or observed values of $K_L\cdot a$ may be significantly different from true values even under well-controlled test conditions. Several researchers present techniques to correct these effects (Brown and Baillod, 1982; Stenstrom and Gilbert, 1981; Brown, 1978).

Due to the difficulties in providing representative information of aerator performance capabilities, the American Society of Civil Engineers has proposed a set of standards for determination of $K_L\cdot a$ and evaluation of aeration systems (ASCE, 1984). The recommended procedure to determine the transfer coefficient for an aeration device is to perform a clean water unsteady-state test in which deoxygenated water is reaerated by operation of the device in a well-mixed containment basin or tank. The resultant rise in DO concentration over time and the absolute time of measurement are recorded. Data collection begins at time $= 0$ when a DO concentration is first detected and continues until the DO approaches
within 98% of the estimated saturation concentration. Assuming the saturation concentration remains constant during the test, the data can then described by a form of Eq [7] with initial boundary conditions at \( t = 0 \) and \( C = C_0 \):

\[
C = C_s \left( 1 - e^{-KLa(t)} \right) + C_{0e}e^{-KLa(t)}
\]  

[8]

Nonlinear statistical methods can then be utilized to regress the concentration data against their time values and provide the best estimates of the transfer coefficient and the saturation concentration. Alternatively, the log-deficit method can be used in which the data are linearized by assuming a saturation concentration and taking the logarithm of the O₂ deficit, the difference between saturation concentration and the observed values. The log values are then linearly regressed against the time value of the observation. The slope of the resulting regression line is equal to the negative value of the transfer coefficient. The exponential method is the recommended method due to the potential for error in transfer coefficient estimation by the log-deficit method. A slight error in the saturation concentration estimation, truncation of data, or a departure from linearity in the log deficit data as the measured concentration approaches saturation can produce errors of 15 to 20% (Brown and Baillod, 1982; Brown, 1978). ASCE (1984) contains computer program listings both in BASIC and FORTRAN for parameter estimation using exponential methods.

Steady-state tests are utilized in applications where unsteady-state testing may be unsuitable. Many devices designed for intensive aquaculture utilize pure O₂ and achieve much higher O₂ transfer rates than conventional aeration devices such as surface agitators or bubble diffusers. These pure O₂ devices are often continuous flow components separate from the tank or basin. Water flow is circulated from
the tank through the oxygenation device and returned to the tank. Under steady-state conditions, if the O₂ saturation concentration within the aeration device is known and the liquid within the device is assumed completely mixed, the effluent DO concentration is representative of the DO concentration of the liquid within the device. The transfer coefficient can then be determined from the following mass balance equation (Watten, 1990).

\[ KL_a = \frac{OTR}{V(C_s - C_e)} \times 6 \times 10^{-5} = \frac{Q(C_e - C_{in})}{V(C_s - C_e)} \times 6 \times 10^{-5} \]  \[ [9] \]

where

- OTR - O₂ transfer rate, kg/hr
- Q - liquid flow through the device, L/min
- Cₑ - effluent DO concentration, mg/L
- C_{in} - influent DO concentration, mg/L
- V - liquid volume inside device into which O₂ transfer occurs, L

The observed value of the transfer coefficient is also affected by temperature of the test conditions and is corrected to a standard temperature of 20°C by the Arrhenius Equation:

\[ KL_a = KL_{a20} \theta T_c^{-20} \]  \[ [10] \]

where

- T_c - liquid temperature, °C.

The value of θ determined by various researcher ranges from 1.01 to 1.05 but the recommended value is 1.024 (ASCE, 1984; Stenstrom and Gilbert, 1981).
Water quality effects upon $K_L a$ are perhaps the most significant factor affecting the $O_2$ transfer coefficient for any particular aeration or oxygenation process and perhaps the most difficult to quantify. The $\alpha$ ratio attempts to quantify water quality effects on $K_L a$ under field operating conditions (fw) in relation to clean water test conditions (cw):

$$\alpha = \frac{K_L a_{fw}}{K_L a_{cw}}$$

[11]

Surface active contaminants such as oils, detergents and proteins commonly found in wastewaters and aquaculture waters appear to affect the rate of molecular diffusion through the liquid film. Other properties such as viscosity or surface tension may also be affected leading to changes in interfacial surface area or hydrodynamic behavior of the gas/liquid interface (Gilbert, 1979). Therefore, the effects of water quality for on $K_L a$ are not consistent and $\alpha$ may vary dramatically. Researchers, in repeated tests for evaluating $\alpha$ for a given waste water and aeration device, often obtain significantly different results (Barnhart, 1986). Ratios for different types of devices aerating the same water may range from 0.6 for to $> 1.0$ (Gilbert, 1979). Ratios typically range from 0.4 to 1.2 for municipal wastewaters (Doyle and Boyle, 1986) and from 0.7 to 1.1 for pond waters under extensive aquaculture production (Shelton and Boyd, 1983). While water quality effects on $O_2$ transfer performance may be significant enough to warrant qualification, attempts at quantification of anything more than trends in $\alpha$ are difficult.

The rate of $O_2$ transfer can be also be increased by increasing the $O_2$ deficit. Since minimum ambient DO concentrations may be established by biological criteria
within a system, the deficit is increased by increasing the saturation or equilibrium concentration. Henry's law states that the saturation concentration of a gas in a liquid is directly proportional to the partial pressure of the gas in the gas phase. Since the partial pressure of a gas in a multicomponent gas mixture is equal to its mole fraction times the total gas pressure, the saturation concentration is given by

$$C_s = \chi_i H (P_t - P_w)$$  \hspace{1cm} [12]

where

- $\chi_i$ - mole fraction of the $i^{th}$ gas in gas phase
- $H$ - Henry's constant, mg / (L-mm Hg)
- $P_t$ - total gas pressure, mm Hg
- $P_w$ - vapor pressure of liquid, mm Hg

Gas solubility is also dependant on liquid temperature and the Henry's constant is corrected for temperature by the Van't Hoff equation. The value of the Henry's constant for $O_2$ appears to be unaffected by gas temperature (Mueller, Quintana, and Di Toro, 1978).

The solubility is also dependant on salinity due to "salting out" effects and the Bunsen coefficient, $B_i$, is often used in lieu of the Henry's constant in another form of Henry's Law. The Bunsen coefficient accounts for the effects of temperature and salinity on the solubilities of $O_2$, $N_2$, $Ar$, (Weis, 1970) and $CO_2$ (Weis, 1974). Consequently, the saturation concentration can also be expressed as:

$$C_s = 1000 K_i B_i \chi_i \frac{(P_t - P_w)}{760}$$  \hspace{1cm} [13]
where

\[ B_i = \frac{L_{gas}}{L_{water, atm}} \]  

\[ K_i = \text{ratio of molecular weight to molecular volume of } i^{th} \text{ gas, mg/mL.} \]

Colt (1984) summarizes formulae for determination of vapor pressure; Bunsen coefficients; solubilities; and partial pressures of dissolved O₂, N₂, Ar, and CO₂. More recent research on O₂ solubility in water (Benson and Kraus, 1984) indicate that the methods presented by Weis (1970) for calculating atmospheric O₂ solubility may produce slightly incorrect values and new methods for calculation of atmospheric O₂ solubilities are presented. Any error attributed to determination of O₂ transfer rates, due to small errors in the saturation value produced by the methods summarized by Colt (1984), would probably be insignificant due to the uncertainty in estimation of other parameters such as \( K_i \).

The saturation value is also affected by contaminants, especially total dissolved solids, commonly found in wastewaters. The beta factor, quantifies the saturation concentration under measured under field conditions (fw) relative to clean water (cw):

\[ \beta = \frac{C_{s fw}}{C_{s cw}} \]  \[ [14] \]

Contaminant effects on beta are usually minor with the possible exception of waters with very high dissolved solids concentrations, such as industrial wastewaters. Values for industrial wastewaters may range from 0.77 to 0.97 while values for municipal wastewaters average about 0.95 (Strenstrom and Gilbert, 1981). Contaminant effects under most aquaculture situations would be
negligible for saturation values of O₂, N₂, Ar, and CO₂ corrected for pressure, salinity, and temperature.

ASCE (1984) proposes standardized definitions and terminology of performance parameters for rating aeration systems. The O₂ transfer rate, OTR, is the mass of O₂ per unit time dissolved in a volume of water under given conditions of temperature, barometric pressure, gas flow rate, power, and DO concentration. The standard O₂ transfer rate, SOTR, is the clean water OTR standardized to conditions of zero DO in the water, water temperature of 20° C and a barometric pressure of 1 atmosphere. The aeration efficiency, AE, is the OTR per unit power input to the device based on either delivered shaft power or wire drawn electrical power with the basis stated. The standard aeration efficiency, SAE, is the SOTR per unit total power input. The O₂ transfer efficiency, OTE, is the fraction of O₂ in an injected gas stream dissolved under given conditions while the SOTE is standardized to conditions as stated for SOTR. In much of the aquaculture literature, the O₂ transfer efficiency is referred to as the O₂ absorption efficiency or absorption efficiency and the aeration efficiency is referred to as the transfer efficiency. The proposed terminology (ASCE, 1984) will be used in this text.

3.3. U-tube Models

U-tubes operate by passing a two-phase mixture of water and bubbles through a deep U-shaped shaft, where the mixture is subjected to an increasing hydrostatic head (Fig. 3.1) as it moves down the u-tube. U-tube performance is a function of the water flow rate, inlet gas concentration, gas/water ratio, u-tube depth, off-gas recycle percentage, diffuser depth, and water quality. The gas/water or GW ratio is defined as the volumetric rate of gas (standardized to conditions of 1 atm and
21.1° C) injected in the u-tube inlet divided by the volumetric flow of water. The basic design of the u-tube aerators in operation at the VTAF is shown in Fig. 3.2.

3.3.1. Basic Model Algorithm

Several authors have developed models to predict u-tube performance (Watten and Beck, 1985; Speece, Eheart, and Giveler, 1983; Mitchell, 1973; Speece and Orosco, 1970). Inlet conditions and dissolved gas concentrations are usually assumed steady-state and only Watten and Beck (1985) attempted to model the effects of off-gas recycle on u-tube performance. Each of these models utilize forward finite difference analysis to simulate transfer within an element of gas and liquid traveling through the u-tube under plug flow conditions. The approximation of the differential equation, Eq [7], defining the rate of gas transfer within the element, is given by:

\[ \Delta C = K_L a (C_s - C) \Delta t. \]  \[15\]

The following discussion of the governing u-tube model equations and algorithm is based primarily upon the model developed by Watten and Beck (1985).

Gas transfer in the u-tube occurs under nonisobaric conditions, thus \( C_s \) and \( K_L a \) change as a function of depth, hydrostatic pressure, and gas transfer. A distance step for travel of an elemental plug is defined by an assumed differential time step, \( \Delta t \), and water flow velocity. At each distance step the following calculations occur (1) total pressure; (2) saturation concentrations of \( O_2 \) and \( N_2 \); (3) overall transfer coefficients; and (4) changes in dissolved gas concentrations of \( O_2 \) and
Fig. 3.1  Supersaturation of water with oxygen by u-tube aeration
Fig. 3.2  Design of u-tube aerators located at the Virginia Tech Aquaculture Facility
\( \text{N}_2 \) at the established deficit during the time step (Watten and Beck, 1985). Once
the change in concentrations are determined, a mass-balance of gases between the
gas and liquid phases establishes the gas phase mole fractions and new dissolved
gas concentrations for the next distance step. The calculations are repeated until
the plug exits the upflow leg of the u-tube. If off-gases are not recycled, exit
conditions and performance are summarized. If off-gases are recycled, a new
effective GW ratio is determined based upon the volume of gases recycled and
primary feed \( \text{O}_2/\text{water} \) ratio, PW. The PW ratio is the GW ratio established by the
pure \( \text{O}_2 \) volumetric rate injected from the primary \( \text{O}_2 \) source. This loop is
repeated until changes between the exit conditions of the previous loop and the
current loop are insignificant.

3.3.2. Modeling U-tube Performance

The gas transfer coefficient is a function of the gas bubble surface area. Mitchell
(1973) and Speece et al. (1970) observed that bubble size remains relatively
constant as the gas liquid mixture moves though the u-tube and only the numbers
of bubbles per unit liquid volume varies. Equations describing hydrodynamic
behavior in mass bubble turbulent flow indicate average bubble radius and the
critical bubble radius required for bubble breakup are relatively independent of gas
density within the bubbles or gas diffuser orifice size and are primarily controlled
by turbulent effects (Azbel, 1981). Assuming, under constant turbulent flow
conditions, that the average single bubble surface area is independent of
hydrostatic pressure or gas transfer and remains constant, then the change in the
total interfacial surface area resulting from gas transfer is directly proportional to
the change in number of bubbles and hence, total gas bubble volume \( V_g \). On the
basis of this assumption and the assumption that the film resistance coefficient, $k_L$, remains constant, the change in the transfer coefficient between consecutive distance steps 1 and 2 can be determined from the ratio presented by (Speece and Orosco, 1970).

\[
\frac{(K_La)_1}{(K_La)_2} = \frac{(V_g)_1}{(V_g)_2}
\]  \[16\]

An initial value of the transfer coefficient for conditions existing at the u-tube inlet is estimated from a regression equation generated from pilot performance data of the u-tube system. Initial estimates are obtained by iteration with the u-tube model to determine a value which predicts the observed exit DO concentration. By regression analysis, Speece and Orosco (1970) determined that the initial transfer coefficient is a function of the u-tube depth, flow velocity, and the initial GW ratio. As flow velocity increased, $K_La$ increased presumably due to increased turbulence, however, outlet DO remained relatively constant. Apparently, effects on performance due to any increase in the transfer coefficient were offset by a decrease in gas/water contact time. Watten and Beck (1985) developed an equation in which the transfer coefficient was a function of u-tube depth and the GW ratio for single downflow velocity. Due to the effects of varying pipe diameters and depths on the turbulent flow regime, each system design may have a unique equation governing estimation of the initial transfer coefficient.

The performance of a pure $O_2$ absorption system will be affected by the simultaneous stripping of other dissolved gases present in the treated flow concurrent with $O_2$ absorption. Dissolved $N_2$, Ar, and $CO_2$ are the principal gases which may be present. Desorption of these gases reduces the mole fraction of $O_2$
in the gas phase, reducing the concentration gradient by decreasing the O₂ saturation concentration.

Packed columns operating at barometric pressure or under a slight vacuum are often utilized to strip large amounts of dissolved N₂+Ar from air-supersaturated water used in fish culture to decrease total gas pressure. U-tubes operate at high pressures and are fair to poor gas strippers when compared to packed columns (Colt and Watten, 1988). However, N₂ and Ar, being slightly soluble gasses, are readily stripped from solution. This stripping should be considered when assessing u-tube performance. Carbon dioxide is a moderately soluble gas and requires very little buildup in the gas phase to terminate further stripping. In modeling multi-component gas transfer in packed columns, stripping of N₂ and Ar into the gas phase was shown to significantly reduce O₂ absorption efficiency and aeration efficiency. Carbon dioxide was shown to reduce the O₂ transfer efficiency no more than 7% even at dissolved concentrations in excess of 100 mg/L or approximately 15,000 % air-saturation (Watten et al., 1991).

The value of Kₐ for all gas species in a multi-component gas system need not each be determined experimentally. If the coefficient of one species, such as O₂, is established experimentally, the values of other species' coefficients can be estimated from their respective liquid phase diffusivities or molecular diameters. Einstein's Law of diffusion states that the liquid phase diffusivity of a solute species is inversely proportional to the size and shape of the diffusing molecule and the viscosity of the medium. Thus if the medium is unchanging, i.e., water in the liquid film, and the overall transfer coefficient for one species is known, transfer coefficients for other species are determined by the following ratio,
\[
\frac{(K_1a)_1}{(K_La)_2} = \frac{d_2}{d_1}
\]  

[17]

where, \(d\), is the molecular diameter of the respective gas species (Tsivoglou et al., 1965). The experimental results of Tsivoglou et al. (1965) support this relationship and each of the u-tube models utilized this convention for relating overall transfer coefficients. However, a model of another multicomponent system for waste treatment processes relates overall coefficients as:

\[
\frac{(K_1a)_1}{(K_La)_2} = \sqrt{\frac{D_1}{D_2}}
\]  

[18]

where \(D\) is the liquid phase diffusivity of the gas species (Mueller et al., 1973). This relationship is supported by the results of Bilstad and Lightfoot (1985) who cite an earlier work (Angelo and Lightfoot, 1968). The earlier work proposes that the liquid film resistance coefficient, \(k_L\), for gas transfer from submerged bubbles is proportional to the square root of the product of liquid phase diffusivity, frequency of bubble oscillation, and a constant, \(4/\pi\).

U-tube performance is quantified by determination of \(O_2\) transfer efficiency and aeration efficiency. To determine aeration efficiency, \(AE\), energy requirements to operate the u-tube system must be quantified. The total energy inputs associated with performance are the energy required to pump water and bubbles through the u-tube and the energy requirements for capture and recycle of off-gases.

The energy required to pump water is the sum of friction energy losses in single phase pipe flow and the energy losses attributed to two-phase flow (Watten and Beck, 1985). These losses are determined by the following equation.
\[
P_{wp} = \frac{Q \gamma_w (h_w + h_g)}{60000 \varepsilon_p}
\]

where

- \(P_{wp}\) - pump power requirements, kW
- \(Q\) - volumetric flowrate of water, L/min
- \(\gamma_w\) - unit weight of water, kN/m\(^3\)
- \(h_w\) - pipe friction headlosses due to liquid flow, m
- \(h_g\) - headlosses attributed to two-phase flow, m
- \(\varepsilon_p\) - overall or combined efficiency of pump

Pipe friction head losses are determined using standard procedures such as the Darcy-Weisbach or Hazen-Williams Formula. However, energy losses associated with two-phase flow are difficult to estimate without performance data. Speece and Orosco (1970) found that total head losses attributed to two-phase flow are a function of the GW ratio, water velocity, and gas diffuser position. For any flow velocity, there was a diffuser depth at which the headloss was zero regardless of GW ratio. Headlosses increased with increasing GW ratio at shallower depths, while at deeper depths, airlift effects resulted in a decrease in headloss as the GW ratio increased. Two-phase flow headlosses decreased as water velocity increased for a constant GW ratio and diffuser depth due to decreasing slip velocity between the bubbles and liquid within the downflow leg. In contrast, Mitchell (1973) found no correlation between downflow leg pressure loss and flow velocity for a 5.2 cm diameter or a 43.3 cm diameter downflow pipe under two-phase flow.
Theoretical derivations of drag forces and rise velocity of turbulent mass bubble flow in large still columns show that the drag forces are a function of bubble radius, rise velocity, and the void fraction of gas to water (Azbel, 1981). The terminal rise velocity occurs when the buoyancy forces and drag forces are equal and energy dissipation from the bubbles to the liquid is at a minimum (Azbel, 1981). Therefore, the decrease in headlosses at higher water velocities observed by Speece and Orosco (1970) could possibly be partially explained by a decrease in bubble slip velocity and/or average bubble radius due to increased turbulence.

A minimum downflow velocity must be established in order to achieve net bubble movement downward. Gas/water ratios near 0.20 may require downflow velocities of 0.9 to 1.2 meters per second (Speece and Orosco, 1970). A downflow velocity of 0.9 meters per second is commonly accepted as a minimum for reliable u-tube operation at GW ratios below 0.20. Wall effects and turbulent flow regimes may make theoretical estimation difficult and headlosses due to two-phase flow in u-tubes have been traditionally predicted from regression equations developed from performance data (Watten and Beck, 1985; Speece and Orosco, 1970).

Energy-losses associated with off-gas recycle include the headloss across the off-gas collector and the energy required to pressurize and reinject that gas back into the downflow leg. The off-gas collector provides sufficient detention time for phase separation and associated headlosses are a function of collector design. Watten and Beck (1985) estimate energy requirements associated with pressurization and injection of the off-gases through the use of the adiabatic compression formula as presented by Yunt (1978):
\[ P_{w_c} = \frac{mRT_k}{\eta_e} \left[ \left( \frac{P_2}{P_1} \right)^n - 1 \right] \]  

where

- \( P_{w_c} \) - power requirement of compression device, kW
- \( m \) - mass flow rate of gas, kg/sec
- \( R \) - gas constant for a particular gas, kJ/kgK
- \( T_k \) - absolute temperature, K
- \( k \) - ratio of specific heats, \( C_p / C_v \)
- \( n \) - constant = \((k-1)/k\)
- \( \eta_e \) - overall or combined efficiency of the compressor
- \( P_2 \) - absolute outlet pressure, kPa
- \( P_1 \) - absolute inlet pressure, kPa

The off-gases contain variable concentrations of \( O_2, N_2 + Ar \), and water vapor with possibly small amounts of \( CO_2 \) and consequently values for \( R, n, \) and \( k \) may vary slightly with changing operating conditions. Watten and Beck (1985) assume the gas to be pure air with values of \( R = 0.287 \), \( n = 0.283 \) and \( k = 1.395 \). The outlet pressure of the compression device, \( P_2 \), is the absolute pressure existing at the u-tube inlet and is the sum of the barometric pressure and available water head. The inlet pressure, \( P_1 \), is calculated as the outlet pressure, \( P_2 \), minus the sum of friction, two-phase flow and off-gas collector head losses.

The energy to move or pump the flow through the u-tube, \( P_{w_p} \), is determined from the friction and two-phase flow head losses in the Watten and Beck (1985) model. Alternatively the real wire-drawn power of a pump could be assumed for
Pw, if true total power requirements are desired. The aeration efficiency (kg O₂ / kWhr) is then calculated by the following equation.

\[
AE = \frac{\text{OTR}}{60000(Pw_p + Pw_c)} \tag{21}
\]

3.3.3. Model Results

Watten and Beck (1985) utilized pilot performance data to calibrate and subsequently verify their u-tube model. During validation tests, performance data were obtained for recycle rates of 0, 25, 50, and 60 % and initial influent O₂/liquid ratios of 2.3 to 6.5 %. Relative error of model predictions of effluent DO ranged from 0.9 to 6.3 % and averaged 2.9 %.

The Watten and Beck model revealed that significant improvements in both OTE and AE could be achieved when off-gases were recycled. For a given u-tube depth and inlet DO concentration, the degree of improvement depended largely on both the PW ratio and the percentage recycle of off-gases.

As the PW ratio is increased, the benefits of off-gas recycle decrease. For example, model simulations indicated an increase in OTE from approximately 42 % at 0 % off-gas recycle to 54 % at 50 % for u-tube operation at a PW ratio of 0.023. Operation at a PW ratio of 0.062, OTE were approximately 27 % at 0 % recycle to 33 % at 50 recycle. While performance is dependent on site specific conditions, Watten and Beck (1985) summarize that for a fixed rate of off-gas recycle, OTE will increase with greater shaft depths, lower PW ratios, and lower influent DO values.
Model simulations revealed u-tube operation at low PW ratios and recycle rates less than approximately 20% resulted in a decrease in AE (Watten and Beck, 1985). The aeration efficiency increases as the recycle rate is increased above approximately 20% and a recycle rate exists that will maximize the AE. Off-gas recycle at high PW ratios can result in a decrease in AE and may result in effective or actual GW ratios that are large enough to cause system failure. Failure is caused by coalescence of bubbles to form large slugs of gas in a transition from a bubbly to slug flow regime. The net effect is excessive head loss and unpredictable flow behavior.

Watten and Beck (1985) showed that for a recycle rate of 90%, total operational costs per kg O₂ transferred were reduced approximately 25 - 41% over a range of required effluent DO concentrations of 20 - 40 mg/L. The increase in total savings increased with increasing effluent DO indicating energy costs per mass of O₂ transferred decreases with increasing GW ratio. Speece et al. (1980; 1983) reported that transfer costs are minimized at relatively large GW ratios of 15-25% and u-tube depths of 25 to 60 meters. Model simulations by Watten and Beck (1985) indicate optimum u-tube operation may best be achieved when designed in a manner to satisfy O₂ demands while operating at relatively high GW ratios established by high off-gas recycle rates.

3.4. Pure Oxygen Aeration Systems

Colt and Watten (1989) and Watten (1991) describe at least six different types of pure O₂ absorption systems that are commercially available and rate typical performance in terms of OTE and AE. A brief review of design and performance
information on pure $O_2$ devices taken from Colt and Watten (1989) and Watten (1991) is presented.

**U-tube:** U-tube systems, originally developed for aeration of large volumetric flows of wastewater (Bruijn and Truinzaad, 1958), utilize hydrostatic pressure to produce a rapid increase in the $O_2$ deficit. While the earliest versions used in wastewater aeration and in fish hatcheries (Speece, 1969) utilized air as the $O_2$ source, limitations on total dissolved gas pressure precludes the use of air in deep u-tubes for high density systems and require $O_2$ rich gas. U-tube systems without off-gas recycle obtain OTE ranging from 30 to 50 % and AR ranging from 1 to 1.5 kg $O_2$/kW.hr for u-tube systems without off-gas recycle. Oxygen transfer efficiency may be increased to 60 to 90 % and AE increased 2.0 to 3.0 kg $O_2$/kW.hr with capture and recycle of off-gases.

**Packed Columns:** Enclosed packed columns utilize an internal packed media which generate increased surface area and some turbulence as water flows down through the column. Oxygen rich gas is the continuous throughout the column as flows are not high enough to saturate the column with water. These devices may be designed to operate at a partial vacuum to enhance $N_2$ gas stripping or at positive pressures to further increase $O_2$ absorption rates. Oxygen transfer efficiencies of 40 to 55 % are typical and may be as high as 95-100 % for pressurized devices. Aeration efficiency ranges from 0.5 to 1.0 kg $O_2$/kW.hr.

**Spray Columns:** Enclosed spray columns create a fine spray of water in contact with an $O_2$ rich atmosphere. These devices are used when the particulate or nutrient content of the water may significantly reduce the void fraction of the media within packed columns due to clogging or biogrowth. These devices may
also be operated at a partial vacuum or pressurized. Performance is similar to packed columns.

**Down Flow Bubble Contactor:** Extended or unlimited gas to water contact time is the principal advantage of this device. Water is pumped through an inverted cone and O₂ is injected in the narrow diameter zone near the top where water velocity and turbulence are high and rapid transfer occurs. As bubbles move down the expanding cone, the downward velocity of the bubbles decreases to zero and the bubbles remain suspended within the device. A small portion of the gases may be wasted to enhance stripping of N₂+Ar and CO₂ into the bubbles. Oxygen transfer efficiency may approach 80 to 90 % and AE can approach 4.0 kg O₂/kWhr.

**Oxygen Injection or Sidestream Oxygen Injection:** In oxygen injection systems, high operating pressure and turbulence achieve rapid O₂ transfer and highly supersaturated effluents. This reduces the volume of water which must be treated. In operation, a stream of water under high pressure (190-890 kPa) is pumped through a venturi nozzle receiving gaseous O₂ injection. High turbulence and shear produces a suspension of very fine bubbles inducing rapid gas transfer. The mixture contact loop provides a contact time 6 to 12 sec to promote further transfer. The mixture is then passed through an eductor or expansion nozzle into the receiving water where a large volume of water is entrained and mixed with the sidestream flow resulting in further transfer from the bubbles. Oxygen transfer efficiency ranges from 15 to 70 % while AE is typically less than 0.5 kg O₂/kWhr.

**Multi-staged Low head Oxygenation (LHO):** The LHO employs the concept of serial off-gas reuse and parallel water flow through several separate treatment stages to maximize O₂ transfer efficiency. These stages are generally designed as
packed or spray column segments. Within single stage devices, high transfer rates are established initially leading to a rapid reduction in the O₂ deficit due to rising DO in the liquid and a simultaneous reduction in the saturation concentration. The transfer rate eventually drops to a point where further contact may not be beneficial even though a significant percentage of the injected O₂ remains in the gas phase. By continually separating and re-exposing off-gases to water with low DO, maximum utilization of the O₂ deficit is obtained as the gaseous O₂ mole fraction decreases. Depending upon design and operation, Oxygen transfer efficiency ranges from 70-100 % and AE ranges from 2 to 5.5 kg O₂/kWhr.

