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Design and Analysis of Microalgal Open Pond Systems for the Purpose of Producing Fuels

A Subcontract Report

J. C. Weissman
R. P. Goebel
Microbial Products, Inc.
Fairfield, California

SERI

Solar Energy Research Institute
A Division of Midwest Research Institute
1617 Cole Boulevard
Golden, Colorado 80401-3393

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PREFACE

The Solar Fuels Research Division of the Solar Energy Research Institute manages a program of research and development on the production of microalgae and their conversion to liquid fuels (e.g., gasoline and diesel) for the U.S. Department of Energy, Biofuels and Municipal Waste Technology Division. As part of the effort, three subcontracts were awarded in 1983 to design and provide cost estimates for the construction of a microalgae facility on a scale suitable for commercialization.

This report is the final report submitted by Microbial Products, Inc. for the project entitled "Design and Analysis of Microalgal Pond Systems for the Purpose of Producing Fuels." This work was supported through SERI subcontract No. XK-3-03153-1 as part of the Aquatic Species Program.

Donna A. Johnson
Aquatic Species Program Coordinator

Approved for

SOLAR ENERGY RESEARCH INSTITUTE

Stanley R. Bull, Director
Solar Fuels Research Division

SUMMARY AND CONCLUSIONS

OBJECTIVES

The purpose of this report is to present a detailed analysis and design of a two acre experimental microalgal production facility that can be used to test the technical and economic potential of a commercial size facility (1000 acres). The design is based on extensive assessment of biological and technical requirements and important considerations such as resource use and availability and the impact of design assumptions on final product cost. Design feasibility and cost estimates are based on actual data obtained from standard cost estimating procedures, consultant input, and vendor quotes.

The proposed facility can be used to test the following important assumptions and concerns about microalgal biomass production.

1. Performance of a lowest cost design.
2. Efficiency of nutrient utilization, particularly carbon.
3. Achievement of high lipid productivity.
4. Reliability of a cost effective biomass harvesting method.
5. Longterm maintanence of biological and engineering standards of performance in a system with a high level of internal recycle.

DESCRIPTION OF THE EXPERIMENT

The site chosen for the experimental system is in Brawley, in California's Imperial Valley, on a one thousand acre farm which borders the Salton Sea. The climate is favorable for microalgal cultivation, with virtually a year round growing season. It is representative of the warmer, drier regions of the American Southwest. Ample land is available for both the experimental system and for potential scale up. Virtually any water resource can be simulated by varying the mixture of the two water resources available: water from the Salton Sea, which is as saline as ocean water, and grondwater from an existing well which produces 1700 gallons per minute. Flexible lease agreements can be negotiated with the land owners for use of the site and facilities.

The proposed experimental system is sized to allow validation of the commercial size facility with a minimum of extrapolation. The choice of two one acre growth ponds is based on an optimization of this validation requirement and cost of the experimentation. Three smaller growth ponds, of $50m^2$ each, are included for testing scaled down subsystem options. These can be constructed and operated at lower cost than the one acre versions, and used to determine which options show the most promise and thus should be tested at the larger scale. The $50m^2$ ponds will also be used in growth and harvesting studies. The harvesting subsystems for the one acre ponds are designed to directly emulate the full size subsystems so that translation of successful results concerning this important aspect, to the commercial scale, is assured.

The experimental system also includes a covered, anaerobic lagoon to be used for determining the extent to which nutrients can be recovered and recycled back to the growth ponds. This is essential to validate the economic predictions of the full scale design.

INNOVATIVE CONCEPTS, DESIGNS, AND TESTS

The designs and systems presented in this report include many concepts and experiments which are innovative. The design and operation of a low cost system is innovative in and of itself. The successful achievement of process and procuctivity goals in such a system would be an important advance in the state of microalgal technology and in its prospects for application. The cost effectiveness is realized by minimizing capital costs of the system and achieving efficient use of inputs. Extensive engineering analysis of carbonation, mixing, and harvesting subsystems has elucidated both the lowest cost, most efficient options and the essential parameters needed to construct, test, and evaluate these subsystems. The use of growth ponds sealed with clay and lined with crushed rock results in construction cost savings of 50% relative to ponds lined with synthetic membranes. In addition a low cost but efficient design allows improvements in technology, like productivity enhancement, to have the fullest impact on final product cost reductions.