**Enclosed surface agitation:** A surface agitator powered by a submersible motor is enclosed within a sealed hood or chamber to expose a water stream to O₂ gas. Unlike the other types of contactors, no significant hydraulic gradient is required. Several units may be staged in the same manner as the LHO to facilitate improved transfer efficiency (Watten and Boyd, 1990). Oxygen transfer efficiency typically ranges from 30-70 % and AE from 0.25-1.0 kg O₂/kWhr.
4. MODEL DEVELOPMENT

The equations and algorithm representing the mass-balance between gas transfer and respirational processes occurring in the RAS located at the Virginia Tech Aquaculture Facility are presented in this chapter. Figure 1.1 shows the RAS components included in the model. The fundamental thesis of the model is that the combined effects of these processes occurring within all components ultimately dictate equilibrium dissolved gas concentrations within each component. Gas transfer and/or respiration changes the concentration of one or more dissolved gas species of a component's water inflow and therefore defines the effluent concentration. The effluent from that component represents the influent to the next component. The process is repeated from component to component until the water returns to the original component again. Thus, the RAS can be modeled as a continuously looping flow circuit containing elements or "component submodels" as shown in Fig. 4.1.

The model was developed in three stages. An overall algorithm for the computer model of the RAS was developed in Stage 1. The mathematical models of individual components were developed in Stage 2. Procedures to facilitate user input, final results analysis, and output were developed in Stage 3. The computer language utilized was Microsoft®QuickBASIC, Version 4.5.
FIG 4.1  Diagram of RAS computer model with RAS component "submodels" arranged in flow circuit to represent RAS configuration.
4.1. Stage 1

Before a model simulation can begin, initial and operating conditions must be defined through user input. Then a model simulation can be initiated and continues until equilibrium conditions are achieved. The program is menu-driven and the main menu allows the user three choices; 1) update or change input parameters to define initial and operating conditions for the simulation run; 2) begin a simulation; or 3) exit the program and end the current program session.

The initial conditions, *i.e.*, dissolved gas concentrations for the very first simulation run in any program session are defined to be the saturation concentrations of the dissolved gases at atmospheric conditions and water temperature. However, the user has the option to change any of the initial dissolved gas concentrations in the tank component prior to a simulation run. The initial conditions in subsequent simulations during any program session are defined by the results of the previous simulation, altered only by changes in operating or system design parameters instituted by the user. Thus, the user can simulate a response to a change in one or more input parameters, such as $O_2$ demand in the tank, flowrate, or GW ratio in the u-tube.

Once steady-state equilibrium conditions are established, the program exits the simulation loop and output procedures are called to analyze effluent water quality of the RAS components. Equilibrium conditions are achieved when the gas concentrations in all of the components stabilize resulting in steady-state effluent concentrations. Steady-state conditions occur when the mass balance ratio of each
gas species, computed as the sum of all the sources (inputs) into the RAS divided by the sum of all sinks (outputs) out the RAS, is greater than 0.97. In any simulation, the flow circuit loop is executed a minimum of 10 times. Upon conclusion of any simulation, the equilibrium DO concentrations in each component are printed to the screen and a second menu presents the following options: 1) summarize performance data and send output summary to the printer; 2) write the progression of tank DO concentration to an comma delimited ASCII File; or 3) return to the Main Menu. The option to write the DO data to an ASCII file is provided so the user can import the data into a plotting program or spreadsheet.

The flow circuit simulation loop contains the CALL statements for procedures representing four components of the RAS; RBC, u-tube, tank, and clarifier. The only parameters passed from one component submodel to the next during loop execution are the respective dissolved concentrations of $O_2$, $N_2+Ar$, and $CO_2$. Execution of a single simulation loop entails execution of a component submodel for one model time step, advancing to the next component and serially moving around the circuit, eventually returning to the original component. Another loop is executed if equilibrium criteria are not satisfied. The model time step is equal to one u-tube detention time since the u-tube is modeled as plug-flow and has the smallest detention time of any of the loop components. As long as all other components are modeled as completely-mixed stirred tank reactors, setting the time step in this manner is reasonable. Inlet conditions remain constant during any given time step and are equal to the exit conditions of the upstream component. The values of the dissolved concentrations of $O_2$, $N_2+Ar$, and $CO_2$
existing each component are stored in their respective arrays during each loop execution.

A modular programming style was employed by creating separate procedures for each specific task, thereby keep the main program code simple. All component submodels are procedures which are called from the main flow circuit which is developed as part of the main program body (Fig 4.1). This modular style of development facilitates incorporation and/or removal of other process components with minimal alteration of the main program. Once a procedure is written to simulate a new component, a CALL statement for that component is placed at the proper location in the loop. Likewise CALL statements and array elements for components to be removed need only be deleted from the simulation loop code. Appendix B contains a complete code listing of the main program and associated procedures.

4.2. Stage 2

Each of the process components except the u-tube were modeled as completely mixed stirred tank reactors. The u-tube model was taken from Watten and Beck (1985) and is simulated with a plug flow reactor model. Gas transfer, consumption, and production are simulated using a finite difference method and mass balance analysis. During execution of any flow circuit loop, each component submodel is also executed for one model time step before moving to the next component. During component loop execution, the change in dissolved gas concentrations over a finite elemental time step, $\Delta t$, are determined by multiplying the sum of mass flow rates of all sources and sinks of each gas by $\Delta t$.
and dividing the sum by the component liquid volume. The change is then added to the current concentration to determine the initial concentration for the next $\Delta t$ loop. The loop is repeated until the summation of $\Delta t$ exceeds the component time step. The analysis then moves to the next component. The finite elemental time step, $\Delta t$, is set equal to 0.25 sec according to the recommendation of Watten and Beck (1985). The general submodel mass balance equation for determining the incremental change in the concentration of any dissolved gas within the completely mixed components is

$$\Delta C = \frac{(Q \ C_{in}) + (K_L a \ (C_s - C) \ V) - DEM + PROD - (QC)}{V} \ \frac{\Delta t}{60} \ \ [22]$$

where

$\Delta C$ - change in dissolved gas concentration, mg/L

$Q$ - water flow rate, L/min

$C_{in}$ - influent dissolved gas concentration, mg/L

$K_L a \ (C_s - C) / V$ - transfer due to atmospheric aeration, mg/min

$K_L a$ - overall transfer coefficient for respective gas, min$^{-1}$

$C_s$ - saturation concentration, mg/L

$C$ - current (effluent) dissolved gas concentration, mg/L

DEM - mass demand rate due to respiration, mg/min (O$_2$ only)

PROD - mass production rate due to respiration, mg/min (CO$_2$ only)

$\Delta t$ - finite difference time step = 0.25 sec

$V$ - component volume, L.

Obviously, only O$_2$ and CO$_2$ are involved with respiration and the terms DEM and PROD are included only in the respective equations for the tank and clarifier. The model is assumed to be a closed system in that the dissolved CO$_2$ partial pressure is not assumed to remain in equilibrium with its atmospheric partial pressure. The
CO₂ (CO₂ + H₂CO₃) concentration is assumed to be dependant upon gas transfer, biological production and carbonate equilibria criteria for a closed system. The final pH is determined from the alkalinity concentration as CaCO₃ and the CO₂ concentration upon termination of the simulation loop. A general description of each component submodel follows.

4.2.1. U-tube

The u-tube model of Wattan and Beck, (1985) was utilized to simulate the u-tube aeration system at the VTAF. Some alterations of the original u-tube model were required to facilitate incorporation into the overall RAS model. First, all user input, performance analysis, and summary output code were removed and recoded into the input subroutine. Second, the nitrogen concentration and its associated equations for establishing stripping by transfer were redefined to represent the "N₂+Ar" concentration. Third, when considering the effects of off-gas recycle, the original u-tube model executed as an iterative loop until equilibrium conditions in the effluent and recycle gases were achieved. This internal loop was modified to conform with the component process procedure described for the other component process modules such that the loop is executed for one component time step during each flow circuit loop. The flow circuit loop will execute a minimum of 10 times to assure adequate assessment of the effects of off-gas recycle. Fourth, regression equations for estimation of initial Kₐ a and two-phase flow head loss for the u-tube design at the VTAF were developed from performance data.
4.2.2. \textit{Tank}

The production tank is modeled as a completely mixed unit. Separate equations are developed from Eq [22] for $O_2$, $N_2$+Ar, and $CO_2$. Atmospheric aeration is included to account for transfer due to in-tank aeration such as supplemental or emergency aeration. The equations assume that any gas transfer occurs between the liquid and the atmosphere at user supplied conditions of barometric pressure and water temperature. Only one term for atmospheric transfer, $K_La(C_s-C)/V$, is included as shown in Eqs [23]-[25]. The $O_2$ transfer coefficient is user input and therefore if more than one aerator is to be simulated, the user must enter an overall or "apparent" value. Values of transfer coefficients for $N_2$+Ar and $CO_2$ are determined from the molecular diameter or liquid diffusivity ratios with respect to $O_2$ as given by Eqs [17] and [18]. Values for these ratios may be entered by the user during execution of the input procedure or the model uses default values of 0.94 for $O_2/(N_2$+Ar) and 1.0 for $O_2/CO_2$. The respective equations developed from Eq [22] are

\begin{align*}
\Delta C_{O_2} &= \frac{(Q C_{in}) + (K_L a (C_s - C) V_t) - DEM_t - (QC)}{V_t} \Delta t \quad [23] \\
\Delta C_{N_2+Ar} &= \frac{(Q C_{in}) + (K_L a (C_s - C) V_t) - (QC)}{V_t} \Delta t \quad [24] \\
\Delta C_{CO_2} &= \frac{(Q C_{in}) + (K_L a (C_s - C) V_t) + PROD_t - (QC)}{V_t} \Delta t \quad [25]
\end{align*}

where
$V_t$ - tank water volume, L

$DEM_t$ - tank $O_2$ demand, mg/min

$PROD_t$ - tank $CO_2$ production rate, mg/min.

The values of $C, C_s, C_m$, and $K_{La}$ for Eqs [23]-[25] are representative of only the indicated dissolved gas species and cannot be assumed to be identical for each equation.

The average daily $O_2$ demand term, $DEM_t$, can be input by the user, or the program can determine a value for $DEM_t$ based upon the $O_2$ feed ratio, OFR, and the daily feed ration. The value for OFR may be input or a default value of 0.3 is assumed. The $O_2$ demand can also be adjusted by a user input scaling factor to account for daily maximum and daily minimum demands. The default value for the scaling factor is 1.0.

The production rate for $CO_2$ is determined from the $O_2$ demand and OFR. Colt and Orwicz (1991) cite a production ratio of 0.34 kg $CO_2$ per kg feed. Based upon the recommended value of OFR $= 0.25$, the ratio of $CO_2$ production rate to $O_2$ consumption rate at a given feeding level would be 1.36. Therefore, the $CO_2$ production rate, $PROD_t$, is $1.36 \cdot DEM_t$. The scaling factor applied to $DEM_t$ is also applied to $PROD_t$.

4.2.3. Multi-tube Clarifier

The multi-tube clarifier is modeled as a completely mixed stirred tank reactor even though flow through the tube media would be more properly modeled as plug flow. Approximately one-third of the total clarifier water volume is occupied by the media within a center horizontal cross-section of the unit. The remaining two
thirds is open water space with one-third located beneath the media where water enters the unit and moves upward into the media. The final third lies between the top of the media and the water surface. A pump near the top of the clarifier delivers flow to the RBC. Water flow across these sections, while not plug flow or completely mixed, approaches some intermediate state. Assuming the clarifier to be a completely mixed reactor simplifies the model by allowing a single component time step to be used for all components. The time step is taken from the u-tube which has the shortest detention time and consequently the shortest time step.

Eq [22] was developed for O$_2$ and CO$_2$ only. No transfer between the gas and liquid phase is assumed to occur in the clarifier. The incremental change in gas concentration is:

\[
\Delta C_{O_2} = \frac{\left( Q \, C_{in} \right) - \text{DEM}_{mc} - (QC)}{V_{mc}} \frac{\Delta t}{60} \tag{26}
\]

\[
\Delta C_{CO_2} = \frac{\left( Q \, C_{in} \right) + \text{PROD}_{mc} - (QC)}{V_{mc}} \frac{\Delta t}{60} \tag{27}
\]

where

- $V_{mc}$ - multi-tube clarifier water volume, L
- DEM$_{mc}$ - multi-tube clarifier O$_2$ demand, mg/min
- PROD$_{mc}$ - multi-tube clarifier CO$_2$ production rate, mg/min.

The DEM$_{mc}$ term for O$_2$ consumption in the clarifier is user input or the model estimates a demand based on the daily feed ration. A regression equation for estimating an overall clarifier O$_2$ demand, DEM$_{mc}$, was developed from O$_2
consumption data measured in the clarifier during actual production conditions during the 1991 production trials. The PROD \text{mc} \text{ term for CO}_2 \text{ production in the clarifier is estimated as described for CO}_2 \text{ production in the tank.}

4.2.4. Rotating Biological Contactor

The RBCs located at the VTAF were constructed as a three stage unit with each stage separated by an internal baffle or partition (Fig. 1.1). The RBC component is modeled as three completely mixed units in series; however, the RBC model has the flexibility to permit simulation of a 0, 1, 2 or 3 stage RBC. One form of Eq [22] gives the incremental change in any of the dissolved gas concentrations:

\[ \Delta C = \frac{(Q \cdot C_{in}) + (K_{l,a} \cdot (C_g - C) \cdot V_r) - (QC)}{V_r} \cdot \frac{\Delta t}{60} \]  

[28]

where

\[ V_r \] 

RBC water volume, L

but the equation is provided separately for O\text{2}, N\text{2}+Ar, and CO\text{2}. The component volume, \[ V_r \], is assumed to be the water volume of an individual stage rather than the total water volume of the entire RBC unit and the volume in each stage are assumed to be equal. The component time step, as determined by the u-tube detention, time applies to the entire RBC unit and not each individual stage.

Effects of biological respiration within the RBC were considered insignificant. Therefore, the DEM term for O\text{2} and the PROD term for CO\text{2} are not included in the RBC transfer equations.

The user has the option to enter a transfer coefficient, \[ K_{l,a} \], for each stage or allow the model to estimate the values based on performance data of the VTAF RBC. A
regression equation to estimate an overall or "apparent" O₂ Kₐ for the first stage was developed from performance data taken from the RBCs at the VTAF. The regression equation is a function of water flow rate due to turbulence and bubble entrainment from the pipe delivering water from the clarifier onto the water surface in the first stage of the RBC. The value of Kₐ for the second and third RBC stages is assumed constant and was also determined from performance data. Values of diffusivity ratios entered for the tank are used to determine Kₐ for N₂+Ar and CO₂ for the RBC from the value of Kₐ for O₂.

4.3. STAGE 3

The third and final stage of model development involved the development of the input, output, and results analysis procedures. Procedures were developed for calculating many background parameters which remain constant during simulation execution.

The input subroutine allows the user to input, update, or a change input parameters describing the design and operation of the RAS and is called from the main program menu if the user desires to change an input parameter upon completion of a simulation. The input subroutine is also called at the beginning of a program session to allow the user to input design and operating data and specify initial conditions before beginning the first simulation run.

Procedures are called to analyze the equilibrium effluent data (dissolved gas concentrations) for each component once equilibrium criteria are met for the current simulation. The routines analyze and summarize component performance.
A procedure is provided for determination of total gas pressure, total gas pressure as percent of saturation, individual partial pressures of dissolved gases, percent saturation of the individual dissolved gases, and the partial pressure ratio of \( \text{N}_2 + \text{Ar} / \text{O}_2 \).

A procedure for analyzing RBC and u-tube transfer performance was developed. The transfer rate of \( \text{O}_2 \), \( \text{N}_2 + \text{Ar} \), \( \text{CO}_2 \) is determined for the RBC while only the transfer rate for \( \text{O}_2 \) and \( \text{N}_2 + \text{Ar} \) are determined for the u-tube. The effects of \( \text{CO}_2 \) on u-tube performance were considered insignificant. The transfer rate of any gas at effluent concentration \( \text{C}_e \) and influent concentration \( \text{C}_{\text{in}} \), through any device is given by

\[
\text{TR} = Q ( \text{C}_e - \text{C}_{\text{in}} ).
\]  

[29]

The aeration efficiency, \( \text{AE} \), for \( \text{O}_2 \) and \( \text{N}_2 + \text{Ar} \), are determined for the u-tube from Eqs [19]-[21]. The oxygen transfer efficiency, \( \text{OTE} \), of the u-tube is also determined based upon the OTR and mass rate of pure \( \text{O}_2 \) injected as determined by the PW ratio. The Total transfer costs for \( \text{O}_2 \) are determined from the \( \text{O}_2 \) injection rate multiplied by the user supplied value of cost of \( \text{O}_2 \) (\$/kg) plus the total power requirements multiplied the user supplied value for power cost (\$/kWhr). The \( \text{O}_2 \) utilization efficiency, \( \text{UE} \), is determined and is defined as the total system biological \( \text{O}_2 \) consumption rate divided by the mass rate of \( \text{O}_2 \) injected into the u-tube.

\[
\text{UE} = \frac{\text{DEM}_{\text{i}} + \text{DEM}_{\text{nc}}}{M_{\text{O}_2}}
\]  

[30]
The procedures to write array data of the $O_2$ concentration in the tank or print a summary report are called only if the user chooses the option at the prompt. The data is written in comma delimited ASCII format to a user named file.
5. MATERIALS AND METHODS

Two sets of tests were conducted to provide sufficient data for model calibration and validation. During the first set of tests, data were collected for model calibration and development of regression equations for predicting the respective component O$_2$ transfer coefficients, $K_{L_a}$, in Eq [22]. Tests were designed to provide the required O$_2$ transfer data for the emergency aerators in the production tank, the RBC, and the u-tube. All tests were conducted using tap water to establish clean water values and all values established for the transfer coefficient were corrected to a standard temperature of 20°C before regression analysis.

The second set of tests was intended to provide overall system performance data for model validation. Tests were designed to simulate scenarios of foodfish production in recirculating aquaculture systems. Artificial O$_2$ demands were established in both the tank and clarifier at rates similar to those observed during the 1991 production trials. Oxygen was provided to the system by u-tube aeration. All tests were conducted in clean tap water to minimize variability in O$_2$ transfer data. Effluent equilibrium DO concentrations of all components, as well as the change in dissolved gas pressure, $\Delta P_m$, in the tank were collected for various simulated production scenarios.

Prior to data collection, the required instrumentation was calibrated. Water flow rate and DO were monitored with commercially available instrumentation. These signals were sampled and recorded by a data logging and control module. Upon conclusion of any experiment or test, the data were dumped through a coaxial network to a central computer connected to multiple modules located throughout the research facility.
Dissolved oxygen concentration data in all components of several recirculating systems were taken during the 1991 production trials. The data were intended to provide typical performance characteristics of a recirculating aquaculture system under field production conditions. U-tube performance was established for field conditions for comparison to the clean water tests. In addition, actual biological O₂ demands were estimated for both the tank and multi-tube clarifier.

5.1. Instrumentation

Water flow rates were measured with a paddle-wheel type flow transmitter, Omega Instruments Model 58, connected to a rate analyzer, Omega Instruments Model DPF 64. The flow transmitter was calibrated over a range of 175 to 350 L/min. Prediction of five flow rates not used in transmitter calibration was ±1.5 % (TABLE A2, p. 160). A brief description of the calibration procedure is also provided (Appendix A, p. 159).

Dissolved oxygen was measured with two commercial DO monitors and a commercial handheld meter. The continuous monitoring systems were the Royce Model 9010 Dissolved Oxygen Analyzer mated with a Royce Model 90 DO Probe and an Oxyguard® 4 Multichannel Dissolved Oxygen Monitor. Manufacturers specifications state an accuracy ±0.5 % of full-scale reading for the Royce and ±1.0 % of full-scale for the Oxyguard when each system is properly calibrated. The handheld DO meter was an Yellow Springs Instrument (YSI) Model 58 connected to a YSI BOD probe. DO measurements made with the YSI were manually recorded. The meter specifications stated an accuracy of ±0.03 mg/L.
Oxygen gas flow was measured with two commercial rotameters. Low to mid range flows (0.5-7 SLPM) were measured with a Matheson Model 757 rotameter. Larger flows (7-15 SLPM) were measured with an Omega Instruments Model FL-115. Custom calibration tables for the Matheson rotameter were not available. Therefore, the meter was calibrated for the range of flows expected to be used during testing (See the Appendix A, p. 155 for a description of the calibration procedure and the resulting calibration table). Measured calibration accuracy was approximately ±5%. A custom calibration was obtained with the Omega Instruments Rotameter with a manufacturers specified accuracy of ±2%. Testing of the rotameter at several flow rates by the calibration method utilized for the Matheson Model 757 yielded results which agreed with the calibration tables within 1%.

Changes in dissolved gas pressure, ΔP_m, were measured with a commercially available Weis Saturometer, and with a continuous monitoring device termed the "gasometer". The gasometer is not available commercially. The gasometers used in this research were constructed in the Department of Agricultural Engineering Shop at Virginia Tech. Tests conducted to evaluate performance revealed agreement in measurement of ΔP_m between gasometers of ±5% while agreement with the saturometer was approximately ±10%. The gasometers substantially reduced the time required to obtain a representative reading of ΔP_m. A 95% of equilibrium reading was obtained by one gasometer within 5 to 6 minutes while the saturometer required over 15 minutes to reach a 95% of equilibrium reading (Fig A2, p. 163). Appendix A, p. 161 contains a more detailed description of the gasometer and its advantages over the saturometer.
5.2. Test Set 1 - Component Calibration

5.2.1. Tank

Four reaeration tests were conducted to evaluate the effect of recirculation flow rate on the $O_2$ transfer coefficient of the emergency aerators located in the production tanks. Unsteady state tests for aerator performance evaluations were conducted in accordance with guidelines as presented in ASCE (1984) to determine the value of the "apparent" $K_t a$. Test procedures for each of the four tests were identical with the exception of varying the recirculation flow rate from 220 L/min to 360 L/min.

Recirculation flow rates were established by pumping water from the end of the tank where water exits into the clarifier directly into the u-tube distribution line entrance that distributes u-tube effluent flow into the tank. Dissolved oxygen concentrations were measured and recorded at two of the tank distribution line inlets. Tank water volumes of approximately 8900 L were deoxygenated with 11.8 mg/L sodium sulfite per mg-DO/L DO to insure complete mixing of excess sulfite before a measurable rise in DO was recorded. Cobalt chloride catalyst concentrations of approximately 0.3 mg/L were established before the addition of sodium sulfite. The DO concentration was sampled and recorded at 1 minute intervals until the DO concentration had risen from non-detectable levels to within 5% of the apparent steady-state saturation value. The data were then retrieved from the logger by the computer and dumped to diskette. A fifth unsteady-state test was conducted without the emergency aerators operating to evaluate the effects of $O_2$ transfer through a quiescent water surface. All test data were then analyzed to determine $K_t a$ by the log-deficit method.
5.2.2. **Rotating Biological Contactor**

Steady state tests were devised to evaluate the effects of flow rate on the value of the $O_2$ $K_t$ $a$ for each stage of the rotating biological contactor. Only the effects of flow rate on gas transfer were considered when designing the tests. The following operational parameters were held constant and represent normal operation during production trials: rotational speed, 6 RPM; RBC stage water level i.e. fraction disk of disk submergence, approximately 30%; aerated water volume, approximately 650 L; and media disk diameter and media surface area per disk, 1.97 m and 336 m$^2$ respectively. Flow rate was varied from 140 to 340 L/min.

Water flows were supersaturated with DO by pumping air-saturated water through the u-tube and injecting $O_2$ to establish a steady-state effluent DO of 150 to 200% of the air saturation value. The flow was discharged into the first stage of the RBC. Steady-state supersaturated inflows were maintained for 8 to 10 stage detention times. The DO concentrations were measured with the YSI Portable DO meter and BOD probe. Two sets of six samples each were collected during each test in 300 mL BOD bottles. Three water samples for each set were taken from the u-tube discharge (RBC inlet) and three samples were taken from a sample siphon positioned at the stage exit for each set. The first set was collected at four stage detention times and the second at 6 to 7 detention times to assure that steady-state influent and effluent conditions were achieved.

Twelve tests were conducted to develop the regression equation for prediction of the 1st stage transfer coefficient and 5 test runs were made to validate prediction of the RBC component model. Gas transfer across the 1st stage occurs due to the RBC aeration effects and due to the inlet water discharge delivered by pumps from
the clarifier. Water is discharged above the water surface downward into the RBC stage. The resulting entrainment of air bubbles and degree of localized transfer was therefore assumed to be a function of the energy imparted to the water due to flow momentum. Four tests were conducted with water discharge just below the water surface to provide a comparison to above surface discharge values. Water was discharged from a 5.08 cm inlet pipe resulting in flow velocities ranging from 1.1 to 2.8 m/s.

The value of 2nd and 3rd stage $K_{L}{\alpha}$ were assumed to be essentially equal since water enters each of these stages at the bottom of the stage and is directed upward toward the surface by a baffle to facilitate mixing. These tests were also conducted on the first stage; however, the supersaturated water was discharged from a horizontal distribution pipe positioned at the bottom of the RBC stage on the inlet side. The water was directed upward toward the surface exiting from a series of 1.27 cm holes drilled into the pipe capped on the end. Three tests were conducted for 2nd and 3rd stage evaluation.

Two separate tests were conducted for RBC component model validation by measuring transfer across the entire 3-stage RBC. Tests were conducted as described for individual stage calibration. Dissolved oxygen measurements were made at the u-tube outlet (RBC inlet), 1st stage exit, and RBC unit exit during one test. A set of samples were also collected at the 2nd stage exit during the 2nd test in addition to those taken for the 1st test. One set of samples at each location was collected at approximately 8 stage detention times.
5.2.3. **U-tube**

U-tube submodel calibration tests were conducted to develop a regression equation for prediction of the inlet or initial $O_2 K_i a$ as a function of water flow rate and GW ratio. Calibration tests on the u-tube were steady-state. Water flow rates from 230 to 320 L/min were evaluated for a range of GW ratios. Steady-state inlet DO's at or near atmospheric saturation were maintained during testing until a steady-state outlet DO was established for a given combination of flowrate and GW ratio. Outlet DO and flowrate were measured and recorded on the data logger. Twenty tests were conducted for regression equation development and prediction comparison.

The u-tubes, as installed and operated within the recirculating systems at the VTAF, are limited to a maximum GW ratio of approximately 0.023 to 0.027 due to the system design. The available water head above the u-tube is approximately 0.76 m as provided by elevation of the RBC. Attempted operation at ratios at higher GW ratios is unreliable and prone to flow termination due to headlosses imparted by bubble holdup. U-tube calibration tests were conducted by pumping flow directly from the inlet water source through the u-tube. The increase in available head allowed tests to be conducted over a GW range of 0.01 to 0.06.

Ratios for GW were established by metering $O_2$ gas into the u-tube inlet at the required flow rate after correcting for u-tube inlet pressure. Pressure correction of gas flows was required due to the variation in u-tube inlet pressure head for different flow rates. The equation for pressure correction is given in Appendix A (p. 157) along with the flow calibration table developed for the Matheson
rotameter. Pressure correction for the Omega rotameter was not required since the calibration provided by the manufacturer specified an exit metering absolute pressure of 2 atm. The outlet pressure for the Omega rotameter was controlled by an inline pressure regulator located between the u-tube and the rotameter.

Headloss due to two-phase flow was also measured to collect data for regression analysis. Total headloss across the u-tube was measured in height of water by manometers positioned at the u-tube inlet and exit. Total headloss was recorded for two-phase flow for each flow rate and GW ratio combination and also recorded at each flow rate without O₂ injection. The headloss due to two-phase flow was assumed to be the difference between the two total headloss measurements.

5.2.4. **Multi-tube Clarifier**

The model assumes no gas transfer occurs within the tube clarifier and that changes in dissolved gasses occur due to biological respiration only. No preliminary calibration tests were designed for the clarifier submodel. A regression equation for estimation of biological O₂ demand rates was developed from DO data collected during the 1991 growth trials. The O₂ demand may be user input or estimated from the regression equation.

5.3. **Test Set 2 - Model Validation**

Model validation from actual production data while desirable would be difficult due to the uncertainties in estimation of biological O₂ demands and α ratios. Experimental tests were designed to control these parameters and provide data for model validation. Predetermined biological O₂ demand rates in both tank and
clarifier were simulated by metering in a sodium sulfite solution at a known concentration and flow rate. Production scenarios were simulated by creating $O_2$ demand rates similar to those observed during the 1991 growth trials. The u-tube was operated at GW ratios similar to those utilized during growths trials to maintain adequate DO concentrations in the recirculating systems. The RBC was operated to account for the effects of reaeration.

Concentrations of sodium sulfite solution were established by dissolving a measured quantity of dry sodium sulfite into a known volume of water. Individual solutions were made for the tank and the clarifier for each test and were metered by a multiple-head variable speed metering pump. Sulfite solution was injected at the bottom of the tank at six locations positioned uniformly over the tank bottom. The tank bottom measures approximately 6.1 m long by 1.6 m wide. A sulfite solution distribution grid (4.9 x 0.9 m) was constructed of 1.27 cm CPVC pipe and exit holes of 0.4 mm diameter were drilled into the grid at six evenly spaced locations. Sulfite solution was metered into the center of the grid. A dye test verified that flow was exiting all points in a uniform manner. Sulfite solution was metered into the clarifier at two points near the clarifier bottom located directly in the water flow inlets leading from the tank. Distribution of sulfite solution in this manner assured complete mixing of the sulfite with the clarifier water inflow before entering the multi-tube media.

Four separate tests were conducted. However, the tests were constructed such that each test provided 2 to 4 sets of validation data. Dissolved oxygen concentrations of the effluent exiting each component were continually monitored and recorded during each test at two minute intervals. Each two minute datum represented an
average of 24 measurements taken every 5 seconds. Water flow rates were recorded continuously in the same manner as for DO. Known GW ratios in the u-tube were established by metering O₂ at the required gas flows after pressure correction. Change in total gas pressure, ΔP_m, in the tank was measured with the gasometer.

The validation data were assumed to be the DO values exiting each component once the system had attained equilibrium for a given combination of O₂ demand, GW ratio, and recirculation flow rate. Once the system has attained equilibrium and the pertinent data recorded, a new combination of these parameters was tested by changing one or more and recording the system response. Two to four different combinations of these parameters were applied to the system during each main test allowing collection of multiple data sets for each main validation test. The system DO concentration response was then recorded by the data loggers as the system approached a new equilibrium condition. Tables 5.1 to 5.44 provide the detailed information on the test parameters for O₂ demands, GW ratio, and flow rate for each test scenario. Each combination was allowed to operate a minimum of 3 tank detention times to assure equilibrium conditions.

Validation Test 1 was intended to simulate a typical average O₂ demand of the system before feeding followed by an increase in demand by a factor of 1.45 due to feeding. The clarifier O₂ demand was held constant. The last scenario in Validation Test 1 was intended to evaluate the model prediction of equilibrium conditions during an emergency situation when recirculation flow was terminated.
**TABLE 5.1** Test parameters used to simulate different production conditions for Validation Test 1.

<table>
<thead>
<tr>
<th>SIMULATION SCENARIO</th>
<th>Recirc Flow L/min</th>
<th>Temp °C</th>
<th>Oxygen Supply</th>
<th>Component</th>
<th>Sulfite conc. g/L</th>
<th>Sulfite rate L/min</th>
<th>Oxygen Demand g/hr</th>
</tr>
</thead>
<tbody>
<tr>
<td>RESTING</td>
<td>240</td>
<td>22.5</td>
<td>GW ratio = 0.014</td>
<td>tank clarifier</td>
<td>48.0</td>
<td>0.28</td>
<td>102.0</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>39.5</td>
<td>0.097</td>
<td>29.8</td>
</tr>
<tr>
<td>FEEDING</td>
<td>240</td>
<td>23.5</td>
<td>GW ratio = 0.014</td>
<td>tank clarifier</td>
<td>71.0</td>
<td>0.28</td>
<td>150.0</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>39.5</td>
<td>0.097</td>
<td>28.8</td>
</tr>
<tr>
<td>EMERGENCY</td>
<td>0</td>
<td>23.5</td>
<td>emergency aerator</td>
<td>tank clarifier</td>
<td>71.0</td>
<td>0.28</td>
<td>150.0</td>
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<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>39.5</td>
<td>0.0</td>
<td>0.0</td>
</tr>
</tbody>
</table>

**TABLE 5.2.** Test parameters used to simulate different production conditions for Validation Test 2.

<table>
<thead>
<tr>
<th>SIMULATION SCENARIO</th>
<th>Recirc Flow L/min</th>
<th>Temp °C</th>
<th>Oxygen Supply</th>
<th>Component</th>
<th>Sulfite conc. g/L</th>
<th>Sulfite rate L/min</th>
<th>Oxygen Demand g/hr</th>
</tr>
</thead>
<tbody>
<tr>
<td>RESTING</td>
<td>320</td>
<td>22.5</td>
<td>GW ratio = 0.01</td>
<td>tank clarifier</td>
<td>88.5</td>
<td>0.20</td>
<td>135.0</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>55.0</td>
<td>0.07</td>
<td>29.4</td>
</tr>
<tr>
<td>FEEDING</td>
<td>320</td>
<td>23.0</td>
<td>GW ratio = 0.01</td>
<td>tank clarifier</td>
<td>88.5</td>
<td>0.28</td>
<td>189.0</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>55.0</td>
<td>0.095</td>
<td>39.6</td>
</tr>
<tr>
<td>FEEDING</td>
<td>320</td>
<td>23.0</td>
<td>GW ratio = 0.02</td>
<td>tank clarifier</td>
<td>88.5</td>
<td>0.28</td>
<td>189.0</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>55.0</td>
<td>0.095</td>
<td>39.6</td>
</tr>
<tr>
<td>RESTING</td>
<td>320</td>
<td>23.5</td>
<td>GW ratio = 0.02</td>
<td>tank clarifier</td>
<td>88.5</td>
<td>0.20</td>
<td>135.0</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>55.0</td>
<td>0.07</td>
<td>29.4</td>
</tr>
</tbody>
</table>
TABLE 5.3. Test parameters used to simulate different production conditions for Validation Test 3.

<table>
<thead>
<tr>
<th>SIMULATION SCENARIO</th>
<th>Recirc Flow L/min</th>
<th>Temp °C</th>
<th>Oxygen Supply GW ratio = 0.013</th>
<th>Component tank clarifier</th>
<th>Sulfite Conc. g/L</th>
<th>Addition Rate L/min</th>
<th>Oxygen Demand g/hr</th>
</tr>
</thead>
<tbody>
<tr>
<td>RESTING</td>
<td>285</td>
<td>22.5</td>
<td></td>
<td></td>
<td>54.0</td>
<td>0.285</td>
<td>117.0</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>47.5</td>
<td>0.095</td>
<td>34.2</td>
</tr>
<tr>
<td>FEEDING</td>
<td>285</td>
<td>22.5</td>
<td></td>
<td></td>
<td>86.0</td>
<td>0.285</td>
<td>186.0</td>
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<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>47.5</td>
<td>0.095</td>
<td>34.2</td>
</tr>
<tr>
<td>RESTING</td>
<td>285</td>
<td>22.5</td>
<td></td>
<td></td>
<td>54.0</td>
<td>0.285</td>
<td>117.0</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>47.5</td>
<td>0.095</td>
<td>34.2</td>
</tr>
</tbody>
</table>

TABLE 5.4. Test parameters used to simulate different production conditions for Validation Test 4.

<table>
<thead>
<tr>
<th>SIMULATION SCENARIO</th>
<th>Recirc Flow L/min</th>
<th>Temp °C</th>
<th>Oxygen Supply GW ratio = 0.025</th>
<th>Component tank clarifier</th>
<th>Sulfite Conc. g/L</th>
<th>Addition Rate L/min</th>
<th>Oxygen Demand mg/min</th>
</tr>
</thead>
<tbody>
<tr>
<td>FEEDING</td>
<td>275</td>
<td>22.0</td>
<td></td>
<td></td>
<td>125.0</td>
<td>0.30</td>
<td>285.6</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>66.5</td>
<td>0.090</td>
<td>46.5</td>
</tr>
<tr>
<td>FEEDING</td>
<td>275</td>
<td>22.0</td>
<td>GW ratio = 0.025 emergency aerator</td>
<td></td>
<td>125.0</td>
<td>0.30</td>
<td>285.6</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>66.5</td>
<td>0.095</td>
<td>46.5</td>
</tr>
</tbody>
</table>

71
and the $O_2$ supply to the tank was limited to transfer by the emergency aerator. Only the equilibrium tank DO and change in dissolved gas pressure was measured during the emergency scenario.

Validation Test 2 was intended to simulate resting and feeding demand at one GW ratio. The system response to an increase in the GW ratio was then recorded at the high $O_2$ demand followed by a decrease in $O_2$ demand to the resting rate while maintaining the increased $O_2$ supply. Oxygen demand in both the tank and clarifier were varied by changing the speed of the metering pump delivering the respective sulfite solutions.

Validation Test 3 simulated a resting rate followed by a feeding response and subsequent return to resting demand rate. The test was intended to verify the system would approach the initial equilibrium condition after an increase in demand. Operation at high GW ratios could not be tested due to the available head limitation. Validation Test 4 provided data for high exerted $O_2$ demands and u-tube operation near the limiting GW ratio.

5.4. Production Trial Data Collection

Dissolved oxygen data were also gathered from eight of the recirculating systems during the 1991 production trials. Hybrid striped bass were stocked in eight of the nine systems at three stocking densities. Therefore, different exerted $O_2$ demands were established at the three densities due to differences in total fish mass and daily feeding ration. The growth trails provided the opportunity to collect data for
estimation of typical O$_2$ demands in both the tank and clarifier as well as to evaluate u-tube performance over a range of GW ratios under field conditions.

Dissolved oxygen measurements were taken at times during the day when the DO in the recirculating systems was at or near equilibrium. Taking data at system equilibrium allowed reasonable estimation of the exerted O$_2$ demand or transfer rate in a component. Dissolved oxygen concentrations of the inflows and outflows of all components of several recirculating systems were measured intermittently throughout the growth trials as fish mass, feeding ration and u-tube GW ratio were increased. In addition to the DO concentration data, the following data were also recorded each time a DO analysis of a system was conducted: daily feeding ration, water temperature, average fish weight, and u-tube GW ratio.
6. RESULTS AND DISCUSSION

Each component subroutine was originally developed as a separate program to allow independent calibration and validation. The experimental test results presented in section 6.1 were utilized for component calibration and validation. The component subroutines were then incorporated into the main program and the results of section 6.3 were utilized for full-model validation.

6.1. Component Calibration - RESULTS

6.1.1. Tank

The values of $K_La$ evaluated for the 0.25 kW surface tank aerator are presented in Table 6.1. Both measured values and values standardized to a temperature of 20°C using Eq [4] are provided. The DO saturation values, $C_s$, presented in Table 6.1 are not theoretical values but were estimated as the asymptotic steady-state DO value of the tank at the conclusion of each test (Fig 6.1). All reaeration tests were allowed to run a sufficient length of time to estimate the saturation concentration in this manner. Estimates of $K_La$ were generated by input of trial and error values of $C_s$ into a spreadsheet model employing the log-deficit method. The predicted values of the time related DO concentration generated by Eq[2] were plotted as in Fig. 6.1 for each combination of $K_La$ and $C_s$. The combination that gave the best visual fit of predicted values versus measured values was accepted as the best estimate.
**TABLE 6.1** Test conditions for reaeration tests and evaluated transfer coefficients of tank aerators.

<table>
<thead>
<tr>
<th>Test</th>
<th>Flow</th>
<th>Temp</th>
<th>Saturation</th>
<th>$K_L a$</th>
<th>$K_L a_{20}$</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>L/min</td>
<td>°C</td>
<td>mg/L</td>
<td>min⁻¹</td>
<td>min⁻¹</td>
</tr>
<tr>
<td>1</td>
<td>361</td>
<td>21.0</td>
<td>7.9</td>
<td>0.033</td>
<td>0.032</td>
</tr>
<tr>
<td>2</td>
<td>219</td>
<td>23.5</td>
<td>7.5</td>
<td>0.031</td>
<td>0.029</td>
</tr>
<tr>
<td>3</td>
<td>221</td>
<td>21.5</td>
<td>7.8</td>
<td>0.029</td>
<td>0.028</td>
</tr>
<tr>
<td>4</td>
<td>358</td>
<td>22.5</td>
<td>7.6</td>
<td>0.033</td>
<td>0.031</td>
</tr>
</tbody>
</table>
Fig. 6.1  Reaeration data from tank reaeration test 1 plotted with predicted values using best fit estimates of $K_{L,a}$ and $C_s$. 
6.1.2. **Rotating Biological Contactor**

The value of $K_La$ in RBC component tests was estimated from measured inlet and outlet DO data with Eq [9]. The 1st stage $K_La$ for above surface discharge is plotted in Fig. 6.2. An equation was developed by linear regression of the temperature corrected values of $K_La$ against flow:

$$K_La_{20} = 0.00201\cdot Q - 0.013$$  \[31\]

The correlation coefficient ($r^2$) was 0.93. The equation predicted $K_La_{20}$ within 6 % of observed values and the RBC component model predicted the change in DO within 2 % during single stage transfer simulation (Table 6.2).

Values of $K_La$ for submerged discharge just beneath the water surface in the 1st stage and for bottom discharge during 2nd and 3rd stage simulation are plotted in Fig. 6.3. Measured values of submerged discharge varied as much as 11 % but no trends with flow were apparent. The value of $K_La$ for the 2nd and 3rd stage test varied less than 5 % and were assumed constant over the flow range tested. The average value (0.167 min$^{-1}$) was accepted as the model calibration value for the 2nd and 3rd stage $K_La$ at 20°C. There appeared to be no correlation between $K_La$ and flow rate for submerged flow tests or 2nd and 3rd stage tests, however, no formal statistical analyses were conducted.

Two tests were conducted for 3 stage RBC component model validation. The data are presented in Table 6.3. The component model predicted the change in DO between the inlet and outlet within 1.6 %.
Fig. 6.2. Observed versus $K_{aL_{20}}$ corrected for 1st stage of RBC with pipe discharge above water surface.
**Fig. 6.3.** Transfer coefficients corrected to 20° C for inflow discharge submerged just beneath water surface in 1st RBC stage and for discharge at RBC bottom to simulate influent flow through entrance baffle into 2nd and 3rd RBC stages.
**TABLE 6.2.** Measured values of $K_{L}a_{20}$ and change in DO ($\Delta$ DO) compared to predicted values of RBC model for single stage RBC.

<table>
<thead>
<tr>
<th>Test</th>
<th>Flow L/min</th>
<th>$K_{L}a_{20}$ (min$^{-1}$)</th>
<th>$\Delta$ DO (mg/L)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>meas</td>
<td>pred</td>
</tr>
<tr>
<td>1</td>
<td>323</td>
<td>0.641</td>
<td>0.636</td>
</tr>
<tr>
<td>2</td>
<td>258</td>
<td>0.479</td>
<td>0.506</td>
</tr>
<tr>
<td>3</td>
<td>195</td>
<td>0.364</td>
<td>0.379</td>
</tr>
</tbody>
</table>

**TABLE 6.3** Measured $\Delta$DO across 3-stage RBC compared to RBC model predictions. $\Delta$DO values are relative to RBC inlet DO concentration rather than individual stage inlet concentration.

<table>
<thead>
<tr>
<th>Test</th>
<th>Flow L/min</th>
<th>stage 1</th>
<th>stage 2</th>
<th>stage 3</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>meas L/mg</td>
<td>pred L/mg</td>
<td>%err</td>
</tr>
<tr>
<td>1</td>
<td>291</td>
<td>4.30</td>
<td>4.30</td>
<td>0.0</td>
</tr>
<tr>
<td>2</td>
<td>205</td>
<td>5.00</td>
<td>5.10</td>
<td>-2.0</td>
</tr>
</tbody>
</table>

*** Data were not taken.
6.1.3. **U-tube**

The value of $K_La$ within the plug-flow u-tube model is not constant but is a function of the GW ratio. Estimates of the initial $K_La_{20}$ at the u-tube inlet were determined by iteration within the u-tube model using performance data. For each combination of inlet and outlet DO measured, the model was used to estimate an inlet $K_La_{20}$ which predicted measured outlet DO at input operating conditions. A predictive equation was developed from the $K_La_{20}$ data with stepwise multiple linear regression analysis using GW ratio and flow rate as independent variables. Both the GW ratio ($p < 0.001$) and flow rate effects ($p = 0.0038$) were significant for prediction of $K_La_{20}$. Regression accounted for over 99% of observed variation. The regression equation was:

$$K_La_{20} = 0.49426*GW + 0.000017*Q - 0.004932$$  \[32\]

where

$K_La_{20}$ - inlet O$_2$ transfer co-efficient at 20°C, sec$^{-1}$

GW - GW ratio.

The calibration data are plotted in Fig. 6.4 as a function of the GW ratio. Five test runs were made to provide data for evaluation of predicted inlet $K_La_{20}$ and ΔDO of the u-tube component model (Table 6.4). A predictive equation for headloss due to two-phase flow was developed from headloss measurements. Only GW ratio proved significant ($r^2 = 0.95$) in predicting headloss, $h_g$ in m of water, over the range of flows tested.

$$h_g = 7.3GW$$  \[33\]
Fig. 6.4. U-tube inlet $K_L a_{20}$ values used in regression of $K_L a_{20}$ against GW ratio and flow rate.
Table 6.4. U-tube regression equation prediction of initial transfer coefficient and component prediction of u-tube $\Delta$ DO.

<table>
<thead>
<tr>
<th>Flow (L/min)</th>
<th>GW ratio</th>
<th>$K_{L_a20}$ (min$^{-1}$)</th>
<th>$\Delta$DO (mg/L)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Meas</td>
<td>Pred</td>
</tr>
<tr>
<td>279</td>
<td>0.010</td>
<td>0.0049</td>
<td>0.0047</td>
</tr>
<tr>
<td>212</td>
<td>0.026</td>
<td>0.0111</td>
<td>0.0115</td>
</tr>
<tr>
<td>210</td>
<td>0.040</td>
<td>0.0197</td>
<td>0.0184</td>
</tr>
<tr>
<td>320</td>
<td>0.016</td>
<td>0.0082</td>
<td>0.0084</td>
</tr>
<tr>
<td>300</td>
<td>0.051</td>
<td>0.0247</td>
<td>0.0254</td>
</tr>
</tbody>
</table>
6.2 Component Calibration DISCUSSION.

6.2.1. Tank

The reaeration tests provide an approximate estimate of $K_L a$ for the surface aerators since there appears to be a slight correlation to flow rate. The increase in flow causes a more complete mixing of the water volume. The increased mixing reduces the concentration gradient during reaeration established between a localized zone of water around the aerator with higher DO than the bulk water volume. Each value of $K_L a$ established in these tests would be termed the "apparent value" reflective of the test conditions. The observed value increased by 14 % as the recirculation flow rate increased from approximately 220 to 360 L/min. Since these devices were considered emergency aerators, development of a predictive equation of $K_L a_{20}$ against flow was not pursued.

$K_L a$ is a function of the aerated water volume in two respects. First, the observed value is directly related to the aerated water volume as given by Eq[1]. Therefore the values established during testing need to be corrected for volume changes if performance within a different tank volume is to be modeled. The $K_L a$ is corrected by multiplication of the ratio of the test volume over the new liquid volume. Second, $K_L a$ is related to the power transferred to the liquid by the device. Massive scale-up or scale down of performance parameters to largely different liquid volumes using the volumetric ratio as described above does not properly account for changes in mixing characteristics and erroneous values may be estimated (Schmidtke, 1986).
The effects of power transfer on mixing, regeneration of the gas to water interface, and surface film thickness reduction or renewal were demonstrated in the reaeration test conducted at high flow rate, approximately 330 L/min, without operation of the surface aerators. The test was intended to quantify the effect of gas transfer through a quiescent water surface in the tank. The DO rose less than 2 mg/L over a 36 hour period even after a measurable rise in concentration of 1 mg/L was established. The effects of quiescent surface transfer were considered insignificant in modeling tank gas transfer.

6.2.2. **RBC**

The effects of power transfer on $K_La$ are strongly reflected in the 1st stage transfer coefficient tests for above surface discharge. The $K_La_{20}$ increases over 40% from a flow of 200 L/min to 340 L/min. The submerged discharge coefficients established for discharge just beneath the water surface are essentially constant with flow and are very close to the constant value established for 2nd and 3rd stage transfer. These factors indicate that the majority of 1st stage transfer that occurs is due to energy and air-bubble entrainment imparted by above surface discharge, within a narrow localized zone around the area where inlet flow strikes the water. The values for 2nd and 3rd stage transfer were shown to be constant and independent of flow (Fig 6.3). The values would, however, be dependant on RBC design or operational parameters such as media surface area, rotational speed or water level. Coefficients for similar RBCs of different sizes can be approximated by dimensional analysis techniques using dimensionless ratios with some success if size differences are not too great (Schmidtke, 1986).
The rate of increase in the 1st stage transfer coefficient is directly proportional to increase in flow, and the effluent DO is largely independent of flow for a steady-state inflow DO concentration. The y-intercept term, in Eq [31] for predication of 1st stage $K_La_{20}$ is less than 5% of the predicted value of 0.402 min$^{-1}$ at a flow of 200 L/min. A t-test indicated the intercept is insignificant in prediction of 1st stage $K_La_{20}$. Substitution of Eq [9] into Eq [31] and solving for exit DO; $C_e$, at $C_s = 8$ mg/L and inlet DO, $C_{in} = 5$ mg/L; shows that predicted increase in steady-state effluent DO for flows of $Q = 200$ L/min to $Q = 400$ L/min is less than 0.002 mg/L. If the intercept term, is dropped from the regression equation, the effluent DO, $C_e$, is independent of flow rate and is constant for any combination of $C_{in}$ or $C_s$. The predicted transfer rate of $O_2$, or any dissolved gas, directly proportional to flow and the effects of decreased detention time at higher flows have little effect on 1st stage transfer.

The $K_La$ value for 2nd and 3rd stage transfer was found to remain constant, $K_La_{20} = 0.167$ min$^{-1}$, over the range of flow rates tested. At $K_La = 0.167$ min$^{-1}$, the steady-state effluent DO predicted by Eq [3] decreases slightly with increasing flow rate due to decreasing detention time. However, the transfer rate increases for increasing flow within the flow range of 200 - 400 L/min.

The transfer rate, TR, in the RBC for any gas into or out of solution is strongly dependant on flow rate (Fig. 6.5) The percent increase is calculated with respect to the TR that would occur at 200 L/min. The TR is determined from flow rate and the difference between the RBC inlet dissolved gas concentration and stage
Fig. 6.5  Increase in gas transfer rate (TR) for one stage RBC with above surface discharge and resulting $K_La_{20}$, one stage RBC with submerged discharge and resulting $K_La_{20}$, and the three stage RBC of the VTAF RBC design.
outlet concentration. Fig. 6.5 is plotted using Eq[31] for $K_L a_{20}$ for 1st stage above surface discharge, and submerged discharge with $K_L a_{20} = 0.167 \text{ min}^{-1}$ (2nd and 3rd stage).

The percentage increase for the 3 stage RBC curve accounts for the total transfer of across 3 stages of the RBC design located at the VTAF. Thus while the total TR for a three stage RBC is greater than transfer across a single stage for above surface discharge, the percentage increase is less. The percentage increase in TR is also independent of any combination of $C_{in}$ or saturation concentration, $C_s$.

The extra $O_2$ added to the water at higher flows may be insignificant due to low DO deficits established by minimum DO criteria of the fish. However, the effect of increasing flow may become important when considering the effects of stripping $CO_2$ to maintain concentrations below safe levels.

6.2.3. U-tube

Errors obtained for predicted $K_L a_{20}$ and predicted outlet ΔDO by the u-tube model suggest that errors in prediction of $K_L a_{20}$ do not appear to magnify model error in predicted ΔDO. The sensitivity of the model to prediction of $K_L a$ is greatest at low GW ratios and low flow rates. At 200 L/min and GW ratio of 0.01, a 10 % increase in the GW ratio to 0.011 increases the predicted outlet ΔDO by 12.5 % which is essentially equal to the increase in predicted $K_L a$. At a GW ratio of 0.04, a 10 % increase to 0.044 increases the predicted ΔDO by 7.7 % while the increase in predicted $K_L a$ is about 11 %. The increased sensitivity of the model to $K_L a_{20}$ and therefore GW ratio at lower GW ratio values is significant because the
measurement error associated with establishing the GW ratio is about ±7-9% due to the accuracy of gas (±5%) and liquid (±1.5%) flow measurement. In all tests, the error between the predicted $K_t a_{20}$ and the measured value is within the magnitude of error that could be attributed to measurement of the GW ratio.

Earlier researchers modeling u-tube performance found that $K_t a$ was proportional to the logarithm of the GW ratio (Speece and Orosco, 1969; Watten and Beck, 1985). However, $K_t a$ was determined to be directly proportional to flow in tests for the VTAF u-tube aerators. The range of values (0.01-0.06) tested was much smaller than the range of values tested by earlier researchers and therefore may represent only a small linear portion of the $K_t a$ GW curve. The relationship between two-phase flow and headloss was also found to be linear and independent of flow. Watten and Beck (1985) found that headloss was exponentially related to GW ratio, flow, and u-tube depth. The effects of depth were not investigated in these tests but the direct linear relationship may again be attributed to the small range of GW ratios and flow rates tested.
6.3. Model Validation-RESULTS

The results of the validation tests for simulated production scenarios and model predictions for equilibrium DO are presented in Tables 6.5 to 6.8. The measured DO value for each component is the equilibrium DO concentration exiting each component after operation at the validation test conditions provided in Tables 5.1-5.4. The operational and production parameters for each validation test were then input into the RAS model to generate the equilibrium DO concentrations that would be predicted for simulated culture.

Two sets of simulation data are presented for each validation test. Each simulation utilizes a different predictive equation of the initial $K_r a_{20}$ in the u-tube component. Simulation set 1 was generated with the regression equation, Eq [32], developed during u-tube component calibration. The model consistently predicted low equilibrium DO values indicating that u-tube component calibration was biased. The average error of model predictions for $\Delta DO$ between the u-tube inlet and outlet was -13.6%. The resulting error in predicting equilibrium DO in the tank averaged -17% and ranged from -33 to -3%.

The emergency simulation in Validation Test 1, predicting a 0 mg/L DO in both simulations, was not included in average error calculations since the system never appeared to reach a true equilibrium. During this simulation, all water flow was terminated resulting in poor mixing of the sulfite in the tank. Poor mixing was evident in DO probe behavior as the reading was unstable ranging from 0.2 to 2 mg/L over short time spans during the emergency scenario.
### TABLE 6.5  
Measured equilibrium DO concentrations and RAS model predictions for Validation Test 1.