In addition to the innovations in construction, the operational efficiency of the design is novel in being both higher and more feasible than that attained by any previous system concept of comparable scale. This efficiency is not achieved by fanciful optimism, but rather by concerted effort to ascertain what can be done and how to do it effectively. The recycle fraction of nutrients such as carbon, nitrogen, phosphorus, and water, must be high, realizable, and not achieved simply by substituting extreme capital and operating cost increases in one area for cost savings in another. The water chemistry analysis has led to operational specifications which minimize water use and virtually eliminate losses of carbon dioxide to the atmosphere. The carbon dioxide injection system is designed for 95% efficiency, but is still low cost. The construction of a large scale, covered anaerobic lagoon, to convert and recycle carbon, nitrogen, and phosphorus has not been attempted at the scale analyzed here. Yet, this recycle of non-product biomass is essential for achieving economic affordability.

The low cost biomass harvesting concept presented in this report is a major innovation, adapting the advances in other disciplines for application in microalgal systems. The concept is to tailor high molecular weight polymers for flocculation and coagulation of algal biomass, inducing characteristics which allow collection of the biomass in sedimentation ponds. These ponds are designed for inexpensive construction and operation. Successful achievement of performance goals will represent another major breakthrough in algal production technology.

Attainment of the lipid productivity assumed in this report, 30 gm/m^2/d average with 50% as lipid, in the system as outlined, would again represent a major advance. Algal strains with high lipid accumlating capability will be used in the experimental system to test productivity goals. The management strategies developed and the environmental conditions existing in the ponds will be evaluated for effect on productivity. Understanding the inhibition of productivity by high levels of dissolved oxygen, and developing means to alleviate it, is a possible route to achieving enhanced productivity.

CONCLUSIONS OF ECONOMIC ANALYSIS

The system described, with designs for efficient use of inputs included, is analyzed in the report in terms of the costs of the algal biomass produced and the unextracted lipid "oils" it contains. At the 1000 acre scale, the production cost of the biomass is $205/mt. The lipid only price is $62/bbl. The sensitivity analysis presented shows that both the recycle of nutrients that do not leave in the lipid product stream, and the low capital cost of the system are essential for attaining further product cost reductions. The impact of higher productivity is only significant under these conditions. Increased productivity must not be accompanied by a decrease in the efficiency of use of inputs, significant increases in system construction costs, nor by reduction in lipid content of the biomass. When system performance is maintained, a 50% increase in biomass productivity results in a 20% decrease in final product cost. If carbon is not recycled and system costs increase in proportion to the increase in productivity, then the cost reduction is well under 10%. Use of a system which is more than twice as expensive to construct as the one presented in this report renders the economics of fuel production untenable. Production costs are most sensitive to the price and recycle effiency of CO_2, the content of lipid in the biomass, and the combination of the cost of capital and the cost of system construction. However, the system must be designed and constructed so that each component which contributes to the final cost is minimized, in order for the concept to approach affordability and for further innovations to have any impact.

RELATIONSHIP OF THE PROPOSED EXPERIMENT TO PROBLEMS IN FUEL PRODUCTION FROM MICROALGAE

Any estimate of the cost of producing lipids from algal biomass is dependent on the assumptions used to generate the inputs to the economic model. It is necessary to determine which are most important and thus deserving of attention, and which are reasonably valid. Four questions must be answered in order to evaluate the overall feasibility of the concept: can inexpensive CO_2 be found and can it be used efficiently, does a large enough water resource exist which can make up for the losses incurred, can strains be found which form lipids at high rates of production and which will dominate the pond flora, and can methods be devised for harvesting the biomass inexpensively.

The resource evaluation questions cannot, for the most part, be answered at the present time. However, it is clear that the efficiency of utilization of the resources will be part of the determnination of how large the resource base must be. In the case of water, the production system must be operated with a high recycle ratio. This has yet to be tested for an extended period of time or on a meaningful scale at the recycle level needed here. This is proposed in the experiment. In the case of CO_2, the system must be designed to recycle as much of the non-lipid carbon as possible, the injection systems must be 90+% efficient, and the management of the pond operations must such that virtually no outgassing occurs. In the engineering analyses given in the report, several means are presented to do all of these. In the experimental plan, efficiency of CO_2 use is given the highest priority, including the testing of an anearobic lagoon for nutrient recycle.