<table>
<thead>
<tr>
<th>Scenario</th>
<th>Component</th>
<th>Measured mg/L</th>
<th>Predicted Simulation 1 mg/L</th>
<th>Error mg/L</th>
<th>Error %</th>
<th>Predicted Simulation 2 mg/L</th>
<th>Error mg/L</th>
<th>Error %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Resting</td>
<td>Tank</td>
<td>10.1</td>
<td>8.6</td>
<td>-1.5</td>
<td>-14.9</td>
<td>10.2</td>
<td>0.1</td>
<td>1.0</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>8.1</td>
<td>6.6</td>
<td>-1.5</td>
<td>-18.5</td>
<td>8.0</td>
<td>-0.1</td>
<td>-1.2</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>8.0</td>
<td>7.8</td>
<td>-0.2</td>
<td>-2.5</td>
<td>8.0</td>
<td>0.0</td>
<td>0.0</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>16.9</td>
<td>15.8</td>
<td>-1.1</td>
<td>-8.5</td>
<td>17.6</td>
<td>0.7</td>
<td>4.1</td>
</tr>
<tr>
<td>Feeding</td>
<td>Tank</td>
<td>6.3</td>
<td>5.3</td>
<td>-1.0</td>
<td>-15.9</td>
<td>7.3</td>
<td>1.0</td>
<td>15.9</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>4.7</td>
<td>3.4</td>
<td>-1.3</td>
<td>-27.7</td>
<td>5.4</td>
<td>0.7</td>
<td>14.9</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>* 7.2</td>
<td>7.1</td>
<td>-0.1</td>
<td>-1.4</td>
<td>7.5</td>
<td>0.3</td>
<td>4.2</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>* 16.4</td>
<td>15.3</td>
<td>-1.1</td>
<td>-8.7</td>
<td>17.3</td>
<td>0.9</td>
<td>5.5</td>
</tr>
<tr>
<td>Emergency</td>
<td>Tank</td>
<td>1.0</td>
<td>0.0</td>
<td>-1.0</td>
<td>-100.0</td>
<td>0.0</td>
<td>-1.0</td>
<td>-100.0</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>***</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>***</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>***</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

* Indicates those data utilized to generate a second prediction equation for inlet u-tube $K_{L}$a

### Data were not taken
TABLE 6.6  Measured equilibrium DO concentrations and RAS model predictions for Validation Test 2

<table>
<thead>
<tr>
<th>Scenario</th>
<th>Component</th>
<th>Measured mg/L</th>
<th>Simulation 1</th>
<th>Simulation 2</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td>Predicted mg/L</td>
<td>Error mg/L</td>
</tr>
<tr>
<td>Resting</td>
<td>Tank</td>
<td>6.7</td>
<td>6.5</td>
<td>-0.2</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>5.3</td>
<td>5.1</td>
<td>-0.2</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>* 7.3</td>
<td>7.4</td>
<td>0.1</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>*13.4</td>
<td>13.1</td>
<td>-0.3</td>
</tr>
<tr>
<td>Feeding</td>
<td>Tank</td>
<td>3.6</td>
<td>2.8</td>
<td>-0.8</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>2.1</td>
<td>0.8</td>
<td>-1.3</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>* 6.5</td>
<td>6.4</td>
<td>-0.1</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>*12.8</td>
<td>12.2</td>
<td>-0.6</td>
</tr>
<tr>
<td>Feeding</td>
<td>Tank</td>
<td>9.4</td>
<td>7.5</td>
<td>-1.9</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>7.4</td>
<td>5.3</td>
<td>-2.1</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>7.8</td>
<td>7.4</td>
<td>-0.4</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>19.6</td>
<td>17.8</td>
<td>-1.8</td>
</tr>
<tr>
<td>Resting</td>
<td>Tank</td>
<td>13.1</td>
<td>11.1</td>
<td>-2.0</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>11.3</td>
<td>9.5</td>
<td>-1.8</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>8.8</td>
<td>8.3</td>
<td>-0.5</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>22.6</td>
<td>18.6</td>
<td>-4.0</td>
</tr>
</tbody>
</table>

* Indicates those data utilized to generate a second prediction equation for inlet u-tube $k_{1\alpha}$
### TABLE 6.7  Measured equilibrium DO concentrations and RAS model predictions for Validation Test 3

<table>
<thead>
<tr>
<th>Scenario</th>
<th>Component</th>
<th>Measured mg/L</th>
<th>Simulation 1</th>
<th>Simulation 2</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Predicted mg/L</td>
<td>Error mg/L</td>
<td>Error %</td>
</tr>
<tr>
<td>Resting</td>
<td>Tank</td>
<td>9.3</td>
<td>8.1</td>
<td>-1.2</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>7.6</td>
<td>6.1</td>
<td>-1.5</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>8.0</td>
<td>7.6</td>
<td>-0.4</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>16.4</td>
<td>14.9</td>
<td>-1.5</td>
</tr>
<tr>
<td>Feeding</td>
<td>Tank</td>
<td>4.8</td>
<td>3.7</td>
<td>-1.1</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>3.3</td>
<td>1.8</td>
<td>-1.5</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>7.0</td>
<td>6.8</td>
<td>-0.2</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>16.7</td>
<td>14.1</td>
<td>-2.6</td>
</tr>
<tr>
<td>Resting</td>
<td>Tank</td>
<td>9.6</td>
<td>7.9</td>
<td>-1.7</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>7.6</td>
<td>5.8</td>
<td>-1.8</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>7.9</td>
<td>7.5</td>
<td>-0.4</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>16.7</td>
<td>14.8</td>
<td>-1.9</td>
</tr>
</tbody>
</table>

### TABLE 6.8  Measured equilibrium DO concentrations and RAS model predictions for Validation Test 4

<table>
<thead>
<tr>
<th>Scenario</th>
<th>Component</th>
<th>Measured mg/L</th>
<th>Simulation 1</th>
<th>Simulation 2</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Predicted mg/L</td>
<td>Error mg/L</td>
<td>Error %</td>
</tr>
<tr>
<td>Resting</td>
<td>Tank</td>
<td>4.5</td>
<td>3.0</td>
<td>-1.5</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>1.7</td>
<td>0.3</td>
<td>-1.4</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>6.5</td>
<td>6.5</td>
<td>0.0</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>21.8</td>
<td>20.2</td>
<td>-1.6</td>
</tr>
<tr>
<td>Feeding</td>
<td>Tank</td>
<td>6.4</td>
<td>5.4</td>
<td>-1.0</td>
</tr>
<tr>
<td></td>
<td>Sump</td>
<td>4.3</td>
<td>2.4</td>
<td>-1.9</td>
</tr>
<tr>
<td></td>
<td>RBC</td>
<td>*7.1</td>
<td>6.8</td>
<td>-0.3</td>
</tr>
<tr>
<td></td>
<td>U-tube</td>
<td>*22.5</td>
<td>20.4</td>
<td>-2.1</td>
</tr>
</tbody>
</table>

* Indicates data utilized to generate a second prediction equation for inlet u-tube $K_{L}a$
In an effort to correct the erroneous u-tube calibration and improve model predictions, a new $K_a a_{20}$ regression equation for the u-tube was developed from measured u-tube inlet and outlet concentration data taken from 4 of the test scenarios. The data were selected to span the range of GW ratios utilized during validation testing. The new regression equation ($r^2 = 0.99$) utilizes only the GW ratio as a predictor.

$$K_a a_{20} = 0.61344*GW - 0.000386$$  \[34\]

Prediction of outlet DO for the remainder of the Test Scenarios was improved substantially. The average relative error for prediction of equilibrium tank DO was less than -0.1 %. The range of absolute error for prediction of tank DO was ±0.7 mg/L error but due to the low tank concentrations involved, the range of relative error was slightly over ±15.5 %. Table 6.9 provides the average and range of error of predicted DO for each component from Simulation 2.

The equilibrium $\Delta P_m$ at the tank outlet was measured in the tank during each test scenario for comparison to predicted values. Table 6.10 lists the measured and values predicted from Simulation 2 for $\Delta P_m$ $N_2$+Ar concentration. The measured $N_2$+Ar concentration were calculated from measured equilibrium tank DO and $\Delta P_m$ using Eqs [5] and Eq [6]. A negative $\Delta P_m$ was measured in all test scenarios and in all but one scenario in Validation 2 the model over predicts $\Delta P_m$. The range of errors is large. The measurements made by the gasometer are suspect of a leak in the gas permeable tubing. In two tests, the saturometer was also used to measure $\Delta P_m$ and in both cases indicated a larger negative change and was somewhat closer in agreement with the predicted values than the saturometer.
TABLE 6.9 Summary of absolute and relative errors of RAS model equilibrium DO predictions of validation tests for Simulation 2.

| Component | Absolute | | | | Relative | |
|-----------|----------|---|---|---|---|
|           | average  | range | | | average | range |
|           | mg/L     | | | | % |
| Tank      | -0.05    | -0.7 to 0.7 | | | -0.1 | -15.9 to 15.6 |
| Sump      | -0.3     | -1.0 to 0.8 | | | -5.0 | -47.6 to 47.1 |
| RBC       | 0.01     | -0.2 to 0.4 | | | 0.5 | -2.6 to 6.2 |
| U-tube    | -0.3     | -2.9 to 0.9 | | | -2.1 | -12.8 to 5.5 |
TABLE 6.10  Measured and predicted equilibrium $\Delta P_m$ and $N_2 + Ar$ concentrations in tank for Simulation 2.

<table>
<thead>
<tr>
<th>Test</th>
<th>Scenario</th>
<th>$\Delta P$</th>
<th>$N_2 + Ar$</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Measured</td>
<td>Predicted</td>
</tr>
<tr>
<td></td>
<td></td>
<td>mm Hg</td>
<td>mm Hg</td>
</tr>
<tr>
<td>1</td>
<td>Resting</td>
<td>-40</td>
<td>-39</td>
</tr>
<tr>
<td></td>
<td></td>
<td>-93</td>
<td>-108</td>
</tr>
<tr>
<td></td>
<td>Emergency</td>
<td>-127</td>
<td>-154</td>
</tr>
<tr>
<td>2</td>
<td>Resting</td>
<td>###</td>
<td>-69</td>
</tr>
<tr>
<td></td>
<td></td>
<td>###</td>
<td>-152</td>
</tr>
<tr>
<td></td>
<td>Feeding</td>
<td>-100</td>
<td>-107</td>
</tr>
<tr>
<td></td>
<td></td>
<td>-62</td>
<td>-47</td>
</tr>
<tr>
<td>3</td>
<td>Resting</td>
<td>-41</td>
<td>-48</td>
</tr>
<tr>
<td></td>
<td></td>
<td>-115</td>
<td>-142</td>
</tr>
<tr>
<td></td>
<td>Feeding</td>
<td>-51</td>
<td>-70</td>
</tr>
<tr>
<td></td>
<td></td>
<td>-143</td>
<td>-186</td>
</tr>
<tr>
<td></td>
<td></td>
<td>-75</td>
<td>-113</td>
</tr>
</tbody>
</table>

### Data were not taken
6.4. Model Validation Discussion

The RAS model provided good prediction of equilibrium DO in Simulation 2 (Table 6.9) when the second $K_La_{20}$ prediction equation, Eq [34], for the u-tube component model was used. The use of 4 validation scenario data sets to develop the second u-tube $K_La_{20}$ prediction equation provided the remaining 7 data sets for model validation. The experimental design, constructed to collect the original u-tube calibration data, was not representative of the conditions under which the u-tubes normally operate. Inlet pressures higher than the normal RBC elevation head of 0.76 m were established at the u-tube inlet during u-tube component testing. Increase of injected O$_2$ mass flow rates due to pressure increase at the u-tube inlet were accounted for according to the equation presented in Appendix A, p. 157. Thus, the increased inlet pressure does not account for the error in the first calibration equation, Eq [32].

The second equation, Eq [34] predicts a value for $K_La_{20}$ approximately 35 % greater than Eq [32] for a given GW ratio. Erroneous over estimation of the GW ratio for those measured $K_La$ values utilized to develop the regression equation would result in an equation that under predicts $K_La$ for the true GW ratio value as was observed for the validation tests. However, in order to induce a positive error of 35 % in the measured $K_La$ value, the required error in the GW ratio measurement would have to be greater than 30 % which seems unlikely. No direct cause for the differences in u-tube performance between the calibration and validation tests is apparent. The error may possibly be attributed to differences in
turbulence levels at u-tube inlet and/or errors in the GW ratio due to erroneous gas flow or liquid flow measurements.

Comparison of prediction errors for Simulation 1 and Simulation 2 (Tables 6.5-6.8) indicates the sensitivity of the model to u-tube performance. Relative error in prediction of u-tube effluent DO averaged -9% while error in predicted OTR across the u-tube averaged about -13% for Simulation 1. The second u-tube K\textsubscript{L}a prediction equation, Eq [34], provided better prediction of u-tube OTR and subsequently better prediction of component equilibrium DO (Tables 6.5-6.8) and the average error in prediction of u-tube effluent DO decreased to -2.1% (Table 6.9). Therefore, accurate u-tube OTR estimation is critical since the magnitude of error in prediction of tank DO decreased from 18% in Simulation 1 to 0.1% in Simulation 2.

The range of relative errors obtained for component DO in Simulation 2 (Table 6.9) appears to suggest poor RAS model performance. The large range of errors established for the sump occurred at DO concentrations less than 4 mg/L. The largest relative error, 47.6%, occurred at an absolute error of 1 mg/L (Table 6.9). Figure 6.6 indicates the magnitude of absolute error in prediction of tank and clarifier DO is, in all but three scenarios, approximately equal to the error in u-tube effluent DO. Thus, at low DO concentrations in the tank and sump, the relative errors would naturally be large even for small absolute errors. At low DO concentrations, absolute rather than relative errors may be a better indicator of performance. Considering the difficulties in obtaining u-tube calibrations which would consistently predicted outlet u-tube DO within ± 1.0 mg/L at high OTR,
Fig 6.6 Error between predicted and measured component DO from Simulation 2 for validation scenarios comparing u-tube to tank, RBC, and clarifier.
the small absolute errors in prediction of component DO suggest model performance may be quite good and validation was successful.

The measured values of \( \Delta P_m \) have been noted as possibly erroneous due to a faulty gasometer. If the measured values are accepted as representative, the magnitude of the error in \( \Delta P_m \) (Table 6.10) may be attributed to an erroneous value of ratio of \( K_L a_{(\text{N}_2)} / K_L a_{(\text{O}_2)} \) as provided by Eq [12] and the liquid phase diffusivities of \( \text{N}_2 \) and \( \text{O}_2 \). A value of 0.94 was input for ratio of \( K_L a_{(\text{N}_2)} / K_L a_{(\text{O}_2)} \). A lower value, such as 0.89 (Watten and Beck, 1985, Speece and Orasco, 1969), may improve overall model performance.

Inclusion of Ar in the \( \text{N}_2 \) concentration and considering these two gases as a single gas \( \text{N}_2 + \text{Ar} \) may possibly impart some slight error due to a differing transfer coefficient. The ratio for Ar alone as calculated by Eq [12] and liquid phase diffusivities for Ar and \( \text{O}_2 \) at 25°C is approximately 1.0. The errors attributed to Ar alone would be negligible since the Ar mole fraction within any gas transferred to or from the liquid would be very near its atmospheric mole fraction of 0.0093.

Errors in \( \text{N}_2 + \text{Ar} \) concentration (Table 6.10) are dependant upon both the error in prediction of DO as well as \( \Delta P_m \) and exhibit no pattern in relation to \( \Delta P_m \) or DO error alone. The largest absolute error in prediction of \( \text{N}_2 + \text{Ar} \) concentration was only 1.4 mg/L. For a given \( \Delta P_m \), a change of 0.5 mg/L in DO results in a change in the \( \text{N}_2 + \text{Ar} \) concentration determined from Eqs [4] and [5] of 0.22 mg/L while a change in \( \Delta P_m \) of 10 mm Hg at constant DO results in a change of 0.26 mg/L. Thus considering the combined effects of small errors in measurement of DO and
ΔP_m, the prediction of N_2+Ar concentration by the RAS model is considered acceptable.

In all validation tests, ΔP_m < 0 were measured, thus total gas pressures were less than 100% even when DO concentrations were supersaturated. The RAS model predicts N_2+Ar stripping rates sufficient to lower N_2+Ar concentrations in the tank 1 to 4 mg/L below saturation resulting in substantial reductions in TGP. The largest tank DO, both measured and predicted, occurred in Validation Test 2 with RAS operating at a u-tube GW ratio of 0.02 and the production tank incurring a resting or average O_2 demand. The measured concentration was 13.1 mg/L or 165% of saturation and the predicted value was 12.9 mg/L or 163% of saturation (Table 6.10). The measured and predicted ΔP_m values were -67 mm HG and -47 mm HG, respectively, representing a TGP < 94% for a barometric pressure of 710 mm HG. The measured and predicted data indicate that TGP > 100% may not occur in the tank until DO concentrations are very supersaturated. Therefore the danger of Gas Bubble Trauma is essentially non-existent under normal operating conditions in a RAS employing u-tube coupled with pure oxygen injection.
6.5. Production Data - Results

Dissolved oxygen data collected from several recirculating systems during the 1991 growth trials were used to establish typical $O_2$ transfer and consumption rates of each component of the RAS under production conditions. The influent and effluent DO concentration of each component was measured and the consumption or transfer rate then determined by the product of the change in concentration and flow rate.

6.5.1. Tank

Figure 6.7 is a plot of DO concentration over a selected 24 hour period within a RAS at the VATF. The magnitude of change in the daily DO concentration is dependant upon mass loading density of fish and feeding ration. Early during growth trials when loading densities were low, the systems exhibited little or no daily DO cycles. As loading density increases due to growth, the daily increase in $O_2$ demand in each system becomes more pronounced. Due to the cyclic nature of DO concentration, DO data for all components were taken at midday or early morning when concentrations were near equilibrium in an attempt to establish representative $O_2$ consumption and transfer rates.

Oxygen demands in the production tank were estimated as the product of the system recirculation flow rate and the difference between the u-tube effluent DO concentration (tank inlet) and the tank exit DO concentration (clarifier inlet). The majority of the data collected were midday concentrations. These data were utilized to establish post feeding $O_2$ consumption rates when daily
Fig. 6.7 Dissolved oxygen concentration in RAS production tank over a selected 24 hour period showing cyclic nature of DO and $O_2$ demand. Loading Density approximately 72 kg/m$^3$. 
demands were maximum. The data are plotted in Fig. 6.8 corresponding to the average of the previous 5 days daily feeding ration. The data were fitted to the following model to establish a correlation between $O_2$ demand and feed ration:

$$DEM_i = \frac{1.44 \cdot OFR \cdot FR}{0.024}$$  \hspace{1cm} [35]

where

- $DEM_i$ - $O_2$ consumption rate in production tank, g/hr
- $OFR$ - Oxygen feed ratio, kg$O_2$/kg-feed
- $FR$ - Daily feed ration, kg/day.

The value of $OFR$ which provided the best fit through the observed data was 0.3. The constant, 1.44, is the recommended scaling factor (Westers, 1981) for estimating post feeding peak $O_2$ demand from the daily average demand estimate provided by the $OFR$ and $FR$.

Regression analysis was also used to developed a predictive equation for estimating daily maximum demand as a function of feed ration and average fish weight from the same demand data in an effort to improve prediction of demand.

$$DEM_i = 16.8 \cdot FR + 0.51 \cdot W - 13.7$$  \hspace{1cm} [36]

where $W$ is the average fish weight, g.

Regression accounts for 93% of the observed variation in the sample and both $FR$ and $W$ tested as significant predictor variables at $p < 0.001$. The predicted values of $DEM_i$ are also plotted in Fig. 6.8 for the $FR$ and $W$ data used in regression.
Fig. 6.8 Oxygen demands established in RAS production tanks during the 1991 growth trials. Predicted values for observed feeding rations are plotted for comparison to observed data.
During the last week of growth trials, tank O$_2$ demands were measured at early morning and midday to establish representative daily peaking factors for estimating maximum daily demands from early morning demands (TABLE 6.11). The values ranged from 1.31 to 1.54 and averaged 1.44. The range of tank DO concentrations established at early morning as compared to midday illustrate the range of the DO demand cycle.

6.5.2. Sump

Oxygen demands in the clarifier were estimated from the measured decrease in DO concentration across the clarifier and often represented greater than 10-15 % of the total O$_2$ demand measured in the RAS. The following regression equation was developed from clarifier demand data to predict clarifier O$_2$ demand

$$DEM_{mc} = 2.52 \cdot FR + 1.1$$  \[36\]

where is DEM$_{mc}$ O$_2$ demand in clarifier, g/hr.

The data exhibited significantly more variation, $r^2 = 0.67$, than demand data estimated for the production tank (Fig. 6.9).

U-tube

Inlet and outlet DO data were used to establish typical performance under production conditions as opposed to clean water conditions under which the u-tube component model was calibrated. Gas/water ratios were determined from system flow rate and oxygen gas injection rate. Oxygen transfer efficiency was determined from the OTR and the mass rate of O$_2$ injected into the u-tube liquid
Table 6.11  Early morning and midday DO concentrations and estimated O$_2$ demands established in the eight RAS. Peaking factors were determined as the quotient of the midday demand and early morning demand.

<table>
<thead>
<tr>
<th>RAS</th>
<th>Flow L/min</th>
<th>7:00 AM</th>
<th>12:00 PM</th>
<th>Peaking Factor</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Inlet DO mg/L</td>
<td>Exit DO mg/L</td>
<td>Demand g/hr</td>
</tr>
<tr>
<td>2</td>
<td>240</td>
<td>16.2</td>
<td>10.1</td>
<td>87.8</td>
</tr>
<tr>
<td>3</td>
<td>235</td>
<td>16.0</td>
<td>10.8</td>
<td>73.3</td>
</tr>
<tr>
<td>4</td>
<td>275</td>
<td>10.7</td>
<td>8.2</td>
<td>41.2</td>
</tr>
<tr>
<td>5</td>
<td>245</td>
<td>17.0</td>
<td>9.6</td>
<td>108.8</td>
</tr>
<tr>
<td>6</td>
<td>230</td>
<td>14.9</td>
<td>10.1</td>
<td>66.2</td>
</tr>
<tr>
<td>7</td>
<td>275</td>
<td>14.4</td>
<td>10.1</td>
<td>70.9</td>
</tr>
<tr>
<td>8</td>
<td>275</td>
<td>16.0</td>
<td>13.5</td>
<td>41.2</td>
</tr>
<tr>
<td>9</td>
<td>270</td>
<td>12.5</td>
<td>9.9</td>
<td>42.1</td>
</tr>
</tbody>
</table>

Average 1.44
Fig. 6.9 Oxygen demands measured in the multi-tube clarifier and regression equation, Eq [37], predictions.
stream. The observed OTE under production conditions was then used to attempt to estimate a value of the \( \alpha \) ratio through iteration within the u-tube component model.

Observed OTE ranged from 30 to 56 %. Fig. 6.10 shows the trend in observed OTE for one of the RAS during the growth trials. Efficiencies above 45 % were observed primarily for Gas/water ratios less than 0.005. Efficiencies less than 35 % were observed for Gas/water ratios primarily for u-tube operation at Gas/water ratios greater than 0.015. Flow rates in the eight RAS ranged from 230 to 280 L/min, making it difficult to establish decrease in OTE due to Gas/water ratio and water quality effects alone.

The observed OTE were initially assumed to be a function of water quality effects on \( K_{l,a} \). A value for the \( \alpha \) ratio was estimated for all u-tube RAS production performance data sets in which the Gas/water ratio was greater than 0.007. Those data for which the u-tube ratio was less than 0.007 were not used since the lowest Gas/water ratio calibration range of the u-tube was approximately 0.009.

The observed operating conditions; inlet DO, flowrate, temperature, and Gas/water ratio were input to the u-tube component model. The \( \alpha \) ratio was incrementally decreased until the difference between the predicted OTE and the observed value was less than 5 %. Predicted OTE increases less than 5 percent for a reduction in inlet \( N_2 + Ar \) from 14 mg/L to 5 mg/L at 20 °C and barometric pressure of 1 atm.

Figure 6.11 indicates the trends of the estimated \( \alpha \) ratio with Gas/water ratio and flow. Theoretically, \( \alpha \) should be independent of either of these factors and only a
Fig. 6.10  Range of OTE observed for one RAS during the 1991 growth trials.
Fig. 6.11  Estimated values of $\alpha$ ratio for RAS under production conditions showing apparent trends with Gas/water ratio and flow rate.

Low flow ($230 < \text{flow} < 245 \text{ L/min}$)  High flow ($270 < \text{flow} < 280 \text{ L/min}$)
function of water quality. The trends in these data suggest some other factor, presumably flow rate, affects the OTE in some manner which is not accounted for by the u-tube model.

Four of the RAS operated at significantly lower flow rates (230-245 L/min) than did the remaining four (270-280 L/min). These four RAS were also the only ones to establish Gas/water ratio greater than 0.015 during the growth trials and tended to exhibit significantly lower OTE. The model estimates of $\alpha$ were significantly lower for those four RAS over the range of Gas/water ratios for which estimates were made. The estimated value of $\alpha$ measured in RAS for the lower flow range averages 0.57 while the average for the high flow range was 0.78. Due to the obvious dependance on flow rate, these values are not realistic estimates of $\alpha$.

6.6. Discussion- Production DO Data

6.6.1. Tank

Dissolved oxygen concentrations in the RAS system are sensitive to changes in $O_2$ demand exerted by the fish. Demands were measured primarily to establish data for typical maximum daily $O_2$ demands and to establish a value for OFR which would predict demands according to Eq [34]. The regression equation, Eq [35] incorporating average fish weight was developed to ascertain if inclusion of average fish weight into a prediction equation would improve prediction of $O_2$ demand by the RAS model. Inclusion of average fish weight accounts for an additional 3.5 % of observed variability in the observed demand data than does feed ration alone. However, the daily feed ration provides the strongest predictor
of O₂ demand and the use of the OFR relationship provided by Eq [35] is commonly accepted (Colt and Orwicz, 1991; Colt and Watten, 1989). Since the regression equation, Eq [36], appears to improve prediction only slightly over the OFR model, Eq [35], and Eq [35] does not require an estimate of average fish weight, the method utilized by the model for estimation of tank O₂ demand will be Eq [35].

Daily average O₂ demands were not determined and therefore a direct estimate of OFR for daily average demand could not be made. Estimation of an average peaking factor was intended to provide a correlation between early morning resting demands and daily maximum demand rate (Table 6.11). The average value for the observed peaking factors, 1.44, is equal to the design value recommended by Westers (1981).

Figure 6.6 and the observed average for peaking factor, 1.44, suggests that the average daily O₂ demand would be somewhat greater than the early morning demand. Therefore, either the value of OFR or the peaking factor would be less than the values obtained from these data if each are correlated with the average and maximum demand rather than the minimum and maximum. Based on the observation in these data for O₂ demand and peaking factor, the default design values of OFR and peaking factor for the RAS system model are 0.3 and 1.44, respectively.

6.6.2. Sump

The observed data provide the only source of information for estimation of the O₂ demand exerted by the multi-tube clarifier. The range and variability in the data is
quite large (Fig 6.9). Correlation of the demand data with clarifier dumping schedules may have improved prediction. The clarifiers were routinely flushed of collected solids in relation to mass rate of feed input to the RAS system. The clarifiers were dumped after every 3 kg feed had been input into the RAS.

Since the RAS received different daily feed rations, all were not dumped on the same schedule. Clarifier demand data were also collected without consideration to dumping schedules. Therefore, in addition to feed ration, the exerted $O_2$ demand for any observation may be correlated to the length of time the clarifier has collected particulates since the previous flushing due to increasing mass loading of collected solids. The variability in the clarifier demand data may have been reduced if $O_2$ demands were measured on a consistent schedule just prior to flushing. The regression equation developed from the observed data is used by the RAS model for prediction of clarifier $O_2$ demand.

6.6.3. U-tube

The attempt to quantify water quality effects alone on u-tube performance by using production condition performance data to estimate $\alpha$ were unsuccessful. The values of $\alpha$ determined in this study exhibited dependance on flow rate through the u-tube. The u-tube component model predicts increasing OTE for decreasing flow rate assuming all other factors, including Gas/water ratio, are remain constant while the OTE decreases for an increasing Gas/water ratio assuming all other factors are constant. Therefore, even though a decrease in $\alpha$ results in a decrease in OTE, the decrease in OTE attributed solely to increased Gas/water ratio should be independent of $\alpha$. Using the calibrated u-tube
component model to estimate $\alpha$ for observed values of OTE was intended to account independently for the effects of variation in OTE due to Gas/water ratio and flow rate. The value of $\alpha$ which was estimated could then be assumed to quantify the effects on OTE due to water quality effects alone.

The observed difference in OTE and $\alpha$ for the two ranges of flow rates can possibly be explained by erroneous data, especially erroneous measurement of the Gas/water ratio, or a numerical model which does not adequately account for the combined effects of Gas/water ratio, flow, u-tube design or $\alpha$. The u-tube model has demonstrated the ability to adequately predict $O_2$ transfer for clean water conditions both in this research and previous studies and therefore should adequately model transfer in "dirty" water under production conditions if realistic estimates of $\alpha$ were known. The observed data were taken over several months time and flow rates in each RAS were determined periodically making the distinct separation of $\alpha$ into two flow ranges due to errors alone unlikely. All u-tubes were constructed identically as in Fig. 3.2 with only minor differences in depth (<1 meter) making differences in OTE due to u-tube design for different RAS under like operating conditions also unlikely. Thus the distinct trends observed for $\alpha$ and OTE in the production data remain fully unexplained.

Two distinct ranges in the estimated value of the $\alpha$ ratio were observed (Fig. 6.11). Each range corresponds to a range of recirculation flow rates in the RAS. Those RAS exhibiting lower flow rates, OTE and $\alpha$ were also the RAS in which loading densities and daily feeding rations were highest, which generally resulted in slightly higher dissolved organic carbon, (DOC) levels. Dissolved organic carbon concentrations averaged 10-15 mg/L in the low density systems and 15-25
mg/L in the high density RAS. The increased DOC levels in the higher density, lower flow rate, RAS may generate lower α ratios than those with lower DOC levels but values in the range of 0.45 to 0.65 due solely to water quality effects are unlikely. The average value of 0.57 is near the low boundary range cited earlier for municipal wastewaters, 0.4 to 1.2 (Doyle and Boyle, 1986). Therefore, these values are not considered adequate estimates of α. If it is desired to force predicted u-tube model performance values of OTE closer to the values observed in this study for actual production conditions, the average estimate of α for each flow range would more properly be considered as simply a correction factor.

The value of α is designated as user input with a default value of 1.0. Based the literature review, α may range from 0.4 to 1.2 for municipal wastewaters (Doyle and Boyle, 1986) and 0.7 to 1.1 for aquaculture pond water (Shelton and Boyd, 1983). Since water in recirculating systems may contain higher DOC levels than pond waters, the typical range of α may be lower, however, a value 0.4 is certainly uncharacteristic of normal aquaculture conditions. A suggested value typical for u-tube operation under RAS field production conditions would be 0.7 due the DOC levels typically found in RAS and the range of α measured by Shelton and Boyd (1983) for pond water.
6.7. Model Simulations

Model simulations were conducted to investigate the effects of recirculation flow rate, equilibrium DO, off gas recycle, automatic control of GW ratio, and decreasing u-tube $\alpha$ ratio on $O_2$ transfer costs and RAS gas transfer performance. Simulations were conducted for increasing $O_2$ demands comprised of production tank $O_2$ demand and clarifier demand estimated from increasing daily feed ration. Average daily demands in the production tank were calculated and multiplied by a peaking factor of 1.44 to represent maximum daily demands for all simulations unless otherwise noted. The range of feed rations investigated represent the typical range of feed rations input to the RAS over the course of the 1991 growth trials. Water temperature was held constant at 25°C. The barometric pressure was assumed to be 710 mm Hg to account for elevation above sea level of the VTAF. The u-tube $\alpha$ ratio was assumed to be 0.7 to account for water quality effects in all simulations unless otherwise noted. The $\alpha$ ratio for the 3-stage RBC was assumed to be 1.0 unless otherwise noted. The $\beta$ ratio was assumed to be 0.95 based upon the recommendation of Gilbert (1979).

Transfer costs were determined by $O_2$ costs and power costs alone and do not account for equipment or capital costs. The cost of $O_2$ was assumed to be $0.253 per kg and the cost of power assumed to be $0.09 per kWhr. The combined efficiency of the pump and the combined efficiency of the off-gas compressor were each assumed to be 0.35 when estimating total energy requirements.

The u-tube aeration system provides an OTR sufficient to meet $O_2$ demands
determined from daily feed ration. The OTR in the u-tube is primarily dependant upon GW ratio and flow rate. Simulations were run to determine the GW ratio required to provide an OTR to maintain predicted equilibrium tank DO concentrations between 6 and 7 mg/L. While u-tube performance is primarily governed by GW ratio and flow, the following simulations and discussions are based on daily feed ration which dictate OTR requirements and subsequently the GW ratio required to meet those requirements and maintain acceptable DO concentrations.

6.7.1. Effects of Flow Rate

Model simulations to investigate effects of recirculation flow rate on gas transfer performance indicate u-tube operation at low flow rates results in reduced oxygen transfer costs over the range of flows investigated (Fig 6.12). Increasing flow rate only slightly decreases the required GW ratio to meet a given O$_2$ demand but results in a rather large decrease in OTE, which in turn, requires higher mass rates of O$_2$ injection in the u-tube. Total costs per kg O$_2$ transferred increase with increasing flow rate and increasing feed ration, i.e. increasing GW ratio, and is attributed primarily to decreasing OTE.

Contact time between O$_2$ gas and water appears to influence OTE to a greater degree than changes in $K_{L,a}$ due to flow rate over the range of flows investigated here. Oxygen transfer efficiency appears to peak between 5 and 7 kg/day feed ration for all flow rates. Decreases in OTE below 5 kg/day may possibly be attributed to decreases in $K_{L,a}$ at very small required GW ratios rather than O$_2$ mole
Fig 6.12  Flow rate effects on required GW ratio, OTE, and total cost per kg $O_2$ transferred for increasing daily feed ration
fraction reduction in the gas bubbles. The predicted O₂ mole fraction in the off-gases increases only slightly from 0.86 at 3 kg/day to 0.87 at 15 kg/day for 250 L/min flow rate. The large fraction of O₂ in the off-gases indicates off-gas recycle would be beneficial in reducing transfer costs by increasing OTE.

Transfer costs increase primarily due to decrease in OTE but energy costs increase dramatically with increased flow rate at low feed rations. At a feed ration of 3 kg/day, energy costs per kg O₂ transferred increase nearly 200% if flow is increased from 250 L/min to 350 L/min (Table 6.12). Increasing the feed ration requires an increase in GW ratio which reduces the effect of flow rate on energy costs. Energy cost increases account for nearly 25% of the total cost increase. At 15 kg/day, energy costs increase approximately 80% when increasing flow from 250 L/min to 350 L/min but account for less than 8% of the total cost increase.

At 15 kg/day, energy costs comprise less than 3% of the total transfer cost for either flow rate due to increased AE. The AE at 250 L/min at 3 kg/day is 2.42 kg O₂ / kWhr and 0.87 kg O₂ / kWhr at 350 L/min while at 15 kg/day the respective AE's are 5.71 kg O₂ / kWhr and 3.14 kg O₂ / kWhr, respectively. Earlier researchers also found decreasing energy costs and increasing AE for increasing GW ratio (Speece and Orosco, 1969; Watten and Beck, 1985).

U-tube operation at lower flow rates results in a decrease in both O₂ and power costs over the range of flows simulated and appears more economical than higher recirculation flow rates. However, Fig. 6.5 suggests flow rate effects on stripping of CO₂ across the RBC may also need to be considered unless some
TABLE 6.12  Breakdown of total transfer costs for each combination of maximum and minimum flow rate and daily feed ration.

<table>
<thead>
<tr>
<th>Flow (L/min)</th>
<th>3 kg/day</th>
<th>15 kg/day</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>O2/kg</td>
<td>kWhr/kg</td>
</tr>
<tr>
<td>250</td>
<td>0.650</td>
<td>0.037</td>
</tr>
<tr>
<td>350</td>
<td>0.851</td>
<td>0.103</td>
</tr>
<tr>
<td>Increase</td>
<td>0.201</td>
<td>0.066</td>
</tr>
<tr>
<td>% Increase</td>
<td>30.9</td>
<td>176.9</td>
</tr>
<tr>
<td>% of total increase</td>
<td>75.3</td>
<td>24.7</td>
</tr>
</tbody>
</table>
other method is utilized for removing biological CO₂. Increasing flow rate from 250 to 350 L/min reduces predicted equilibrium CO₂ concentrations 20 to 25 % over the range of feed rations investigated (Fig 6.13). At a feed ration of 15 kg/day and flow rate of 250 L/min, the CO₂ concentration is approximately 32 mg/L but decreases to 24 mg/L at 350 L/min. Both concentrations exceed the recommended allowable maximum of 20 mg/L (Colt and Orwicz, 1991). However, some researchers found little effect on growth and health at CO₂ levels in excess of 50 mg/L (Smart et al., 1979). Therefore, there remains uncertainty as to what concentration CO₂ is detrimental to production, especially if DO and alkalinity levels are adequate.

Increased supplemental O₂ transfer by the RBC due to increased flow exerts little or no effect on total O₂ transfer costs of the U-tube. Decreasing flow rate results in a lower DO in the clarifier effluent (RBC influent) due to increased clarifier detention time. The predicted O₂ demand in the clarifier is 40 g/hr at 15 kg/day feed ration resulting in a predicted clarifier effluent DO of 3.7 mg/L at 250 L/min, 4.3 mg/L at 300 L/min, and 4.6 mg/L at 350 L/min. Thus, increased DO deficits in the RBC result in an equivalent supplemental OTR (43 g/hr) at each flow rate. The supplemental OTR at 15 kg/day is about 13 % of the total system oxygen demand or required OTR. Supplemental O₂ transfer in the RBC provides 26% of the total OTR at 250 L/min and increases slightly to 30% at 350 L/min but due to the decrease in u-tube performance, the increased OTR of the RBC is negligible.
FIG 6.13  Predicted equilibrium CO$_2$ concentration in production tank due to flow rate effects on CO$_2$ stripping in the RBC.
6.7.2. Effects of Equilibrium DO

Simulations were conducted to investigate the effect of equilibrium DO in the production tank on supplemental RBC O₂ transfer and required U-tube GW ratio. Increasing the equilibrium DO in the production tank would reduce the DO deficit of the RBC influent decreasing supplemental O₂ transfer of the RBC and increasing O₂ transfer requirements of the u-tube. Simulations were conducted at a flow of 280 L/min to determine the supplemental RBC OTR and required GW ratio at equilibrium DO concentrations of 6.5 and 8.5 mg/L (±0.2 mg/L). The u-tube α ratio was 0.7 and the RBC α ratio was 1.0.

The RBC provides supplemental O₂ transfer rates ranging from 22 to 42 g/hr over the range of feed rations providing 22% to 13% of the required OTR (Fig 6.14) at 6.5 mg/L DO in the tank. Increasing the equilibrium DO to 8.5 mg/L results in predicted RBC OTRs ranging from -5 g/hr at 5 kg/day to 16 g/hr at 15 kg/day. The RBC strips some O₂ provided by the u-tube at 5 kg/day and supplies less than 6% of the required total OTR at 15 kg/day (Fig 6.14).

Reductions in supplemental OTR require a small but relatively constant increase in required GW ratio (Fig 6.14). The increase in GW ratio appears small but daily total transfer costs increase 33% from $1.49 to $1.94 per day at 5 kg/day feed ration and approximately 13% at 15 kg/day from $ 5.02 to $ 5.70. Control of equilibrium DO within ±0.2 mg/L may be possible only in a computer model. However, these simulations indicate savings in O₂ transfer costs may be recognized when equilibrium DO concentrations are maintained slightly below saturation.
FIG 6.14 Percent of total required OTR supplied by RBC and resulting GW ratio which are required to maintain tank DO at approximately 6.5 and 8.5 mg/L.
6.7.3. Off-gas Recycle and Automated Control of GW Ratio

Watten and Beck (1985) demonstrated that OTE may be increased and O\textsubscript{2} transfer costs could be reduced through capture and recycle off-gases exiting with the u-tube effluent. Flow rate simulations predicted OTE of 30-40\% and mole O\textsubscript{2} fractions in off-gases in excess of 0.88 indicating off-gas recycle may reduce O\textsubscript{2} costs. Simulations were made for increasing feed ration at 0 and 75\% off-gas recycle rates to determine the PW ratio required to maintain equilibrium tank DO concentrations between 6 and 7 mg/L. Flow rate was 280 L/min, u-tube \( \alpha \) ratio = 0.7, and RBC \( \alpha \) ratio = 1.0.

Predicted OTE at 0\% recycle ranges from 36-33\% for increasing feed ration while 75\% recycle increases the range to 69-64\%. The PW ratio at 75\% recycle is reduced approximately 47\% at all feed rations (Fig. 6.15). Seventy-five off-gas percent recycle increases the effective GW ratio approximately 27\% over the GW ratio at 0\% recycle. The O\textsubscript{2} mole fraction in off-gases decrease from approximately 0.88 at 0\% recycle to 0.84 at 75\% recycle and N\textsubscript{2}+Ar mole fractions increase from 0.12 to 0.16.

Aeration efficiency increases with increasing feed ration for a given recycle rate due to an increase in the required PW ratio which results in increased OTR. However, AE decreases for a given feed ration at 75\% as compared to 0\% recycle due to an increases in the effective GW ratio (Fig. 6.16). Aeration efficiency decreases because OTR remains approximately the same while energy requirements increase. Energy costs increase due to increased two-phase flow headloss and off-gases compression for recycle. Thus, energy costs per kg O\textsubscript{2} are increased slightly for off-gas recycle but are offset by reduced OTE.
Fig. 6.15 Required GW ratios to maintain tank DO concentrations between 6 and 7 mg/L for 0 and 75% off-gas recycle.
FIG. 6.16  Decrease in AE effected by increase in effective GW Ratio due to off-gas recycle
Seventy-five percent recycle decreases transfer costs approximately 45% due to increased OTE. Costs per kg O₂ transferred range from $0.74-0.78 at 0% recycle and $0.41-0.42 at 75% recycle. Predicted daily total transfer costs are provided in Fig. 6.17 computed as the product of total cost per kg O₂ transferred and the mass of O₂ transferred per day.

A simple method for control of the GW ratio to adjust u-tube OTR in conjunction with daily demand cycles was investigated to estimate possible reductions in transfer cost. The method proposed involves a timer that toggles a second injection line connected to the U-tube open or closed increasing or decreasing the GW ratio by a preset amount. The primary injection line operates continuously providing the GW ratio required to meet daily average requirements. The second line is opened by the timer at approximately 9:00 AM daily (just before feeding, Fig 6.7) increasing the GW ratio to meet maximum demands and maintain DO levels between 6 and 7 mg/L. The timer then closes the second injection line at 6:00 PM as demands decrease to minimum levels.

Daily O₂ transfer costs were calculated by assuming u-tube operation for 15 hr/day (6:00 PM until 9:00 AM next day,) at the GW ratio to meet minimum demands and 9 hr/day at the GW ratio to meet maximum demands (See Fig. 6.7). Transfer costs per kg O₂ at each feed ration were approximately equal for timer controlled operation and u-tube operation at the maximum GW ratio. However, the total amount of O₂ injected daily was reduced approximately 20% thus reducing total daily transfer costs (Fig. 6.17).
Fig. 6.17 Daily total O₂ transfer costs predicted for U-tube operation at 0% off-gas recycle, 75% recycle, and timer controlled operation.
6.7.4. **Effects of Water Quality**

Model simulations were made to investigate water quality effects on RAS system performance by decreasing u-tube α ratio from 1.0 to 0.5 for increasing feed ration. The GW ratio required to maintain tank DO concentrations between 6 and 7 mg/L was determined for each combination of α and feed ration (Fig. 6.18). The data in Fig. 6.18 exhibit some variance or scatter rather than forming a straight line since a small range of GW ratio for each combination of α and feed ration would provide a equilibrium DO between 6 and 7 mg/L. If a value of the GW ratio were determined at each feed ration which provided exactly the same equilibrium tank DO at constant α values, the data would exhibit stronger linearity.

Oxygen transfer efficiency decreases for increasing feed ration at constant α, as was observed in earlier simulations, but the effects on OTE due to decreasing α result in substantial increases in the required GW ratio to meet O₂ demands (Fig 6.18). Reduction in OTE for constant α values due to an increase in feed ration from 5 to 15 kg/day is less than 10 % for all α (Table 6.13). However, the average reduction in OTE relative to α = 1.0 for α = 0.85, 0.7, and 0.5 are 11, 22, and 40 %, respectively, illustrating the effect of water quality on OTE, and subsequently O₂ costs. The range of OTE for α=0.85 to 0.7, 42 % to 33 %, is representative of the range of OTE routinely established in the VTAf u-tube aerators under production conditions.
Fig. 6.18  Estimated GW ratio required to maintain tank DO between 6 and 7 mg/L for various \( \alpha \) values due to increasing feed ration. Oxygen demands determined from feed ration were maximum daily demands.
Table 6.13  Predicted OTE due to decreasing $\alpha$ and increasing daily feed ration.

<table>
<thead>
<tr>
<th>Feed Ration kg/day</th>
<th>$\alpha$</th>
<th>OTE %</th>
<th>OTE %</th>
<th>OTE %</th>
<th>OTE %</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>1</td>
<td>0.85</td>
<td>0.7</td>
<td>0.5</td>
<td></td>
</tr>
<tr>
<td>5</td>
<td></td>
<td>46.4</td>
<td>41.6</td>
<td>36.3</td>
<td>28.1</td>
</tr>
<tr>
<td>7</td>
<td></td>
<td>45.7</td>
<td>41.6</td>
<td>36.3</td>
<td>27.9</td>
</tr>
<tr>
<td>9</td>
<td></td>
<td>46.1</td>
<td>41.4</td>
<td>35.6</td>
<td>27.4</td>
</tr>
<tr>
<td>11</td>
<td></td>
<td>45.2</td>
<td>40.4</td>
<td>35.0</td>
<td>26.9</td>
</tr>
<tr>
<td>13</td>
<td></td>
<td>44.2</td>
<td>39.1</td>
<td>33.9</td>
<td>26.1</td>
</tr>
<tr>
<td>15</td>
<td></td>
<td>43.9</td>
<td>38.4</td>
<td>33.4</td>
<td>25.5</td>
</tr>
</tbody>
</table>
6.7.5. Effects of RBC Nitrogen Reabsorption

Model simulations to investigate effects of flow rate and equilibrium DO on system performance indicated that RBC reaeration could provide sufficient \( O_2 \) transfer to reduce u-tube transfer requirements as much as 10-30 % under proper operating conditions. In addition to providing supplemental \( O_2 \), RBC reaeration replaces a portion of the \( N_2+Ar \) removed from the water during u-tube oxygenation. U-tube \( N_2+Ar \) stripping rates increase with increasing GW ratio at a given recycle rate but mole fractions of \( O_2 \) and \( N_2+Ar \) remain relatively constant and mole fractions of \( O_2 \) are reduced and \( N_2+Ar \) increases by employing off-gas recycle. Under equilibrium conditions, \( N_2+Ar \) stripped in the u-tube is replaced by the RBC establishing equilibrium concentrations within each component's effluent. Watten et. al, (1991) demonstrated \( N_2 \) desorption within a pure \( O_2 \) packed column adsorber substantially reduced OTE in a column operating at 1 atm pressure and influent dissolved nitrogen concentrations greater than 70 %.

Nitrogen and Argon stripping reduces the mole fraction of \( O_2 \) in the gas phase decreasing effective \( O_2 \) deficits in the liquid phase. Therefore, higher GW ratios in the adsorption unit were required to provide a given exit DO resulting in higher operating costs.

The design parameter controlling \( N_2+Ar \) transfer in the RBC is \( K_{l,a} \). Decreasing \( K_{l,a} \) would result in lower equilibrium \( N_2+Ar \) concentrations in the RBC effluent (u-tube influent) since the deficit must increase to match u-tube stripping rates. Lower \( N_2+Ar \) concentrations in the u-tube influent may increase u-tube OTE by
establishing higher O$_2$ mole fractions and lower N$_2$+Ar mole fractions in the gas bubbles.

Simulations were conducted at 0 % and 90 % off-gas recycle for decreasing RBC $K_{L,a}$ values to investigate effects on required GW ratio, u-tube performance, and equilibrium concentrations of N$_2$+Ar in the RAS system. Operating conditions were identical to those for off-gas recycle simulations with the following exceptions. Daily feed ration was held constant at 11 kg/day. The RBC was modeled as a single stage unit rather than a 3 stage to investigate the effects of varying a single $K_{L,a}$ value rather than a set of values. A single stage RBC would require a $K_{L,a}$ of approximately 1.2 min$^{-1}$ to provide an equivalent transfer rate obtained by the VTAF three-stage unit operating with a through put flow rate of 280 L/min. Therefore, simulations were conducted to determine the GW ratio required to maintain equilibrium tank DO between 6 and 7 mg/L for $K_{L,a}$ ranging from 1.5 to 0.005 min$^{-1}$.

Decreasing RBC stage $K_{L,a}$ below about 0.25 min$^{-1}$ lowers equilibrium dissolved N$_2$+Ar concentrations in the u-tube influent and increases the O$_2$ mole fraction in u-tube off-gases (Fig 6.19). Increases in O$_2$ mole fractions in the u-tube due to decreasing RBC $K_{L,a}$ are larger at 90 % off-gas recycle, however, effects on u-tube performance for either 0 or 90 % recycle are negligible. U-tube oxygen transfer efficiency increases slightly over the range of decreasing RBC $K_{L,a}$ values, however, reductions in supplemental O$_2$ transfer from the RBC increases the u-tube GW ratio required to maintain equilibrium DO concentrations in the tank (Table 6.14). Required PW ratios increase 17 % at
Fig. 6.19 Effect of off-gas recycle on dissolved N\textsubscript{2}+Ar concentration in u-tube influent as a percent of saturation due to decrease in RBC stage $K_{L}a$ and resulting O\textsubscript{2} mole fraction in off-gases exiting in u-tube effluent.
**TABLE 6.14** Summary of RAS O$_2$ transfer performance as a function of off-gas recycle for decreasing RBC $K_La$ values. Total daily costs ($/day) are computed as the product of the cost per kg O$_2$ transferred ($/kg$) and total daily kg of O$_2$ transferred.

<table>
<thead>
<tr>
<th>RBC $K_La$</th>
<th>U-tube 0% recycle</th>
<th>U-tube 90% recycle</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>PW OTE $/kg$ $$/day$</td>
<td>PW OTE $/kg$ $$/day$</td>
</tr>
<tr>
<td></td>
<td>Ratio</td>
<td>Ratio</td>
</tr>
<tr>
<td>1.5</td>
<td>0.0240 35.2 0.748 3.38</td>
<td>0.0105 81.7 0.346 1.58</td>
</tr>
<tr>
<td>1.0</td>
<td>0.0245 35.3 0.746 3.45</td>
<td>0.0105 82.1 0.344 1.58</td>
</tr>
<tr>
<td>0.5</td>
<td>0.0255 35.4 0.742 3.58</td>
<td>0.0110 82.3 0.342 1.65</td>
</tr>
<tr>
<td>0.25</td>
<td>0.0265 35.6 0.738 3.72</td>
<td>0.0115 82.6 0.339 1.72</td>
</tr>
<tr>
<td>0.10</td>
<td>0.0270 36.1 0.727 3.79</td>
<td>0.0118 83.4 0.334 1.75</td>
</tr>
<tr>
<td>0.05</td>
<td>0.0275 36.2 0.725 3.86</td>
<td>0.0119 83.9 0.331 1.76</td>
</tr>
<tr>
<td>0.005</td>
<td>0.0280 36.3 0.723 3.93</td>
<td>0.0120 84.8 0.325 1.77</td>
</tr>
</tbody>
</table>
0 % recycle and 14 % at 90 % recycle when $K_{L}a$ is decreased from 1.5 to 0.005 min$^{-1}$. Total costs per kg $O_2$ transferred decrease slightly due to increasing OTE and decreasing energy costs. The increased OTE appears insufficient to compensate for decreases in supplemental $O_2$ transfer and daily total costs increase due to increases u-tube $O_2$ transfer requirements.

Since the transfer capacity of the RBC is reduced, the equilibrium $CO_2$ concentration in the production tank increases due to decreased $CO_2$ stripping. The equilibrium $CO_2$ concentration in the production tank at $K_{L}a = 1.5$ min$^{-1}$ is approximately 21 mg/L. Decreasing $K_{L}a$ to 0.25 min$^{-1}$ increases the equilibrium $CO_2$ concentration over 100 % to approximately 45 mg/L. At $K_{L}a$ below about 0.1 min$^{-1}$, predicted $CO_2$ concentrations exceed 100 mg/L possibly becoming detrimental to fish production.
7. SUMMARY

Experimental tests designed to validate the RAS gas transfer model verified that the model simulates gas transfer in a recirculating aquaculture system. The model predicted equilibrium DO concentrations in the production tank within ±0.7 mg/L (±15 %) under simulated production conditions. The largest relative errors (±48 %) occurred in prediction of the multi-tube clarifier DO when actual DO concentrations were less than 4 mg/L and absolute errors were ±1.0 mg/L. The largest absolute error (- 3.0 mg/L) occurred in prediction of u-tube effluent DO.

The u-tube component model was validated in previous research and found to adequately simulate the effects of off-gas recycle for differing U-tube depths, flow rates, and GW ratios. Development of a predictive equation for $K_{L}a_{20}$ was required; however. The model was assumed to adequately simulate effects off-gas recycle and additional tests for recycle validation were not performed.

Validation tests indicated that predicted equilibrium DO in all components of the RAS depends primarily on accurate calibration of the u-tube component model. An initial u-tube calibration was erroneous and resulted in large errors in predicting u-tube OTR. Errors in prediction of u-tube OTR were found to magnify errors in prediction of component DO in the production tank and multi-tube clarifier. A second calibration equation developed upon a limited set of data (4 data points) were taken from a portion of the validation test data which were more characteristic of production conditions the u-tube calibration run conditions,
proved more successful. The remaining validation data were then used for model validation.

The second u-tube calibration equation reduced errors in prediction of u-tube OTR resulting in substantial improvement in prediction of remaining component DO concentrations. The smallest errors occurred in predication of RBC effluent DO due to the nature of RBC gas transfer in which effluent DO asymptotically approaches the atmospheric saturation value for given conditions. Model validation was considered successful since model predictions for equilibrium tank DO were within ± 0.7 mg/L and prediction of equilibrium dissolved N₂+Ar concentrations were within ± 1.4 mg/L. However, this equation also under predicted OTR in the u-tube. If another inlet Kₜ,a calibration equation were developed from a more extensive set of performance data, errors between model predictions and the validation data may be reduced even further.

The RAS gas transfer model attempts to simulate the combined effects of gas transfer and biological respiration in a recirculating aquaculture system by mass balance analysis rather than predicting gas transfer alone. While the model has been demonstrated to correctly simulate gas transfer processes, the equivalent cannot be stated about simulation of biological respiration. Biological respiration is influenced by many factors which are not considered in the model and the model only provides an estimation of oxygen demand rate based on daily feed ration. Oxygen demand in these systems while strongly correlated to feed input, is widely variable as seen from the actual biological demands estimated in this research.
Techniques provided in the literature for estimating biological O₂ demands of the fish were shown to adequately predict demands in the range of those observed during the 1991 growth trials. The value of the oxygen feed ratio, OFR = 0.3, estimated from production data is within the range of values, 0.25-0.4, provided in the literature for warmwater culture. The value of the peaking factor, 1.44, determined from resting and midday post feeding O₂ demands is also in agreement with the value recommended in the literature. No techniques were available from the literature for estimating O₂ demands in the clarifier. Therefore, a regression equation to predict clarifier O₂ demand from daily feed ration was developed from measured clarifier demands. Demands measured in the clarifier often accounted for over 10% of the total RAS O₂ demand, (tank + clarifier), estimated under production conditions during the growth trials. Thus, clarifier O₂ demands are considered essential when estimating total system O₂ consumption.