The harvesting of the algal suspension is a problem that will require continuous attention in terms of applying new developments in polymer chemistry to a changeable pond environment. A repertoire of polymers must be available for use when a contaminant arises which has different flocculating characteristics from the desired strain.

The discovery of productive, lipid forming strains, which are highly competitive as well, is a basic requirement. However, productive accumulation of carbohydrates or protein has been looked for and found. There is no reason to believe lipid cannot be prolifically produced as well. The sensitivity analysis indicates that high lipid content, 40-50% of the biomass is necessary, but that overall lipid productivity need not be higher than 25 t/acre/yr. Before increases in productivity have too much effect, the operating costs of the system must be reasonably low. Once these costs have been lowered, the capital costs become more important. While on the surface it may appear that increased productivity is a panacea, closer examination reveals that this is so only under one condition: lowest CO_2 requirement and cost, combined with low capital cost which is not increased significantly in order to attain the higher productivity. The system designed, and to be tested during the proposed experiment, is a low cost system, operating efficiently in terms of CO_2, and limited in terms of productivity only by the availability of productive algal strains.

The last requirement, that the strain be dominant, may depend on how unusual lipid accumulation by algae turns out to be. Enough strains which are known to store lipids must be identified, so that once a universal harvesting method is developed (as the polymer flocculation method potentially is), one could be relatively sure that nitrogen depletion will lead to the desired product despite the identity of the particular strain. This needs to be demonstrated.

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SECTION 1.0

INTRODUCTION

1.1 OBJECTIVES

The objective of this report, as described in the statment of work, is to develop a cost effective design for a microalgae culture facility, that meets specified location, resource, and product requirements. The conceptual design and operating options presented are based on field and laboratory experiments, conducted by Microbial Products, Inc., engineering cost analysis, and assessment of the work in the literature.

In developing the design and analysis, the major decision in choosing among identified alternatives has been potential cost effectiveness and economic efficiency. This approach is based on the fact that the ultimate success of microalgae-based fuels will be determined by market economics.

1.2 STATE OF THE ART

An evaluation of the state of the art of algal biomass production can be based on information from actual producing plants and from extrapolations of results of experimental facilities. In the former, the costs of constructing the system approach $100x10^3$/acre, the resource base is well-defined but limited and relatively expensive, and the final cost of the biomass is $2-5/dry lb. The costs derived from extrapolation vary extensively and are generally based on unrealistic assumptions used. For this reason, only the results of Weissman and Goebel [1], Benemann et al [2], and Laws [3] will be used in the following discussion. These investigations are, in any case, among the most recent and thus the latest in the state of the art.

The three major technical and economic concerns of algal biomass production are: the cost and availability of the two major resources (CO_2 and water), the costs of constructing and maintaining the system, and the design and costs of the biomass harvesting and processing systems. At present none of these concerns has been resolved for a process which must produce the biomass at a cost of $.05/lb as must be accomplished for fuel precurser biomass.

1.2.1 Major Resources

Despite the effort that has been put into identifying the water resources available for large scale algal biomass production, no suitable resource has yet been identified. The reasons for this are 1) data on groundwaters are sparse, non-random, and difficult to compile and 2) the selection criteria have not been fully applied so that the resource data have not been properly evaluated. It is not possible to conduct a comprehensive investigation of water resource availability. Rather it seems more useful to state the criteria which any water resource must meet in order to be suitable for use in a low cost algal biomass production system. In discussing the water availability, it is important to consider the interdependencies among the water, land and CO_2 resources required for algal production systems.

In existing production facilities, system sizes vary from 1-20 acres and the amount of water required is small. Thus water from wells with special characteristics or inexpensive irrigation waters, already available for agriculture are used. When unique growth media are required, as in the high bicarbonate medium for Spirulina production, recycling of the growth medium is necessary, further lowering the water demand. For fuel production from microalgae, the scale increases 50-5000 fold, changing the resource question entirely. Now the agricultural water, which is not too costly per se, is not utilizable because the competition from conventional agriculture makes the land too expensive. This leaves only three water resources left for consideration: seawater, groundwater that is too saline for agriculture, and agricultural drainage water.

In the U.S., coastal property with access to seawater is extremely expensive. However, since so much work has been done in seawater, and since this is really the only unlimited water resource, we will consider the consequences of using it. One thing that must be kept in mind when considering the use of seawater is that much more water must be used since the initial salinity is already high. One really cannot consider even a 50-100% increase in the salinity, due to evaporation, for two reasons. First, salinity restricts the number of species that will tolerate the medium and there are already very many species screening criteria. Second, the growth rate and productivity of algae on waters with increasing salinity, above seawater, decreases rapidly even for species which tolerate the higher salinity [4].

Saline groundwater is the prime candidate for algal biomass production, because it may be the only resource available in large enough quantities at low enough cost, where land is inexpensive. The groundwater resources identified to date [5], are most likely useable, after conditioning, as will be shown below. However, the size of the resource base is only speculative at present. Nonetheless use of saline groundwater will be a basic assumption in this report and serves as an example of a theme to be repeated throughout the report: the assumptions used will often be speculative, but it will be shown that the alternatives are much too costly to be seriously considered. The major effort will be to demonstrate that the speculative assumptions are the only economic alternative.

Agricultural drainage waters are becoming a larger and larger resource as irrigation waters are made more available in the semi-deserts of the U.S. southwest. These drainage waters are becoming recognized as a major waste disposal problem. Although use of these waters poses some of the same problems as the use of irrigation water (available where land is relatively expensive) the credits for evaporating some of this water may outweigh the expenses. The following analysis will thus cover the use of seawater and several types of saline groundwaters, delineating the consequences that the characteristics have on the design of low cost algal biomass systems.

The major resource necessary for fuel precurser production from microalgae is carbon dioxide. Even at $40/ton, CO_2 is likely to comprise over 50% of the operating costs. Its requirement is proportional to biomass production so that increasing the output of a system, per acre, does little to alleviate the cost burden of CO_2. CO_2 is available in moderate to large quantities from three resources: wells, refinery cracking processes, and power plant off gases. The logistics of siting a pond system close to both water and CO_2 resources poses one of the most difficult problems. We will not delve into this area except to say that at least in the future enough CO_2 is available for large system use even if power plants must be built to supply it. The design considerations will emphasize efficient use of CO_2, assumed to be available at various specified costs. Algal biomass production systems in operation today use commercial CO_2 produced at refineries and trucked or trained to the pond site. Costs of this range from $50-100/ton. This is the CO_2 resource that will be specified for the experimental system developed.

1.2.2. Pond Construction Costs.

State of the art systems are small and very expensive per unit area to construct. Although the recovery of the cost of construction can be spread over many years, none of the existing systems could make a profit if the product value were not greater than about $10/dry lb. High construction costs and the large scale of the systems required to produce energy combine to discourage private investment. Thus the designs specified in this report were devised to keep capital costs of construction as low as possible while still achieving production goals. For example, pond lining with plastic membranes was considered too expensive to buy and install initially, as well as being a potential maintenance problem. A crushed rock lining, on the other hand, is much less expensive. The theme of our approach is repeated: intensive algal production has not been attempted in large ponds, sealed with clay and lined with crushed rock, but we feel that in order for the process to approach affordability, this option must be specified.

1.2.3. Biomass Harvesting.

Perhaps the foremost technical problem in low-cost algal biomass production has been recovering the cells from suspension. Besides simple sedimentation, there is no method yet developed that is inexpensive. All straining, filtering, and chemical flocculation methods presently being analyzed are too expensive. Induced bioflocculation, possibly aided by small doses of chemical flocculant, followed by sedimentation or flotation are potentially effective means of harvesting which would not be too costly. Both conventional and the proposed means will be analyzed in the body of the report.