The effects of water quality on transfer performance are often unknown or difficult to establish. The unsuccessful attempt at estimation of α ratios for u-tube operation under production conditions indicates the difficulty in separating water quality effects on oxygen transfer performance from effects of operational parameters such as flow rate or GW ratio. However, the effect of water quality due to variation in α was shown by model simulation to strongly influence O₂ transfer performance of the u-tube and subsequently O₂ transfer costs. Substantial reductions of OTE to meet a given oxygen demand occur due to decreases in α. The predicted reduction in OTE for α = 0.7, a value concluded to be representative of production conditions, was nearly 22%.
Model simulations were conducted to assess effects on performance and O₂ transfer costs of several operational parameters and several methods to improve O₂ transfer performance. Recirculation flow rate was demonstrated to influence u-tube OTE and O₂ transfer costs. Decreasing flow from 350 to 250 L/min decreased total transfer costs by 25-30 % but increased equilibrium CO₂ concentrations approximately 20-25%. Equilibrium CO₂ concentrations under conditions of low flow rate and high respiration rates were predicted to exceed 30 mg/L. The U-tube component model predicted a 40 % decrease in O₂ costs for off-gas recycle rate at 75% compared to operation with no recycle. The model also predicted that a 20% reduction in costs may be obtained if u-tube operation were controlled by a timer switch operating in conjunction with maximum and minimum O₂ demand cycles exhibited in the RAS. Reducing equilibrium DO maintained in the production tank from 8.5 mg/L to 6.5 mg/L substantially increased supplemental O₂ transfer of the RBC. Reducing the equilibrium DO to 6.5 mg/L reduced u-tube O₂ transfer requirements 13-22 % and O₂ transfer costs 13-33 % depending upon O₂ demands.

The combined effects O₂ and N₂+Ar absorption in the RBC were shown to affect O₂ transfer costs. Decreasing RBC reaeration rates by decreasing RBC K₄a reduced both supplemental O₂ transfer and reabsorption of N₂+Ar stripped in the u-tube. Reducing RBC K₄a slightly increased u-tube OTE by reducing the N₂+Ar mole fraction in gas bubbles in the u-tube. Despite slight increases in OTE, u-tube GW ratios required to meet RAS O₂ demands were increased due to decreases in supplemental O₂ transfer from the RBC and resulted in increased total O₂ costs.
The model is intended to provide only general quantification of dissolved gas concentrations and system transfer performance rather than exact values. The variability in model parameters such as respiration or $\alpha$ ratio results in some uncertainty in model predictions. However, the model can be used to establish relative trends in RAS component effects on gas transfer and consumption processes. Numerical or graphical representation of such trends can aid in qualifying component performance and provide reasonable estimates of operational costs for $O_2$ transfer, or assessment of proposed changes in operational or design parameters.

The RAS model was calibrated primarily with design parameters such as $K_t a$ measured at the VTAF in RAS configuration located there. The calibration equations developed for estimation of RBC and u-tube $K_t a$ as well as multi-tube clarifier oxygen demand make the model somewhat site specific. However, the modular programming structure allows procedures simulating gas transfer and/or respiration for other types of component's to be incorporated into or deleted from the overall RAS model or calibration values for existing components can be input directly. This programming structure offers the flexibility to analyze an almost unlimited combination of RAS components. Coupled with such flexibility, the RAS Gas Transfer Model's ability to simulate the combined effects of multi-component gas transfer and biological respiration provides a useful tool in assessing gas transfer and oxygenation performance in recirculating aquaculture systems.
8. CONCLUSIONS

1. Validation was successful and the model adequately simulated gas transfer in the all components of the RAS.

2. The model provides the ability to simulate changes in system gas transfer performance due to changes in RAS component operation, configuration, or system O$_2$ demands.

3. The model provides a tool for accounting for water quality effects on O$_2$ transfer performance and provides realistic representations of performance efficiencies expected under production conditions.

4. The biological data collected for O$_2$ demand in the tank and clarifier indicated that oxygen demands could be estimated from daily feed inputs into the RAS. This provides a simple model input from which total system O$_2$ demand can be estimated.

5. Model simulations indicate that O$_2$ transfer costs at the VTAFF may be reduced substantially by one or more of the following:

   a. Reducing u-tube flow rate. Costs may be reduced as much as 25 % by reducing flow from 350 to 250 L/min and 15 % by reducing flow from 300 to 350 L/min.

   b. Maintaining equilibrium DO in the production tank below saturation. Costs may be reduced as much as 13-30% due to increased supplemental O$_2$ transfer in the RBC.

   c. Installing off-gas recycle. Seventy-five percent off-gas recycle may reduce transfer costs as much as 44%.

   d. Controlling u-tube oxygen transfer in sequence with RAS demand cycles. Costs may be reduced as much as 20 % with a simple timer increasing or decreasing u-tube GW ratio on a set daily schedule.
6. Accounting for water quality effects on $K_{La}$ and saturation values estimated from clean water tests was shown to reduce u-tube OTE as much as 11% at $\alpha = 0.85$ and 22% at $\alpha = 0.7$ when $\beta = 0.95$.

7. The RBC was shown to be useful in maintaining equilibrium CO$_2$ concentrations at acceptable levels in the RAS as well reducing OTR requirements of the u-tube.

8. The RAS model's modular program style allows simple integration or removal of subroutines representing individual process components. This structure allows investigation of alternative system components providing preliminary design analysis data or specification of O$_2$ transfer requirements when selecting commercial oxygenation devices.
9. RECOMMENDATIONS

The research results presented shows the potential of the model to simulate gas transfer in the VTAF Recirculating Aquaculture Systems. The following recommendations for future investigation are suggested to support or supplement the information presented in this thesis.

Good prediction of RAS equilibrium DO’s was observed for the validation tests using the second u-tube calibration. However, this equation was developed from four validation data points for $K_{\text{ia}}$ taken from a possible set of 12 and the remaining utilized for validation. Another U-tube calibration based on more values of $K_{\text{ia}}$ over a larger range of flows, and GW ratios, may provide improved model performance and better validate performance.

Total daily O$_2$ transfer costs were projected by model simulations over a range of feed input rations for typical operating conditions of $\alpha$ ratio, flow, and maximum daily oxygen demand. These data could be used to estimate oxygenation costs per kg of fish produced over an entire grow out period and compared to actual total oxygenation costs incurred during the 1991 growth trials. Assuming model predictions compare favorably with actual cost estimates, such a comparison would further support the RAS model validation and emphasize the model's potential for O$_2$ transfer costs analysis.

The literature indicate that other oxygenation devices may the provide required O$_2$ transfer rates at significantly lower costs than the u-tube. Several numerical
models of some of these devices have been developed and validated by other researchers. If these models are similar in modeling approach to the RAS component models, they could easily be incorporated into the RAS gas transfer model. Thus, simulations could be conducted as if the u-tube were replaced by an alternative oxygenation device providing data for performance and cost comparisons.
Literature Cited


Colt, John and Barnaby J. Watten. 1988. *Applications of pure oxygen in fish culture.* Aquacultural Engineering. 7:397-441.


APPENDIX A

Instrumentation Calibration Procedures
Calibration Procedure for Rotameter

The rotameter utilized for metering oxygen at the low to mid-range oxygen flows, 0.5-7 standard liters per minute (SLPM) was the Matheson Model 757. These rotameters were placed in each of the recirculating systems. A customized calibration for each rotameter was not specified at the time of purchase. The manufacturer provided a general calibration table to provide approximate flows for the particular model at standard conditions. The rotameter was designed to operate simultaneously utilizing two metering balls of different densities, glass and stainless steel, to provide volumetric gas flow measurement over two flow ranges. If an oxygen flow was established through the rotameter which suspended both balls such that a flow reading could be made for each, the flows indicated by the manufacturer's tables often differed by as much as 10 to 15 %. The accuracy of the tables, ±10 % as suggested by the manufacturer, was determined to be inadequate for calibration requirements of the U-tube submodel and therefore a customized calibration table was developed.

The following procedure was used to provide calibration data.

Materials Required. Large graduated cylinder, size is dependant upon flow range of rotameter and the accuracy required. A cylinder of 10 L was used in this calibration with a graduation scale of 50 mL. A large container filled with water. The container should be sufficiently deep and wide to allow the graduated cylinder to be filled with water and inverted without air entrainment. A small amount of red or green food coloring added to the water allows easier reading of collected gas volumes in the graduated cylinder. A stopwatch is also required. A length of flexible tubing connected to the exit line of the rotameter will make the procedure less cumbersome. While one person may be able to run the calibration, the procedure is much more easily performed by two persons.
**Procedure:** Submerge the graduated cylinder into the large container and invert the graduated cylinder so that it is filled with water without any gas bubbles collected in the top (cylinder base or bottom). Establish a gas flow through the rotameter with the exit hose several inches below the water surface in the large container but do not allow bubbles to enter the graduated cylinder. Record the glass ball reading (GBR). Position the cylinder so that the submerged rim (top edge) is near the exit hose. Simultaneously start the stopwatch and place exit hose under the graduated cylinder so that all gas bubbles are collected in the top of the graduated cylinder. Care must be taken not to move the position of the exit hose vertically in the water column as the glass ball reading will vary considerably with changing hydrostatic pressure. A constant glass ball reading should be maintained throughout gas collection. Collect exit gas for a specified time or volume and record both. Once the exit hose is removed the volume of collected gas should be read by positioning the cylinder vertically such that the water level in the cylinder is at the same level as in the larger container. This must be done to assure that gas pressure inside the container is essentially equal to the atmospheric pressure. The procedure should be repeated several times for each GBR over a range of readings to estimate measurement accuracy of the resulting calibration. The local barometric pressure and gas temperature must be measured in order to correct the gas volume to Standard Conditions. The gas volume can be corrected by the Ideal Gas Law due to the low pressures involved. Once the collected gas volume is calculated and corrected, the volumetric flowrate in SLPM can be determined for a particular GBR by dividing the volume by the time recorded for gas collection.

The above procedure was performed using the Matheson rotameter to develop a calibration table over a range of glass ball readings (Table A1). This table was then utilized to determine all gas flow measurements for the Matheson rotameter during model calibration and validation testing. The repeatability of the volumetric flow rates obtained for a given GBR was within ±1%. Slight inaccuracies may occur due to addition of water vapor to the collected gas. The partial pressure of water vapor in water saturated air at 20° C is 17.5 mm Hg, approximately 2.3 % of the total pressure at 1 atm (760 mm Hg). The vapor
pressure in the collected gas volumes was not measured but was assumed to be less than the saturation value due to the short contact time between the oxygen bubbles and the water. Since the vapor pressure was not known, the gas volumes collected during calibration were not corrected for addition of water vapor. The accuracy of the calibration would be within ±5% of the full-scale reading even if the collected gases were water vapor saturated and not corrected.

A formula provided by the manufacturer for correcting gas flows due metered at conditions other than other than standard conditions of 1 atm and 21.1 °C is provided below. The calibration table provides two sets of flows for a given GBR (Table A1). The second set, *SLPM*, corresponds to the volumetric gas flow in SLPM at the approximate metering pressure at the U-tube inlet on the recirculating aquaculture systems during normal production. This formula was used to correct gas flows at conditions of U-tube inlet absolute pressures ranging from 1.07 to 1.2 atm during calibration of the U-tube submodel

\[ *\text{SLPM} = \frac{\text{SLPM}}{\sqrt{\frac{T_c}{21.1^\circ\text{C}}} \cdot \frac{1}{P}} \]

where
*SLPM* - Equivalent flow at Standard Conditions for a gas metered at exit pressure, P and temperature, T_c.
P - Exit pressure, atm
T_c - Gas temperature, °C
TABLE A1. Calibration table developed for Matheson Rotameter using water calibration procedure for a range of glass ball readings (GBR). Values labeled as SLPM are flows at standard conditions. Values labeled as *SLPM* flows to exit conditions of 1.07 atm absolute pressure (0.76 m H₂O gauge) which is approximated metering pressure in the u-tubes at the VTAF under production conditions.

<table>
<thead>
<tr>
<th>GBR</th>
<th>SLPM</th>
<th><em>SLPM</em></th>
</tr>
</thead>
<tbody>
<tr>
<td>10</td>
<td>0.37</td>
<td>0.38</td>
</tr>
<tr>
<td>15</td>
<td>0.64</td>
<td>0.67</td>
</tr>
<tr>
<td>20.5</td>
<td>0.95</td>
<td>0.98</td>
</tr>
<tr>
<td>25</td>
<td>1.25</td>
<td>1.29</td>
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<tr>
<td>30</td>
<td>1.59</td>
<td>1.64</td>
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<tr>
<td>35</td>
<td>1.91</td>
<td>1.98</td>
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<tr>
<td>40</td>
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<td>2.98</td>
</tr>
<tr>
<td>50</td>
<td>3.03</td>
<td>3.14</td>
</tr>
<tr>
<td>55</td>
<td>3.19</td>
<td>3.3</td>
</tr>
<tr>
<td>59.5</td>
<td>3.32</td>
<td>3.44</td>
</tr>
<tr>
<td>65</td>
<td>3.65</td>
<td>3.78</td>
</tr>
<tr>
<td>70</td>
<td>3.95</td>
<td>4.09</td>
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<tr>
<td>75</td>
<td>4.25</td>
<td>4.4</td>
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<tr>
<td>80</td>
<td>4.54</td>
<td>4.71</td>
</tr>
<tr>
<td>90</td>
<td>5.12</td>
<td>5.31</td>
</tr>
<tr>
<td>95</td>
<td>5.41</td>
<td>5.6</td>
</tr>
<tr>
<td>100</td>
<td>5.7</td>
<td>5.9</td>
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<tr>
<td>105</td>
<td>5.95</td>
<td>6.17</td>
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<td>110</td>
<td>6.2</td>
<td>6.43</td>
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<td>115</td>
<td>6.45</td>
<td>6.68</td>
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<td>120</td>
<td>6.69</td>
<td>6.93</td>
</tr>
<tr>
<td>125</td>
<td>6.99</td>
<td>7.24</td>
</tr>
<tr>
<td>130</td>
<td>7.24</td>
<td>7.51</td>
</tr>
</tbody>
</table>
Water Flow Meter Calibration

Water flows were measured with the Omega Instruments Model 58 paddle wheel flow transmitter, a 5.08 cm flow cell, and a Model 64 Rate analyzer (Omega Instruments, Co. New York, N.Y.). The signal output from the Model 58 is a 4 to 20 mA current proportional to flowrate. This current signal was passed through a 100 ohm resistive circuit to generate an output voltage. The output voltage was measured and recorded by the data logging and control module.

Calibration of the flow meter was performed by the manufacturer but a calibration test indicated the calibration was poor. Therefore, the flow meter was calibrated over a flow range of 150 to 350 L/min. The calibration procedure consisted of filling a container to a known volume at a constant flow. The flow rate was determined by recording the time to fill the container. The output voltage was sampled and recorded on the data logger over the length of time required to perform a test for a single flow rate. The measured flow rates were then regressed against the average output voltage to develop a transform equation for predicting flow from voltage output. The plot of the transform equation and the calibration values are given in Fig. A1. Several tests were also conducted to evaluate the accuracy of the equation. The resulting equation predicted the flow rate within ±1.5 % of the measured value as shown in Table A2.
Fig. A1  Calibration data for Model 58 flow transmitter. The manufacturer's calibration represents the transform equation that would be developed from the manufacturer's voltage setpoints for zero and full scale flow rate.

Table A2  Prediction evaluation of calibrated Model 58 flow transmitter. Evaluations were made at time of calibration and later during experimental trials for RAS model calibration.

<table>
<thead>
<tr>
<th>Date</th>
<th>mV</th>
<th>Measured (L/min)</th>
<th>Predicted (L/min)</th>
<th>Residual (L/min)</th>
<th>% err</th>
</tr>
</thead>
<tbody>
<tr>
<td>6/31/91</td>
<td>1266</td>
<td>248</td>
<td>244</td>
<td>3.8</td>
<td>1.53</td>
</tr>
<tr>
<td>6/31/91</td>
<td>1126</td>
<td>203</td>
<td>205</td>
<td>2.2</td>
<td>-1.06</td>
</tr>
<tr>
<td>6/31/91</td>
<td>1657</td>
<td>357</td>
<td>353</td>
<td>3.7</td>
<td>1.04</td>
</tr>
<tr>
<td>8/7/91</td>
<td>1175</td>
<td>218</td>
<td>219</td>
<td>0.6</td>
<td>-0.29</td>
</tr>
<tr>
<td>8/7/91</td>
<td>1165</td>
<td>215</td>
<td>216</td>
<td>1.2</td>
<td>-0.57</td>
</tr>
<tr>
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<td>1289</td>
<td>252</td>
<td>251</td>
<td>1.7</td>
<td>0.66</td>
</tr>
<tr>
<td>8/7/91</td>
<td>1488</td>
<td>304</td>
<td>306</td>
<td>2.2</td>
<td>-0.71</td>
</tr>
</tbody>
</table>
Gasometer Evaluation

Three gasometers were constructed by shop personnel at the Department of Agricultural Engineering of Virginia Polytechnic Institute and State University. Parts list and procedures for construction are provided by Bouck (1980). The gasometer offers several principal advantages over the Weis Saturometer when measuring change in dissolved gas pressure, $\Delta P_m$. Both devices utilize the membrane diffusion method for $\Delta P_m$ measurement. However, the Weis Saturometer requires a significant amount of time *in situ* to obtain a representative reading; 15 - 30 minutes, while the gasometer can obtain readings in as little as 7 to 20 minutes depending on conditions (Bouck, 1980). The time required to obtain a final reading is directly proportional to the total surface area of a gas permeable tubing exposed to the sample and to the total internal volume which must be pressurized. The gasometer provides an extended length of tubing exposed to sample which substantially increases the surface area with only a negligible increase in internal volume. The saturometer requires constant operator attention to provide sample flow over the gas permeable tubing. At low positive $\Delta P_m$, the saturometer may be prone to significant errors due to bubble formation on the tubing if care is not taken to provide sufficient flow or hydrostatic pressure by lowering the unit deep enough in the water column (Bouck, 1980). The gasometer provides flow over the gas permeable tubing with a submersible pump or it may be installed directly into a pressurized water line. Hydrostatic pressure and high flow velocity prevent bubble formation and help to decrease the time required to obtain a pressure measurement. The gasometers constructed for use at

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the VTAF utilize mercury manometers for pressure measurement. Therefore, calibration was not theoretically required. The saturometer utilizes a spring pressure gauge for pressure measurement and may be prone to errors.

All three of the gasometers were tested against the saturometer for measurement of $\Delta P_m$. Two of the gasometers operated correctly and appeared to provide accurate measurements (Fig A2.). The third appeared to have a leak in the gas permeable tubing as no $\Delta P_m$ was measured. Gasometer #1 was constructed so that the gas permeable tubing was directly connected to the mercury manometer tubing and provide 95% of the final reading within 6 minutes under the conditions tested. Gasometer #2 contained a pressure release toggle valve located inline between the manometer and gas permeable tubing. The valve allows the unit to be rezeroed quickly to atmospheric pressure after a measurement is taken. Gasometer #2 required a substantially longer time to obtain a 95% of final reading due to the increased internal volume provided by the valve. Measurement of $\Delta P_m$ agreed within 5% between the gasometers but the saturometer measured $\Delta P_m$ nearly 10% greater under the conditions tested. The gasometers were assumed to provide more accurate measurements of $\Delta P_m$ due to the construction design and utilization of a true mercury manometer.
Fig A2  Evaluation of gasometers for measurement of ΔP by comparison to Weis Saturometer.  Water temperature - 25° C  DO - 5.3 mg/L
APPENDIX B

Computer Model Code Listing
"THIS PROGRAM MODELS GAS TRANSFER WITHIN A RECRIRCULATING
'AQUACULTURE SYSTEM WHEN GIVEN DESIGN AND OPERATING VALUES
'FOR THE FOLLOWING:

' TANK
' UTUBE AERATOR
' 3 STAGE RBC
' MULTI-TUBE CLARIFIER
' STOCKING DENSITY AND DAILY FEED RATION
' RECRIRCULATION RATE

' Written by Luther G. Wood, OCTOBER 1991
' Aquaculture Research Facility
' Dept. of Agricultural Engineering and Dept of Fisheries and Wildlife
' Virginia Polytechnic Institute and State University

'BEGIN MAIN PROGRAM CODE

'BEGIN Procedure Declarations

DECLARE SUB AIRSAT ()
'Determines atmospheric saturation concentrations
DECLARE SUB BACKGROUND ()
'Performs preliminary background calculations needed during simulation
'which remain onstant during any 'simulation run.
DECLARE SUB DISPLAY ()
'Prints model description and author information to the screen
DECLARE SUB INPARAMETERS ()
'Inputs user defined system design and operating data
DECLARE SUB SWITCH (SWX!)
'Switches user inputed value from dummy variable to input variable
DECLARE SUB SETUP ()
'Assigns initial dissolved gas concentrations 'for first array elements for any simulation 'run
DECLARE SUB REPORT ()
'Provides output format menu and initiates output printing
DECLARE SUB SHORTSUM ()
'Summarizes and prints an abbreviated
DECLARE SUB UTUBE () 'model output report
DECLARE SUB UTUBESUM () 'Models U-tube performance
DECLARE SUB OFFGAS () 'Summarizes equilibrium exit conditions
DECLARE SUB SHORTOFFGAS () of U-tube effluent and prints output
DECLARE SUB TANK () 'Analyzes off-gases from U-tube and
directs transfer performance
DECLARE SUB TANKSUM () 'Provides brief output report on u-tube gas
DECLARE SUB PH (CxF, PHYD) transfer performance
DECLARE SUB SUMP () 'Models transfer and consumption in tank
DECLARE SUB SUMPSUM () 'Summarizes equilibrium conditions in
tank and prints output
DECLARE SUB RBC () 'Predicts final tank pH based on total
DECLARE SUB RBCSUM () 'alkalinity and equilibrium carbon dioxide
DECLARE SUB PRESSURE (ExF, FxF, CxF, TGP, O2T, N2T, DPO2, DPN2, TODP) concentration
DECLARE SUB ASCFLE () 'Models oxygen consumption in solids-
' END Procedure Declarations settling

'Following Variables are declared as Global Variables
'These global variables are user input or are optionally determined within input
'subroutine INPARAMETERS
COMMON SHARED E, F, CD, Q, C, D, Z, AIK, VOLT, KLOT, KLCT, DEMt, FEED, VOLS, DEMs, A, B, ACF, BCF, I, kN,kC, HWC, OGR, OGCHL, CSTKG, CKWH, AEC, LID, SID, SOD, KLO1, KLO2, KLO3, VOLR, EFFc, EFFp

'These global variables are determined within PROCEDURE SETUP or AIRSAT
COMMON SHARED Pref, Qm, DK, BN, BO, BNA, BC, ANA, AC, AO, FSAT, FSATD, CSAT, CSATD, ESAT, ESATD, INC, I, SAO, UPA, g,
H, DOWN, UP, GL, LAI, SAI, L05, PROD1, PROD2, KLNT, KLN1, KLCT, KLN2, KLN3, KLC1, KLC2, KLC3

'These global variables are determined within the PROCEDURE UTUBE and are
'required by PROCEDURE OFFGAS
COMMON SHARED   P, XO, XN, MOLO2, MOLN2, TOTV
'Set array dimensions

DIM   ET(550), FT(550), CT(550), ES(550), FS(550), CS(550), ER(550), FR(550),
      CR(550), EU(550), FU(550), CU(550)

'These are global array variables
COMMON SHARED   ET(), FT(), CT(), ES(), FS(), CS(), ER(), FR(), CR(), EU(),
                  FU(), CU()

'Variable Identification :
'Ex = oxygen concentration (mg/L)
'Fx = nitrogen + argon concentration (mg/L)
'Cx = carbon dioxide concentration (mg/L)
'Where x = T, S, R, or U
'\(x_T\) = Bulk TANK conc. also CLARIFIER influent
'\(x_S\) = Bulk SUMP conc. also RBC influent
'\(x_R\) = RBC effluent also U-TUBE influent
'\(x_U\) = U-TUBE effluent also TANK influent
'Where x = E or F

DEFINT U      'Defines U as integer to serve as counter for array elements
COMMON SHARED U   'Defines U as global variable

CALL DISPLAY
CALL INPARAMETERS
CALL BACKGROUND

10      CLS
cho = 0
PRINT : PRINT : PRINT : PRINT
PRINT " MAIN MENU"
PRINT : PRINT : PRINT
PRINT ; TAB(10); "1. Change input parameters": PRINT
PRINT ; TAB(10); "2. Run simulation": PRINT
PRINT ; TAB(10); "3. Exit program": PRINT PRINT
INPUT " ENTER CHOICE "; cho
IF cho = 1 THEN
  CALL INPARAMETERS
  CALL BACKGROUND
ELSEIF cho = 2 THEN
GOTO 100
ELSEIF cho = 3 THEN
GOTO 212
ELSE
GOTO 10
END IF
GOTO 10

100 CALL SETUP
CLS
U = 0
LOCATE 5, 30: PRINT "SIMULATION IN PROGRESS"
LOCATE 12, 35: PRINT "Loop Count"
LOCATE 17, 25: PRINT "Maximum allowable Count = 500"
DO
U = U + 1
LOCATE 13, 40: PRINT ; U
CALL UTUBE
EU(U) = E
FU(U) = F
CU(U) = CD
CALL TANK
ET(U) = E
FT(U) = F
CT(U) = CD
CALL SUMP
ES(U) = E
FS(U) = F
CS(U) = CD
CALL RBC
ER(U) = E
FR(U) = F
CR(U) = CD

'Mass balance Ratio on sources and sinks of gases for equilibrium check
ETEST = ABS((Qm * (EU(U) - ES(U)) + KLOT*(ESATD - ET(U))*VOLT) / ((DEMt + DEMs) * 16.67))
FTEST = ABS(Qm*(FR(U) - FS(U)) + KLNT*(FSATD - FT(U))*VOLT) / (FR(U) - FU(U))
CTEST = ABS((Qm * (CR(U) - CS(U)) + KLCT*(CSATD - CT(U))*VOLT) / ((PRODt + PRODs) * 16.67))
IF ETEST > .98 AND FTEST > .98 AND CTEST > .98 AND U > 10 THEN
  TEST$ = "TRUE"
ELSEIF U > 500 THEN
  TEST$ = "TRUE"
ELSE
  TEST$ = "FALSE"
END IF

LOOP WHILE TEST$ = "FALSE"

CALL REPORT
CLS
GOTO 10
212 END

'*********** END MAIN PROGRAM***********
PROCEDURE AIRSAT
CALCULATE ATMOSPHERIC SATURATION CONCENTRATIONS

SUB AIRSAT

'Temperature in Kelvin
DK = C + 273.16

'Water vapor pressure, Z, for given conditions
V1 = 24.4543 - 67.4509 * (100 / DK) - 4.8489 * LOG(DK / 100)
Z = (2.71828 ^ (V1)) * 760

'Bunsen coefficients for Oxygen, Nitrogen, Argon, and Carbon dioxide
BO = 2.7183 ^ (85.8079 * (100 / DK) + 23.8439 * LOG(DK / 100) - 58.3877 + S * (-.034892 + .015568 * (DK / 100) - .0019387# * (DK / 100) ^ 2))
BN = 2.7183 ^ (85.7661 * (100 / DK) + 24.3696 * LOG(DK / 100) - 59.6274 + S * (-.05158 + .026329 * (DK / 100) - .0037252# * (DK / 100) ^ 2))
BA = 2.7183 ^ (82.0262 * (100 / DK) + 22.5929 * LOG(DK / 100) - 55.6578 + S * (-.036267 + .016241 * (DK / 100) - .0020114# * (DK / 100) ^ 2))
BC = 22.263 * 2.7183 ^ (90.5069 * (100 / DK) + 22.294 * LOG(DK / 100) - 58.0931 + S * (.027766 - .025888 * (DK / 100) + .0050578# * (DK / 100) ^ 2))

'Weighted nitrogen+argon values for Bunsen coefficient, molecular wt / molecular volume ratio, and constant A
BNA = (BN * .78084 + BA * .00934) / .7902
KNA = (BN * .78084 + BA * .00934) / (1.25043 * .78084 * BN + 1.798419 * .00934 * BA)

'Calculate A constants for oxygen, nitrogen+argon, and carbon dioxide
AO = .5318
AC = .3845
ANA = .76 * KNA

'Saturation conc. for Oxygen, Nitrogen+Argon, and Carbon dioxide in air
ESAT = 1000 * 1.42903 * BO * .20946 * (D - Z) / 760
FSAT = 1000 * BNA * .79018 * (D - Z) / (760 * KNA)
CSAT = 1000 * 1.97681 * BC * .00032 * (D - Z) / 760

END SUB  'AIRSAT
'PROCEDURE ASCFLE
WRITE DO CONCENTRATION DATA TO ASCII FILE'

DEFINT U
SUB ASCFLE

'Maintain values for time and filename variable
STATIC Firstpass, time, NAME$

'Provides a default filename and path specification on initial
IF Firstpass = 0 THEN NAME$ = "B:NONE"

CLS
PRINT: PRINT: PRINT
PRINT "You may specify the drive, path and name of the data file you wish"
PRINT "simulation data to be written."
PRINT
PRINT "If the data file already exists the data will be appended else the file will
PRINT "be created and written to the specified drive."
PRINT
PRINT " IF the simulation is a response to some change of an input parameter for the"
PRINT "previous simulation then Press Enter to keep Displayed Path and Name and file"
PRINT "will be appended"
PRINT
PRINT " NOTE: File name must be eight characters or less and DO NOT specify a file"
PRINT "extension as the extension '.DAT' will be added automatically."
PRINT: PRINT

PRINT "Path and File Name are "; NAME$
INPUT " ENTER PATH AND FILE"; SWTCH$
IF SWTCH$ <> "" THEN NAME$ = SWTCH$

Set counter for DO...WHILE Loop which writes array elements to ASCII file and reset
'U = 0 to access first element in each array

TOTAL = U

U = 0

OPEN NAME$ + ".DAT" FOR APPEND AS #1
WRITE #1, time, ET(U), EU(U), ES(U), ER(U)
time = time + INC / 60
DO
  U = U + 1
  WRITE #1, time, ET(U), EU(U), ES(U), ER(U)
  time = time + 4 * INC / 60
  LOOP WHILE U < TOTAL
CLOSE #1
Firstpass = 1

END SUB 'ASCFLE
PROCEDURE BACKGROUND
'CALCULATE BACKGROUND PARAMETERS FOR SIMULATION

DEFINT U
SUB BACKGROUND

'Apply BETA correction factors to saturation values
ESATD = BCF * ESAT
FSATD = BCF * FSAT
CSATD = BCF * CSAT

'Calculate nitrogen and carbon dioxide transfer coefficients
'RBC
KLN1 = kN * KLO1 : KLC1 = kC * KLO1
KLN2 = kN * KLO2 : KLC2 = kC * KLO2
KLN3 = kN * KLO3 : KLC3 = kC * KLO3
'Tank
KLNT = kN * KLOT : KLCT = kC * KLOT
'Coefficients for u-tube are determined in PROC U-TUBE

'Carbon dioxide producton due to respiration in tank and sump
PRODt = 1.36 * DEMt
PRODs = 1.36 * DEMs

'Pressure at U-tube Inlet (mm Hg)
Pref = D + (J * 3.048) * 22.89

'Flow calculations for U-tube
Q = Qm / 3.758
LAI = ((LID / 2.54) / 24) ^ 2 * 3.1417
SAI = ((SID / 2.54) / 24) ^ 2 * 3.1417
SAO = ((SOD / 2.54) / 24) ^ 2 * 3.1417
UPA = LAI - SAO

g = Q / 60 / 7.48 / SAI  'Downflow velocity in U-tube (ft/sec)
H = Q / 60 / 7.48 / UPA  'Uplow velocity in U-tube (ft/sec)

'Downflow Leg Detention Time (sec)
IF Q > 0 THEN
   DOWN = A * 3.048 / g
   UP = A * 3.048 / H
ELSE
   DOWN = 30  'Establishes incremental time value
   UP = 30  'under conditions of 0 flow rate
END IF
'Incremental time value for component time step is equal to U-tube Detention Time (sec)
INC = DOWN + UP
'Finite difference elemental time (sec)
I = 0.25
'Compute Liquid Frictional Headloss per foot length in downflow leg
HLD = (Q / (.281 * HWC * (SID / 2.54) ^ 2.63)) ^ 1.85
'Compute hydraulic radius of upflow crossection
hr = UPA / (2 * 3.1417 * ((LID / 2.54) / 24) + 2 * 3.1417 * ((SOD / 2.54) / 24))
'Liquid frictional headloss per foot in upflow leg
HLU = (H / (1.318 * HWC * hr ^ .63)) ^ 1.851

END SUB     'BACKGROUND
PROCEDURE DISPLAY
'PRINTS MODEL DESCRIPTION TO SCREEN

SUB DISPLAY
CLS
PRINT : PRINT
PRINT "THIS PROGRAM MODELS GAS TRANSFER WITHIN A"
PRINT "RECIRCULATING AQUACULTURE"
PRINT "SYSTEM WHEN GIVEN DESIGN AND OPERATING VALUES FOR"
PRINT "THE FOLLOWING:"
PRINT : PRINT
PRINT ; TAB(10); "TANK"
PRINT ; TAB(10); "U-TUBE AERATOR"
PRINT ; TAB(10); "3 STAGE RBC"
PRINT ; TAB(10); "MULTI-TUBE CLARIFIER"
PRINT ; TAB(10); "STOCKING DENSITY AND DAILY FEED RATION"
PRINT ; TAB(10); "RECIRCULATION RATE"
PRINT : PRINT
PRINT "Written by Luther G. Wood, OCTOBER 1991"
PRINT "Aquaculture Research Facility"
PRINT "Dept. of Agricultural Engineering and Dept of Fisheries and Wildlife"
PRINT "Virgina Polytechnic Insitute and State University"
PRINT : PRINT
PRINT "PRESS ANY KEY TO CONTINUE"
X$ = INPUT$(1)

END SUB 'DISPLAY
'PROCEDURE INPARAMETERS
'RECEIVES USER INPUT DATA FROM KEYBOARD

DEFINT U
SUB INPARAMETERS
STATIC Firstpass, SCAL, KLOTIN, KLO1IN, KLO2IN, KLO3IN

1950 CLS
    cho = 0
    IF Firstpass = 0 THEN
        cho = 4
        GOTO 2080
    END IF

PRINT "CHOOSE A COMPONENT TO ENTER DESIGN INFORMATION"
PRINT : PRINT : PRINT
PRINT ; TAB(10); "1. TANK AND SUMP": PRINT
PRINT ; TAB(10); "2. RBC": PRINT
PRINT ; TAB(10); "3. U-TUBE": PRINT
PRINT ; TAB(10); "4. ALL": PRINT
INPUT "ENTER CHOICE": cho

'IF statements control menu selections
    IF cho = 1 OR cho = 4 THEN
        GOTO 2080
    ELSEIF cho = 2 THEN
        GOTO 2350
    ELSEIF cho = 3 THEN
        GOTO 2340
    ELSE
        GOTO 1950
    END IF

2080 REM ***** INPUT CODE FOR TANK AND SUMP *****

CLS
IF Firstpass = 0 THEN
    EFFp = .35
    D = 760
    C = 25
    VOLT = 8900
    Qm = 280
AIK = 100
SCAL = 1

END IF

LOCATE 4, 20:PRINT "TANK AND SUMP INPUT"
LOCATE 9, 10:PRINT "TANK VOLUME IS "; VOLT; "Liters"
LOCATE 10, 10:PRINT "Va. Tech Design Value = 8900 L"
CALL SWITCH(VOLT)
LOCATE 9, 10:PRINT "LIQUID VOLUME FLOW RATE IS "; Qm; " (LPM)"
CALL SWITCH(Qm)
Q = Qm * 3.785
LOCATE 9, 10:PRINT "COMBINED EFFICIENCY OF PUMP IS"; EFFP
CALL SWITCH(EFFP)
LOCATE 9, 10:PRINT "WATER TEMPERATURE IS "; TEMP; " DEGREES C"
CALL SWITCH(TEMP)
LOCATE 9, 10:PRINT "ATMOSPHERIC PRESSURE IS "; PMA; " mm Hg"
CALL SWITCH(PMA)

'Determine atmospheric saturation concentrations for temperature and pressure
CALL AIRS
IF Firstpass = 0 THEN
    ET(0) = ESAT
    FT(0) = FSAT
    CT(0) = CSAT
END IF

LOCATE 8, 1:PRINT "From previous simulation, the Tank equilibrium D.O.
    Conc (mg/L) is "; USING "###.###"; ET(U)
LOCATE 10, 10:PRINT "Saturation Conc (mg/L) = "; USING "###.###"; ESAT; ;
CALL SWITCH(ET(U))
LOCATE 8, 1:PRINT "From previous simulation, the Tank equilibrium
    (D.N+Ar.) Conc (mg/L) is "; USING "###.###"; FT(U)
LOCATE 10, 10:PRINT "Nitrogen + Argon Saturation Conc (mg/L) = ";
    USING "###.###"; FSAT
CALL SWITCH(FT(U))
LOCATE 8, 1:PRINT "From previous simulation, the Tank equilibrium (D.C.)
    Conc (mg/L) is "; USING "###.###"; CT(U)
LOCATE 10, 10:PRINT "Carbon Dioxide Saturation Conc (mg/L) = "; USING
    "###.###"; CSAT
CALL SWITCH(CT(U))
LOCATE 9, 10:PRINT "Total Alkalinity as CaCO3 is "; AIK; "mg/L"
CALL SWITCH(AIK)

PRINT "You have the option to enter an oxygen consumption rate (mg per min)"

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PRINT "for the production tank or you can choose to have the program estimate
PRINT "demands for you based upon daily feed ration."
PRINT
CHS = 0

DO WHILE CHS < 1 OR CHS > 2
  PRINT " 1) Enter value for Oxygen Demand": PRINT
  PRINT " 2) Have program estimate Oxygen Demand"
  PRINT
  INPUT " Enter the number of your choice": CHS
  LOOP

IF CHS <> 2 THEN
  PRINT
  LOCATE 12, 10: PRINT " Enter oxygen Demand in Tank (g/hr) ";
  DEMt
  CALL SWITCH(DEMt)
  FEED = 0
ELSE
  CLS
  PRINT
  PRINT "An average oxygen demand rate is determined from average
daily feeding ration"
  LOCATE 9, 10: PRINT "AVERAGE FEEDING RATION IS "; FEED;
  kg PER DAY 
  CALL SWITCH(FEED)

PRINT " A scaling factor ( 0.1 to 5 ) can be applied to the average
demand to simulate peak or low demands. Respiration rates may
increase to a peak value of 1.4 to 2 times the average after heavy
feeding activity and decrease by a factor less than 1 during
extended periods without feed or during exposure to low D.O.
concentration. The recommended post-feeding peaking factor is
approximately 1.44."
PRINT " ENTER VALUE OF 1 FOR NO SCALING."

DO
  LOCATE 9, 10: PRINT "DEMAND SCALING FACTOR IS "; SCAL
  CALL SWITCH(SCAL)
  IF SCAL < .1 OR SCAL > 10 THEN
    LOCATE 8, 10: PRINT "ENTER A VALUE OF .1 TO 10"
  END IF
LOOP WHILE SCAL < .1 OR SCAL > 10

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'Estimate unscaled average oxygen demand (g/hr)
  DEMx = .3 * FEED * 1000! / (24)

'Determine scaled oxygen demand (g/hr)
  DEMt = SCAL * DEMx
END IF

LOCATE 8, 10: PRINT " The transfer coefficient at 20 C for intank aeration is ";
  KLOTx " min^-1"
LOCATE 9, 10: PRINT " Va Tech Design Value = 0.03 min^-1"
LOCATE 11, 10: PRINT " Enter a value of 0 for no intank aeration"
  CALL SWITCH(KLOT)
  KLOT = KLOT * 1.024 ^ (C - 20)

IF Firstpass = 0 THEN VOLS = 2500

LOCATE 9, 10: PRINT "Volume of sump or solids separation unit is "; VOLS;
LOCATE 10, 10: PRINT " Va Tech Design Value = 2500 L"
  CALL SWITCH(VOLS)

'Estimate clarifier oxygen demand based on daily feed ration

  IF FEED <> 0 THEN DEMs = 2.52 * FEED + 1.1

LOCATE 8, 10: PRINT "Estimated oxygen demand in solids-setting unit 
LOCATE 9, 10: PRINT " based on entered daily feed ration is "; DEMs; "g/hr"
  CALL SWITCH(DEMs)

CLS

IF cho = 1 THEN GOTO 2670  'If statements allow menu selections to
  'be executed properly
IF cho = 2 THEN GOTO 2670
IF cho = 3 THEN GOTO 2670

2340 REM ***** INPUT CODE FOR U-TUBE ********

IF Firstpass = 0 THEN
  ACF = 1
  BCF = 1
  HWC = 150
  EFFc = .35
  A = 40 / 3.048
  LID = 6 * 2.54
  SID = 3 * 2.54
  SOD = 3.5 * 2.54
  J = 1.5 / 3.048

END IF
IF cho = 1 THEN GOTO 2670  
IF cho = 2 THEN GOTO 2670  
IF cho = 3 THEN GOTO 2670

2350 REM ***** INPUT CODE FOR RBC **********

CLS
LOCATE 4, 32: PRINT " RBC INPUT"
   KLO1DES = (.00201 * Qm - .0013)
   KLO2DES = .167
   KLO3DES = KLO2DES
'Set Firstpass values equal to design values of RBC Design at Va. Tech
'Aquaculture' Research Facility
IF Firstpass = 0 THEN
   KLO1IN = KLO1DES
   KLO2IN = KLO2DES
   KLO3IN = KLO2DES
   VOLR = 650
   KN = .94
   KC = 1
END IF

LOCATE 7, 10: PRINT "VOLUME OF SINGLE RBC STAGE IS "; VOLR; "L"
LOCATE 9, 10: PRINT "Va Tech. Design Value = 650 L"
   CALL SWITCH(VOLR)
LOCATE 7, 10: PRINT "OXYGEN TRANSFER COEFFICIENT AT 20 C FOR STAGE 1 IS "; KLO1IN
LOCATE 9, 10: PRINT "Va Tech. Design Value for Entered Flow Rate = "; KLO1DES
   CALL SWITCH(KLO1IN)
PRINT "Enter a value of 0 for subsequent RBC transfer coefficients if RBC is single stage"
PRINT
LOCATE 7, 10: PRINT "OXYGEN TRANSFER COEFFICIENT AT 20 C FOR STAGE 2 IS "; KLO2IN; " MIN^-1"
LOCATE 9, 10: PRINT "Va Tech. Design Value = "; KLO2DES
   CALL SWITCH(KLO2IN)
LOCATE 7, 10: PRINT "OXYGEN TRANSFER COEFFICIENT AT 20 C FOR STAGE 3 IS "; KLO3IN; " MIN^-1"
LOCATE 9, 10: PRINT "Va Tech. Design Value = "; KLO3DES
   CALL SWITCH(KLO3IN)
LOCATE 9, 10: PRINT "RELATIVE K2 FOR NITROGEN (0.8-0.94) IS "; KN
   CALL SWITCH(kN)
LOCATE 9, 10: PRINT "RELATIVE K2 FOR NITROGEN IS "; kC
CALL SWITCH(kC)
KLO1 = KLO1IN * 1.024 ^ (C - 20)
KLO2 = KLO2IN * 1.024 ^ (C - 20)
KLO3 = KLO3IN * 1.024 ^ (C - 20)

IF cho = 1 THEN GOTO 2670 If statements control menu selections
IF cho = 2 THEN GOTO 2670
IF cho = 3 THEN GOTO 2670

2670 Firstpass = 1

END SUB 'INPARAMETERS
DEFINT U
SUB OFFGAS

Total volumetric rate of off-gasses exiting u-tube (CFM)
OOPS = TOTV * Q / 7.48

'Headloss due to two-phase flow (ft)
HLGL = (287 * GL - .0303) / 12

'Headloss through off-gas collector and two-phase flow (ft)
HLGLP = HLGL * 22.89 + (OGCHL * 3.048) * 22.89

'Standard Liters per minute of oxygen and nitrogen+argon exiting U-tube
SLPMO2 = MOLO2 * .0821 * DK * Qm
SLPMN2 = MOLN2 * .0821 * DK * Qm
VEX = OOPS * P / (P - HLGLP)

'Pressure at compressor inlet (kPa)
POGCE = (P - HLGLP) / 25.4 / .00029613# / 1000

'Convert pressure at U-tube inlet from mm Hg to kPa
PREFF = Pref / 25.4 / .00029613# / 1000

'Mass rate of off-gas recycled (kg/sec)
MASQ = (MOLO2 * 32 + MOLN2 * 28) * Qm / 60000! * (OGR / 100)
GCR = .2598 * XO + .2969 * XN

'Energy requirement to compress off-gasses (kW)
POGR = MASQ * GCR * DK / (.283 * EFFc) * ((PREFF / POGCE) ^ .283 - 1)
ERROG = POGR

'Energy requirement to pump flow
PUMP = ((Pref - P + HLGLP) / 22.89 * (Q / 60) * 8.34 / EFFp) / 550 / 1.341

'Total Energy requirement to pump Flow and recycle off-gasses
TER = PUMP + ERROG

'Mass Rate of oxygen injection into u-tube
MRO2 = (B * Qm * 32 * 60) / (.08206 * 264.2 * 1000)

'kgs of oxygen and Nitrogen+Argon Transferred per hour
kgO2 = (EU(U) - ER(U)) * Qm * 60 / 1000000
kgN2 = (FU(U) - FR(U)) * Qm * 60 / 1000000

'Aeration efficiency (kgs / kWhr)
TRANCEO = kgO2 / TER
TRANEN = kgN2 / TER

'Cost of injected oxygen per hr
O000 = B * (Qm * 60) * .0013253# * (CSTKG)
LPRINT "Volume of Off-Gas (N2+O2) Exiting U-tube = "; VEX * 28.32; "LPM at", P - HLGLP; "mm Hg"
LPRINT "Off-Gas Composition = "; XO * 100; "% OXYGEN"; XN * 100; "% NITROGEN"
LPRINT "Volume of Gas Recycled = "; ((SLPMO2 + SLPMN2) * (OGR / 100)); "SCFM" LPRINT "Effective Gas-Liquid Ratio = "; GL * 100; "%"
LPRINT LPRINT "********U-TUBE ENERGY REQUIREMENTS**********"
LPRINT "Velocity in DOWN LEG (m/S) = "; g / 3.048 LPRINT "Velocity in UP LEG (m/S) = "; H / 3.048
LPRINT "Head Loss Due To Pipe Friction = "; ((Pref - P) / 22.89 + (OGCHL * 3.048) / 3.048); "m"
LPRINT "HEAD LOSS Due To Two-Phase Flow = "; HLGL / 3.048; "m"
LPRINT "Power Required to Recycle off-Gas = "; ERROG; "kW"
LPRINT "Power Requirement (TOTAL) = "; TER; "kW"
LPRINT LPRINT "******TRANSFER EFFICIENCIES******"
LPRINT "U-tube Oxygen injection Rate = "; MRO2*1000; "g/hr"
LPRINT "Oxygen Transfer Rate"; kgO2 * 1000; "g/hr"
LPRINT "Oxygen Transfer Efficiency = "; kgO2 * 100 / MRO2; "%"
LPRINT "Oxygen Aeration Efficiency = "; TRANEO; "kg O2/kW-hr"
LPRINT LPRINT "Nitrogen Transfer Rate "; kgN2 * 1000; "g/hr"
LPRINT "Nitrogen Aeration Efficiency = "; TRANEN; "kg N2/kW-hr"
LPRINT LPRINT "******TRANSFER ECONOMY******"
LPRINT "Oxygen Cost per kg O2 Transfer = "; OOOO / kgO2
LPRINT "Energy Cost per kg O2 Transfer = "; TER * CKWH / kgO2
LPRINT "Total Cost per kg O2 Transfer = "; OOOO / kgO2 + TER * CKWH / kgO2
LPRINT LPRINT "Energy Cost per kg N2 Transfer = "; TER * CKWH / kgN2

END SUB 'OFFGAS
'PROCEDURE PH
'CALCULATE pH IN TANK

DEFINT U

'Passes carbon dioxide conc. to PROC PH and returns pH value to PROC TANK

SUB PH (CxF, PHYD)

'NOTE: In determining final pH, the total alkalinity is assumed to exist primarily as 'bicarbonate. If non-carbonate forms are not considered then the assumption is 'valid under most conditions encountered in fish culture. The model does not 'account for ammonia production or alkalinity consumption by nitrification.

MOLHC03 = AIK / 50000  'Moles alkalinity
H2CO3 = CxF / 44000   'Mole carbon dioxide
HYD = (10 ^ -6.3) * (H2CO3) / MOLHC03 'Moles hydrogen ion
PHYD = -1 * LOG(HYD) / 2.3026  'pH

END SUB  'PH
PROCEDURE PRESSURE
CALCULATE DISSOLVED GAS PRESSURES

DEFINT U

'Pass dissolved gas concentrations to PROC PRESSURE and return total and partial pressure back to PROCS TANKSUM, U-TUBESUM, SUMPSUM and RBCSUM

SUB PRESSURE (ExF, FxF, CxF, TGP, O2T, N2T, DPO2, DPN2, TODP)

'Calculate partial pressures
O2T = ExF / BO * AO
N2T = FxF / BNA * ANA
CO2T = CxF / BC * AC

'Excess partial pressures
DPO2 = O2T - ESAT / BO * AO
DPN2 = N2T - FSAT / BNA * ANA
DPCO2 = CO2T - CSAT / BC * AC

'Total excess pressure
TODP = DPO2 + DPN2 + DPCO2

'Total gas pressure as percent of barometric pressure
TGP = (O2T + N2T + CO2T + Z) / D * 100

END SUB 'PRESSURE
'PROCEDURE RBC
'MODELS RBC GAS TRANSFER

DEFINT U
SUB RBC

STATIC ER1, ER2, ER3, FR1, FR2, FR3, CR1, CR2, CR3

'Initialize stage gas concentrations at start of any simulation
IF U = 1 THEN
   ER1 = ER(0)
   FR1 = FR(0)
   CR1 = CR(0)
   ER2 = ER(0)
   FR2 = FR(0)
   CR2 = CR(0)
   ER3 = ER(0)
   FR3 = FR(0)
   CR3 = CR(0)
END IF

   t = 0

'Begins simulation loop for one component time step
DO WHILE t <= INC
   IF KLO1 = 0 THEN
      ER1 = E
      FR1 = F
      CR1 = CD
   ELSE
      DELDO1 = (Qm * E + (ACF * KLO1 * (ESATD - ER1) * VOLR) - Qm * ER1) * (I/60) / VOLR
      DELDN1 = (Qm * F + (ACF * KLN1 * (FSATD - FR1) * VOLR) - Qm * FR1) * (I/60) / VOLR
      DELDC1 = (Qm * CD + (ACF * KLC1 * (CSATD - CR1) * VOLR) - Qm * CR1) * (I/60) / VOLR
      ER1 = ER1 + DELDO1
      FR1 = FR1 + DELDN1
      CR1 = CR1 + DELDC1
   END IF

   IF KLO2 = 0 THEN
      ER2 = ER1
   END IF

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FR2 = FR1
CR2 = CR1

ELSE

DELD02 = (Qm * ER1 + (ACF * KLO2 * (ESATD - ER2) * VOLR) - Qm * ER2) * (I / 60) / VOLR
DELDN2 = (Qm * FR1 + (ACF * KLN2 * (FSATD - FR2) * VOLR) - Qm * FR2) * (I / 60) / VOLR
DELDC2 = (Qm * CR1 + (ACF * KLC2 * (CSATD - CR2) * VOLR) - Qm * CR2) * (I / 60) / VOLR

ER2 = ER2 + DELDO2
FR2 = FR2 + DELDN2
CR2 = CR2 + DELDC2

END IF

IF KLO3 = 0 THEN

ER3 = ER2
FR3 = FR2
CR3 = CR2

ELSE

DELD03 = (Qm * ER2 + (ACF * KLO3 * (ESATD - ER3) * VOLR) - Qm * ER3) * (I / 60) / VOLR
DELDN3 = (Qm * FR2 + (ACF * KLN3 * (FSATD - FR3) * VOLR) - Qm * FR3) * (I / 60) / VOLR
DELDC3 = (Qm * CR2 + (ACF * KLC3 * (CSATD - CR3) * VOLR) - Qm * CR3) * (I / 60) / VOLR

ER3 = ER3 + DELDO3
FR3 = FR3 + DELDN3
CR3 = CR3 + DELDC3

END IF

t = t + 1

LOOP

'Set effluent concentrations equal to third stage concentration
E = ER3
F = FR3
CD = CR3

END SUB 'RBC
PROCEDURE RBCSUM
'SUMMARIZE AND PRINT OUTPUT FOR RBC

DEFINT U
SUB RBCSUM

LPRINT"+++++++++++++++++++++++++++++++++++++++++++++++++++
LPRINT" RBC EFFLUENT CONDITIONS"
LPRINT"+++++++++++++++++++++++++++++++++++++++++++++++++++"
'Set gas concentrations to local variables to pass to PROC PRESSURE
ERF = ER(U)
FRF = FR(U)
CRF = CR(U)
CALL PRESSURE(ERF, FRF, CSF, TGP, O2T, N2T, DPO2, DPN2, TODP)
LPRINT "DISSOLVED OXYGEN LEVELS"
LPRINT " Initial - '; ER(0); " mg/L"
LPRINT " Equilibrium - '; ERF; " mg/L" LPRINT "Oxygen Tension = '; O2T; "mm Hg"
LPRINT " Excess Oxygen Tension = '; DPO2; "mm Hg"
LPRINT "Oxygen as Percent of Saturation = '; ERF * 100 / ESAT; "%"
LPRINT "Mass oxygen transferred to water = '; (ER(U) - ES(U)) * Qm * 60 / 1000; "g/hr"
LPRINT
LPRINT "DISSOLVED NITROGEN+ARGON LEVELS"
LPRINT " Initial = '; FR(0); " mg/L"
LPRINT " Equilibrium = '; FRF; " mg/L"
LPRINT "Nitrogen Tension = '; N2T; "mm Hg"
LPRINT "Excess Nitrogen Tension = '; DPN2; "mm Hg"
LPRINT "Nitrogen as Percent of Saturation = '; FRF * 100 / FSAT; "%"
LPRINT "Mass of Nitrogen transferred to water = '; (FR(U)-FS(U)) * Qm*60/1000; "g/hr"
LPRINT
LPRINT "DISSOLVED CARBON DIOXIDE LEVELS"
LPRINT " Initial = '; CR(0); " mg/L"
LPRINT " Equilibrium = '; CRF; " mg/L"
LPRINT "Carbon Dioxide as Percent of Saturation = '; FRF * 100 / FSAT; "%"
LPRINT "Mass of Carbon Dioxide transferred to water = '; (CR(U) - CS(U)) * Qm * 60 / 1000; "g/hr"
LPRINT
LPRINT "*****TOTAL GAS PRESSURE *****"
LPRINT "Total Excess Gas Pressure (DELTA P) = '; TODP; "mm Hg"
LPRINT "Total Gas Pressure (Percent Saturation) = '; TGP; "%"
IF ERF > .1 THEN LPRINT "Nitrogen-Oxygen Ratio = '; N2T / O2T LPRINT :
LPRINT

END SUB    'RBCSUM
PROCEDURE REPORT

OUTPUT FORMAT MENU AND INITIALIZATION OF OUTPUT GENERATION

DEFINT U
SUB REPORT

CLS
BEEP: BEEP: BEEP

'Print Gas concentration to screen for user inspection
LOCATE 6, 10: PRINT "EQUILIBRIUM CONCENTRATIONS IN TANK (mg/L)"
LOCATE 8, 15: PRINT " DO "; USING "###.#"; ET(U)
LOCATE 9, 15: PRINT " DN "; USING "###.#"; FT(U)
LOCATE 10, 15: PRINT " DC "; USING "###.#"; CT(U)

Provide REPORT MENU
LOCATE 12, 10: PRINT "FORMAT OUTPUT BEFORE RETURNING TO MAIN MENU"
LOCATE 14, 20: PRINT "1) Short Output Summary to Printer"
LOCATE 15, 20: PRINT "2) Detailed Output Summary to Printer"
LOCATE 16, 20: PRINT "3) No Output"
DO
LOCATE 18, 20: INPUT "ENTER CHOICE"; CHO1
LOOP UNTIL CHO1 <> 1 OR CHO1 <> 2 OR CHO1 <> 3
'Call PROCS to generate printed output in report format of users choice
IF CHO1 = 1 THEN
   CALL SHORTSUM
ELSEIF CHO1 = 2 THEN
   CALL U-TUBESUM
   CALL TANKSUM
   CALL RBCSUM
   CALL SUMPSUM
END IF

'Provide option to write array, ET(U), to ASCII file
IF CHO1 = 1 OR CHO1 = 2 THEN
   DO
      CLS
      LOCATE 9, 10: PRINT "Do you wish to write transgression of DO concentration from initial"
      LOCATE 10, 10: INPUT "to equilibrium to a Comma-Delimited ASCII File (Y or N)?"; X$
LOOP UNTIL X$ = "y" OR X$ = "Y" OR X$ = "N" OR X$ = "n"
END IF
IF X$ = "y" OR X$ = "Y" THEN CALL ASCFLE: X$ = ""

ENDSUB 'REPORT
'PROCEDURE SETUP

'INITIALIZES GAS CONCENTRATION AT BEGINNING OF SIMULATION

DEFINT U

SUB SETUP

STATIC Firstpass

IF Firstpass = 0 THEN
  ER(0) = ESATD
  FR(0) = FSATD
  CR(0) = CSATD
  EU(0) = ESATD
  FU(0) = FSATD
  CU(0) = CSATD
  FS(0) = FT(0)
  ES(0) = ET(0)
  CS(0) = CT(0)
  Firstpass = 1
ELSE
  ER(0) = ER(U)
  FR(0) = FR(U)
  CR(0) = CR(U)
  EU(0) = EU(U)
  FU(0) = FU(U)
  CU(0) = CU(U)
  ET(0) = ET(U)
  FT(0) = FT(U)
  CT(0) = CT(U)
  ES(0) = ES(U)
  FS(0) = FS(U)
  CS(0) = CS(U)
END IF

'In first simulation of any program session, initial values of tank and clarifier effluent concentrations are defined in PROC INPARAMETERS and RBC and U-tube effluent set equal to saturation conc.

Else

'In subsequent simulations in any program session, initial values are set equal to equilibrium values of previous simulation

'Set values of effluent variables, E, F, and CD equal to RBC effluent to define influent for U-tube component which is first component called in flow circuit loop

E = ER(0)
F = FR(0)
CD = CR(0)

END SUB    'SETUP
PROCEDURE SHORTOFFGAS

'Provides brief report on u-tube gas transfer performance

DEFINT U
SUB SHORTOFFGAS

'Total volumetric rate of off-gasses exiting u-tube (CFM)

OOPS = TOTV * Q / 7.48

'Headloss due to two-phase flow (ft)

HLGL = (287 * GL - .0303) / 12

'Headloss through off-gas collector and two-phase flow (ft)

HLGLP = HLGL * 22.89 + (OGCHL * 3.048) * 22.89

VEX = OOPS * P / (P - HLGLP)

'Pressure at compressor inlet (kPa)

POGCE = (P - HLGLP) / 25.4 / .00029613# / 1000

'Convert pressure at U-tube inlet from mm Hg to kPa

PREFF = Pref / 25.4 / .00029613# / 1000

'Mass rate of off-gas recycled (kg/sec)

MASQ = (MOLO2 * 32 + MOLN2 * 28) * Qm / 60000! * (OCR / 100)

GCR = .2598 * XO + .2969 * XN

'Energy requirement to compress off-gasses (kW)

POGR = MASQ * GCR * DK / (.283 * EFFc) * ((PREFF / POGCE) ^ .283 - 1)

ERROG = POGR

'Energy requirement to pump flow (kW)

PUMP = ((Pref - P + HLGLP) / 22.89 * (Q / 60) * 8.34 / EFFp) / 550 / 1.431

'Total energy requirement to pump flow and recycle off-gasses

TER = PUMP + ERROG

'Mass Rate of oxygen injection into u-tube

MRO2 = (B*Qm*32*60) / (.08206*264.2*1000)

'Kgs of oxygen and nitrogen + Argon transferred per hour

kgO2 = (EU(U) - ER(U)) * Qm / 60! / 1000000

'Oxygen aeration efficiency

TRANEO = kgO2 / TER

'Cost of injected oxygen per hr

OOOO = B * (Qm * 60) * .0013253# * (CSTKG)

LPRINT "Effective Gas-Liquid Ratio (%) = "; USING "##.##"; GL * 100
LPRINT"U-tube Oxygen injection Rate = ";MRO2*1000;"g/hr"
LPRINT "Oxygen Transfer Rate"; kgO2 * 1000; "g/hr"
LPRINT "Oxygen Transfer Efficiency = "; kgO2 * 100 / MRO2; "%
LPRINT "Oxygen Aerated Efficiency = "; TRANEO; "kg O2/kW-hr"
LPRINT "Oxygen Cost per kg Transfer = "; USING "##.####"; OOOO / kgO2
LPRINT "Energy Cost per kg O2 Transfer = "; USING"###.####"; TER*CKWH/kgO2
LPRINT "Total Cost per kg O2 Transfer = "; USING "####.#####"; OOOO / kgO2 +
TER * CKWH / kgO2

END SUB  'SHORTOFFGAS
PROCEDURE SHORTSUM
'PROVIDES OUTPUT SUMMARY

DEFINT U
SUB SHORTSUM

ETF = ET(U) : FTF = FT(U) : CTF = CT(U)
ESF = ES(U) : FSF = FT(U) : CSF = CT(U)
EUF = EU(U) : FUF = FT(U) : CUF = CT(U)
ERF = ER(U) : FRF = FT(U) : CRF = CT(U)
CALL PRESSURE(ETF, FTF, CTF, TGPT, O2T, N2T, DPO2, DPN2, TODP)
CALL PRESSURE(ESF, FSF, CSF, TGPS, O2T, N2T, DPO2, DPN2, TODP)
CALL PRESSURE(EUF, FUF, CUF, TGPU, O2T, N2T, DPO2, DPN2, TODP)
CALL PRESSURE(ERF, FRF, CRF, TGPR, O2T, N2T, DPO2, DPN2, TODP)
LPRINT:"Count = ";U
LPRINT "****EQUILIBRIUM DISSOLVED GAS CONCENTRATIONS
(mg/L)****"
LPRINT
LPRINT: TAB(11); "TANK"; TAB(20); "CLARIFIER"; TAB(37); "RBC"; TAB(50);
"U-TUBE"
LPRINT" DO";USING; "###.#"; TAB(10); ET(U); TAB(22); ES(U); TAB(35); ER(U);
TAB(51); EU(U)
LPRINT" DN";USING; "###.#"; TAB(10); FT(U); TAB(22); FS(U); TAB(35); FR(U);
TAB(51); FU(U)
LPRINT" DC";USING; "###.#"; TAB(10); CT(U); TAB(22); CS(U); TAB(35); CR(U);
TAB(51); CU(U)
LPRINT"TGP(%)";USING; "###.#"; TAB(10); TGPT; TAB(22); TGPS; TAB(35); TGPR;
TAB(51); TGPU
LPRINT:LPRINT
LPRINT" Flow = ";Qm; " L/min"
CALL PH(CTF, PHYD)
LPRINT "pH = "; USING "#.##"; PHYD
IF FEED <> 0 THEN LPRINT "Daily Feed Ration"; FEED; "kg/day"
LPRINT "Estimated Oxygen Demand "; DEMt; "g/hr"
LPRINT "Estimated Carbon Dioxide Production Rate "; PRODt; "g/hr"
LPRINT
LPRINT" U-tube Performance "
LPRINT "EXIT DO (mg/L)"; USING "##.#"; EU(U)
CALL SHOROFFGAS

END SUB 'SHORTSUM
DEFINT U
SUB SUMP

STATIC ES, CS

'Initialize component concentrations

IF U = 1 THEN'
    ES = ES(0)
    CS = CS(0)
END IF

t = 0

'BEGIN clarifier simulation loop for one component time step
DO
    DELDO = (Qm * E - (DEMs * 16.67) - ES * Qm) * (I / 60) / VOLS
    DELDC = (Qm * CD + (PRODs * 16.67) - CS * Qm) * (I / 60) / VOLS
    ES = ES + DELDO
    CS = CS + DELDC
    IF ES < 0 THEN ES = 0
    IF CS < 0 THEN CS = 0
    t = t + I
LOOP WHILE t < INC

'Set effluent value equal to component values
    E = ES
    CD = CS

END SUB 'SUMP
`PROCEDURE SUMPSUM
'SUMMARIZE AND PRINT OUTPUT FOR MULT-TUBE CLARIFIER
`******************************************

DEFINT U
SUB SUMPSUM

LPRINT"+++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++",
LPRINT "    CLARIFIER EFFLUENT CONDITIONS"
LPRINT"+++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++",
LPRINT
     ESF = ES(U)
     FSF = FS(U)
     CSF = CS(U)
CALL PRESSURE(ESF, FSF, CSF, TGP, O2T, N2T, DPO2, DPN2, TODP)
LPRINT "DISSOLVED OXYGEN LEVELS"
LPRINT "    Initial = "; ES(0); " mg/L"
LPRINT "    Equilibrium = "; ESF;"mg/L"
LPRINT "    Oxygen Tension = "; O2T; " mm Hg"
LPRINT "    Excess Oxygen Tension = "; DPO2; " mm Hg"
LPRINT "    Oxygen as Percent of Saturation = "; ESF * 100 / ESAT; "%
LPRINT "    Estimated Demand is "; DEMs; "g/hr"
LPRINT
LPRINT "DISSOLVED NITROGEN LEVELS"
LPRINT "    Initial = "; FS(0); " mg/L"
LPRINT "    Equilibrium = "; FSF; " mg/L"
LPRINT "    Nitrogen Tension = "; N2T; " mm Hg"
LPRINT "    Excess Nitrogen Tension = "; DPN2; " mm Hg"
LPRINT "    Nitrogen as Percent of Saturation = "; FSF * 100 / FSAT; "%
LPRINT
LPRINT "DISSOLVED CARBON DIOXIDE LEVELS"
LPRINT "    Initial = "; CS(0); " mg/L"
LPRINT "    Equilibrium = "; CSF; " mg/L"
LPRINT "    Carbon dioxide production rate = "; PRODs; " g/hr" LPRINT
LPRINT "****TOTAL GAS PRESSURE *****"
LPRINT "    Total Excess Gas Pressure (DELTA P) = "; TODP; " mm Hg"
LPRINT "    Total Gas Pressure (Percent Saturation) = "; TGP; "%
IF ESF > .1 THEN LPRINT "    Nitrogen-Oxygen Ratio = "; N2T / O2T
LPRINT : LPRINT

END SUB    'SUMPSUM
PROCEDURE SWITCH
SWITCHES INPUTTED VALUE FROM DUMMY VARIABLE TO MODEL INPUT VARIABLE

SUB SWITCH (SWX)
IN$ = ""

'Forces user to input only numeric characters and allows user to retain displayed value by
'only pressing ENTER key
DO
  LOCATE 20, 10: PRINT "TO RETAIN DISPLAYED VALUE PRESS ENTER"
  LOCATE 21, 10: PRINT " OR"
  LOCATE 22, 10: INPUT "TYPE NEW VALUE AND PRESS ENTER"; IN$
  PRINT
  hx = LEN(IN$)
  tx = VAL(IN$)
  tx$ = STR$(tx)
  hy = LEN(tx$) - 1
  IF IN$ <> "" AND IN$ = "0" THEN
    TEST = 1
  ELSEIF IN$ = "" THEN
    TEST = 1
  ELSEIF hy = hx AND tx <> 0 THEN
    TEST = 1
  ELSEIF tx < 0 AND hy = hx - 1 THEN
    TEST = 1
  ELSE
    LOCATE 12, 10: PRINT "Enter only numeric characters and omit
    insignificant zeros"
    LOCATE 13, 10: PRINT " Example - RIGHT WRONG"
    LOCATE 14, 10: PRINT " 1.05  1.050"
    LOCATE 15, 10: PRINT " .0005  5e-4"
    LOCATE 16, 10: PRINT " -.05  - .05"
    LOCATE 17, 10: PRINT
  END IF
LOOP UNTIL TEST = 1

IF IN$ <> "" THEN SWX = VAL(IN$)
CLS

END SUB 'SWITCH
DEFINT U
SUB TANK

STATIC ET, FT, CT

Initialize dissolved gas concentrations
  IF U = 1 THEN'
    ET = ET(0)
    FT = FT(0)
    CT = CT(0)
  END IF

t = 0

'Execute tank simulation for one component time step
  DO
    DELDO = (Qm * E + (ACF * KLOT * (ESATD - ET) * VOLT) - (DEMt * 16.67) - Qm * ET) * (I / 60) / VOLT
    DELDN = (Qm * F + (ACF * KLNT * (FSATD - FT) * VOLT) - Qm * FT) * (I / 60) / VOLT
    DELDC = (Qm * CD + (ACF * KLCT * (CSATD - CT) * VOLT) + (PRODl * 16.67) - Qm * CT) * (I / 60) / VOLT
    IF ET + DELDO < 0 THEN ET = 0 ELSE ET = ET + DELDO
    IF FT + DELDN < 0 THEN FT = 0 ELSE FT = FT + DELDN
    IF CT + DELDC < 0 THEN CT = 0 ELSE CT = CT + DELDC
    t = t + 1
  LOOP WHILE t < INC

'Set effluent values equal to component values
  E = ET
  F = FT
  CD = CT

END SUB 'TANK
**PROCEDURE TANKSUM**

'SUMMARIZE AND PRINT OUTPUT FOR PRODUCTION TANK'

```
******************************************************************************
*DEFINT U
*SUB TANKSUM
*LPRINT"*****************************************************************************"
*LPRINT " Conditions in Tank "
*LPRINT"*****************************************************************************"
*LPRINT
ETF = ET(U)
FTF = FT(U)
CTF = CT(U)
CALL PRESSURE(ETF, FTF, CTF, TGP, O2T, N2T, DPO2, DPN2, TDP)
*LPRINT
IF FEED <> 0 THEN LPRINT "Daily Feed Ration = "; FEED; "kg/day"
*LPRINT
LPRINT "DISSOLVED OXYGEN LEVELS"
LPRINT " Initial = "; ET(0); " mg/L"
LPRINT " Equilibrium = "; ETF; "mg/L"
LPRINT "Oxygen Tension = "; O2T; "mm Hg"
LPRINT "Excess Oxygen Tension = "; DPO2; "mm Hg"
LPRINT "Oxygen as Percent of Saturation = "; ETF * 100 / ESAT; "%"
LPRINT "Estimated Oxygen Demand"; DEMt; "g/hr"
*LPRINT
LPRINT "DISSOLVED NITROGEN LEVELS"
LPRINT " Initial = "; FT(0); " mg/L"
LPRINT " Equilibrium = "; FTF; " mg/L"
LPRINT "Nitrogen Tension = "; N2T; "mm Hg"
LPRINT "Excess Nitrogen Tension = "; DPN2; "mm Hg"
LPRINT "Nitrogen as Percent of Saturation = "; FTF * 100 / FSAT; "%"
*LPRINT
LPRINT "DISSOLVED CARBON DIOXIDE LEVELS"
LPRINT " Initial = "; CT(0); " mg/L"
LPRINT " Equilibrium = "; CTF; " mg/L"
LPRINT "Carbon dioxide production rate = "; PROD; " g/hr"
CALL PH(CTF, PHYD)
LPRINT "pH = "; PHYD
LPRINT "*****TOTAL GAS PRESSURE *****"
LPRINT "Total Excess Gas Pressure (DELTA P) = "; TDP; "mm Hg"
LPRINT "Total Gas Pressure (Percent Saturation) = "; TGP; "%" IF ETF > .1 THEN
LPRINT "Nitrogen-Oxygen Ratio = "; N2T / O2T LPRINT : LPRINT
END SUB "TANKSUM
```
PROCEDURE UTUBE
MODELS U-TUBE GAS TRANSFER

SUB U-TUBE

TOTAL Firstpass, VOLREF, MO2, MN2, M

'Bypasses subroutine calculation if flow rate = 0 to avoid division by 0
IF Q = 0 THEN GOTO 950

'Initialize GW ratio, gas bubble volume, molar concentration of oxygen and nitrogen,
'primary feed ratio on first execution of any simulation or conditions of no off-gas recycle

IF Firstpass = 0 OR OGR = 0 THEN
XO = 1
XN = 0
GL = B
MN2 = 0
MO2 = GL * 1325.247
MOLREF = MO2 / 32000
VO2 = DK * 760 * GL / 294.2 / Pref
VOLREF = VO2
M = g
firstpass = 1
END IF

'Establish pressure to inlet reference pressure
P = Pref

'Set element plug volume concentration equal to inlet concentration
YO = E
YN = F

'Initial estimation of inlet kla, set kla references for kla and gas volume relationship,
'and compute kla for nitrogen
590 L = ACF * (.000386 + (.61344 * GL)) * 1.024 ^ (C - 20)
LL = L
n = k * L
NN = n
630 CEO = 1429.03 * BO * XO * ((P - Z) / 760) * BCF
CEN = 1250.43 * BN * XN * ((P - Z) / 760) * BCF
DELDO = L * (CEO - E) * I
DELDN = n * (CEN - F) * I
CO = E + DELDO
E = CO
CN = F + DELDN
F = CN
MO = MO2 - DELDO
MO2 = MO
MN = MN2 - DELDN
MN2 = MN
MOLO2 = MO2 / 32000
MOLN2 = MN2 / 28000
XO = MOLO2 / (MOLO2 + MOLN2)
XN = MOLN2 / (MOLO2 + MOLN2)
VO2 = MOLO2 * .0821 * DK / (P / 760) VN2 = MOLN2 * .0821 * DK / (P / 760)
TOTV = VO2 + VN2
VOLFRA = TOTV / VOLREF
L = LL * VOLFRA
n = NN * VOLFRA
t = t + I

'Is pass through u-tube completed?
   IF t > INC THEN GOTO 920

'Is pass through downleg completed?
   IF DOWN > t THEN GOTO 900


   GOTO 630

900   P = D + (J * 3.048) * 22.89 + g * t * 22.89 - g * t * HLD * 22.89

   GOTO 630

'Are off-gasses recycled? IF not then goto line label for END SUB
920   IF OGR = 0 THEN GOTO 950

   MOLO2 = MOLO2 * (OGR / 100) + B * 1325.247 / 32000
   MO2 = MOLO2 * 32000
   MOLN2 = MOLN2 * (OGR / 100)
   MN2 = MOLN2 * 28000
   VO2 = MOLO2 * .0821 * 294.2
   VN2 = MOLN2 * .0821 * 294.2
   TTOV = VO2 + VN2
   GL = TTOV
   TOTV = MOLO2 * .0821 * DK / (Pref / 760) + MOLN2 * .0821 * DK / (Pref / 760) VOLREF = TOTV
   XO = MOLO2 / (MOLO2 + MOLN2)
\[ XN = \frac{MOLN2}{(MOLO2 + MOLN2)} \]

t = 0

950 \hspace{1em} \text{END SUB} \hspace{1em} \text{UTUBE}
PROCEDURE UTUBESUM
'SUMMARIZE AND PRINT OUTPUT FOR U-TUBE

DEFINT U
SUB UTUBESUM

EUF = EU(U)
FUF = FU(U)
CRF = CU(U)

LPRINT"+++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++
LPRINT " U-TUBE EFFLUENT"
LPRINT"+++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++++
EUF = EU(U)
FUF = FU(U)
CALL PRESSURE(EUF, FUF, CRF, TGP, O2T, N2T, DPO2, DPN2, TODP)
LPRINT
LPRINT "Recirculation Flow rate = "; Qm; "L/min"
LPRINT
LPRINT "DISSOLVED OXYGEN LEVELS"
LPRINT " Initial = "; EU(0); " mg/L"
LPRINT " Equilibrium = "; EUF; " mg/L"
LPRINT "Oxygen Tension = "; O2T; "mm Hg"
LPRINT "Excess Oxygen Tension = "; DPO2; "mm Hg"
LPRINT "Oxygen as Percent of Saturation = "; EUF * 100 / ESAT; "%
LPRINT
LPRINT "DISSOLVED NITROGEN LEVELS"
LPRINT " Initial = "; FU(0); " mg/L"
LPRINT " Equilibrium = "; FUF; " mg/L"
LPRINT "Nitrogen Tension = "; N2T; "mm Hg"
LPRINT "Excess Nitrogen Tension = "; DPN2; "mm Hg"
LPRINT "Nitrogen as Percent of Saturation = "; FUF * 100 / FSAT; "%
LPRINT
LPRINT "****TOTAL GAS PRESSURE *****"
LPRINT "Total Excess Gas Pressure (DELTA P) = "; TODP; "mm Hg"
LPRINT "Total Gas Pressure (Percent Saturation) = "; TGP; "%
IF EUF > .1 THEN LPRINT "Nitrogen-Oxygen Ratio = "; N2T / O2T
LPRINT
LPRINT " ****OFF_GAS CHARACTERISTICS*****"
CALL OFFGAS
LPRINT : LPRINT

END SUB "UTUBESUM
APPENDIX C

Definitions of symbols and acronyms
Symbols and Acronyms

a - interfacial area between the gas and liquid phases, m²

AE - aeration efficiency, kg/(kWhr)

A_i - constant = 760/(1000 K_i), for i_th gas

Ar - Argon

B_i - Bunsen coefficient of i_th gas, L_gas / (L_{water} atm)

BP - local barometric pressure, mm Hg

Ci - dissolved concentration of i_th gas, mg/L

ΔC - change in dissolved gas concentration, mg/L

C_e - effluent DO concentration, mg/L

C_{in} - influent DO concentration, mg/L

C_o - initial dissolved gas concentration, mg/L

C_s - equilibrium or saturation concentration of dissolved gas, mg/L

CO_2 - carbon dioxide

D - liquid phase diffusivity

d - molecular diameter of gas molecule, Å

dC/dt - time rate of change in dissolved gas concentration, mg/(L-min)

DEM - oxygen consumption rate, g/hr or g/kg_{fish}/hr

DEM_i - oxygen consumption rate in production tank, g/hr
DEM<sub>mc</sub> - oxygen consumption rate in multi-tube clarifier, g/hr

DO - dissolved oxygen concentration, mg/L

e<sub>c</sub> - overall or combined efficiency of the compressor, dimensionless

e<sub>p</sub> - overall or combined efficiency of pump, dimensionless

FR - daily Feed Ration, kg/day.

g - acceleration due to gravity, m/s<sup>2</sup>

GBT - Gas Bubble Trauma

GW - gas to water volumetric ratio, dimensionless

H - Henry's constant, mg / (L mm Hg)

h<sub>g</sub> - headlosses attributed to two-phase flow, m

h<sub>w</sub> - pipe friction headlosses due to liquid flow, m

i - gas species identifier

k - ratio of specific heats of air, C<sub>p</sub>/C<sub>v</sub>

K<sub>i</sub> - ratio of molecular weight to molecular volume of i<sup>th</sup> gas, mg/mL

k<sub>L</sub> - liquid film resistance coefficient, m/min

K<sub>La</sub> - apparent gas transfer coefficient for respective gas, min<sup>-1</sup>

K<sub>La</sub><sub>20</sub> - apparent gas transfer coefficient at 20° C, min<sup>-1</sup>

m - mass flow rate of gas, kg/s

n - constant = (k-1)/k, dimensionless

N<sub>2</sub> - nitrogen
\( N_2 + \text{Ar} \) - nitrogen and argon combined as single gas species

\( O_2 \) - oxygen

OFR - oxygen feed ratio, \( \text{kg} \ O_2 / \text{g-feed} \)

OTE - oxygen transfer efficiency, %

OTR - oxygen transfer rate, mg/min

\( P_i \) - aqueous partial pressure of ith gas

\( P_i \) - absolute inlet pressure, kPa

\( P_2 \) - absolute outlet pressure, kPa

\( \Delta P_m \) - measured total excess dissolved gas pressure, mm Hg

\( \Delta P_{\text{uncomp}} \) - uncompensated pressure change, mm Hg

\( P_{N2+Ar} \) - combined partial pressures of nitrogen and argon, mm Hg

\( P_t \) - total gas pressure, mm Hg or atm

\( P_w \) - vapor pressure of water, mm Hg

PROD - mass production of carbon dioxide rate due to biological respiration, g/hr

\( \text{PROD}_t \) - carbon dioxide production rate in production tank, g/hr

\( \text{PROD}_{mc} \) - carbon dioxide production rate in multi-tube clarifier, g/hr

\( PW \) - primary oxygen feed gas water ratio, dimensionless

\( P_{wc} \) - power requirement of off-gas compression device, kW

\( P_{wp} \) - pump energy requirements, kW

\( Q \) - water flow rate, L/min
R - Gas Constant for a particular gas, kJ / kg.K
RAS - recirculating aquaculture system
RBC - rotating biological contactor
t - time, min
$\Delta t$ - finite element differential time step = 0.25 s
SLPM - volumetric flow rate of gas at Standard Conditions, L/min
*SLPM* - volumetric flow rate of gas measured at non-standard conditions corrected to an equivalent flow at Standard Conditions, L/min
$T_c$ - temperature, °C
$T_k$ - absolute temperature, K
TR - transfer rate, mg/min
TGP - percent saturation of total dissolved gas pressure, %
V - volume, L or m³
$V_b$ - gas bubble volume, L
$V_{mc}$ - multi-tube clarifier water volume, L
$V_r$ - water volume of one stage of rotating biological contactor, L
$V_t$ - production tank volume, L or m³
VTAF - Virginia Tech Aquaculture Facility
W - average fish weight, g
UE - utilization efficiency, %
Z - depth, m

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\( \alpha \) - ratio of \( K_{i,a} \) values (field water/clean water), dimensionless

\( \beta \) - ratio of saturation concentration (field water/clean water), dimensionless

\( \theta \) - constant = 1.024

\( \rho \) - density of water, kg / m\(^3\)

\( \chi_i \) - mole fraction of the \( i^{th} \) gas in gas phase

\( \gamma_w \) - unit weight of water, knN/m\(^3\)
Vita

Luther Graham Wood was born on September 12, 1964. The son of Wayne and Effie Wood, he grew up on a small farm near New Bern, NC where he spent much of his extra-curricular time either working on the farm, hunting, or fishing. He attended North Carolina State University from 1982-1987 completing requirements for Bachelor's of Science Degree in Agricultural Engineering and working part-time as a research technician in the Agricultural Engineering Department.

Much of his undergraduate course work and work experience were intended as preparation for a career in farm machinery design. However, during the final semester of his senior year he became interested in aquaculture and decided to pursue a career in Aquacultural Engineering thereby combining his love for fishing with engineering and research. Upon graduation, he accepted a job with the Department of Sea Grant at NCSU as an aquaculture technician working at the Pamlico Aquaculture Research and Demonstration Center. He worked at the Research center until 1989 gaining experience in fishery growout and hatchery management. In August of 1989, he was accepted by the Department of Agricultural Engineering at Virginia Tech as a candidate for Master's of Science Degree. At Virginia Tech, he gained experience in water quality analysis, ozone treatment as well as gas transfer processes for aquaculture while working part time as a graduate research assistant for the Virginia Tech Aquaculture Reseach Facility. He completed graduate course work in primarily in water quality and water/wastewater treatment to prepare for a career in aquacultural and/or enviromental engineering.

Luther Graham Wood