1.3 ORGANIZATION OF THE REPORT

The report contains thirteen sections including this introduction. In sections 2.0 - 5.0 the analyses which serve as the basis for design criteria are presented. The chemistry of the water resource is the subject of Section 2.0. The water used and the blowdown ratio assumed determines the carbon storage capacity and pH arange of operation. This determines the transit time between carbonation stations, which is needed as an input to determining the pond size.. The design criteria for two different types of carbonators is the subject of Section 3.0. In Section 4.0 two methods for mixing are analyzed: paddlewheels and gas-lift pumps. The results are stated in terms of efficiency and power requirements to obtain channel mixing velocities of 20 cm/s. Harvesting systems is the subject of Section 5.0. The methods analyzed include flocculation, with and without small chemical doses, followed by either a two-step sedimentation system, air/DO flotation using a foam collection device, microstraining, or belt filtration. Section 6.0 summarizes the analyses of preceding sections, provides an overview of other important considerations not previously presented, and discusses the implications of the analytic results on the design and operation of a large scale algal production system. In Sections 7.0-9.0, the design of a large scale ponding system is presented along with the construction costs. The cost of operating such is presented in Section 10.0. The sensitivity of the economics of algal biomass production to changes in the initial assumptions and design is the subject of Section 11.0. Finally, the experimental system which is to be used to validate the operation of the proposed system, is presented in Sections 12.0 and 13.0.

SECTION 2.0

WATER RESOURCES

2.1 CHARACTERISTICS OF WATER RESOURCES

The chemical composition of the water resource has several primary consequences for both the design and operation of the system. Pond size and/or the number of carbonation stations per pond is determined by the carbon storage capacity of the water. Since it is least expensive to restrict the "power" corridor to one end of the channels, one carbonation station per pond is optimal unless this criterion restricts pond size to less than about 10 acres. In that case, pond construction will be more expensive because more carbonation stations will be needed per pond, requiring a more extensive CO_2 and electrical distribution system. The water chemistry will also exert a selection criterion on the algal species. Thus the species must be screened for performance on the anticipated water resource. The water available may also require some conditioning prior to use. For algal systems this is likely to be the case when the water hardness (Ca in particular) and carbonate exceed the solubility product at the projected pH range. The situation is more acute when the calcium concentration exceeds the alkalinity. In both cases some water softening will be required. The conditioning is achieved at lowest cost by the addition of sodium carbonate, followed by settling. If removal of magnesium is also required, then both lime and sodium carbonate need to be added.

In Table 2-1, several water resources are shown along with the corresponding carbon chemistry and water conditioning requirements. Also shown are the outgassing consequences, the precipitation of calcium carbonate, and the levels of dissolved CO_2. Figures 2-1,2,3 show the pH dependency of these factors and thus the data upon which the specified pH range of operation was chosen. This range is limited on the acid side by loss of CO_2 due to outgassing and on the basic side to 1) low CO_2 concentration and 2) calcium carbonate precipitation. The latter is critical in waters in which there is a lot of calcium which could precipitate causing a turbidity problem and especially critical where the alkalinity is low, as in seawater. Here calcium carbonate precipitation results in a major loss of buffering and carbon storage capacity. When the calcium concentration is low (1-4mM) and the alkalinity moderately high (12-40mM), precipitation is not a problem. The figures were constructed assuming infinite solubility of calcium carbonate. At finite solubility (indicated in the figures) the figures are still accurate when carbonate alkalinity is much greater than calcium concentration. On the other hand, seawater chemistry changes considerably as calcium carbonate precipitates.

Table 2-1. Chemistry of Water Resources

	Seawater	Salton Sea	Type I L	Low TDS	Type II L	Exp't.
BDR	.95	.95	.125	.25	.125	.125
Ca removed ppm	-	600	500	0-40	125	1100
Na_2CO_3 added,ppm	-	1500	1250	0-100	0	2750
Cost/acre/yr,$	-	5000	2400	200	0	300
TDS,gm/l	35	34	32	10	32	32
$K1x10^{-6}$	1.05	1.05	1.05	.88	1.05	1.05
$K2x10^{-9}$	9.55	9.55	9.55	.18	9.55	9.55
$KSx10^{-6}$	.5	.5	.5	.05	.5	.5
Ca,mM	10.3	10.3	.5	1-4	1-4	1-4
Alk.,mM	2.3	2.3	32	12	32	24
pH range	6.5-8.0	6.5-8.0	8-8.5 8-9	7.5-8.5 7.5-9.0 7.5-9.5	8.0-8.5 8-9	8.0-8.5 8.0-9.0
C stor,mM	.85	.85	4 8	0.9 2 3.5	4 8	3 6
C out, $gm/m^2/d$	5	5	2 2	3 3 3	2 2	<2 2
CO_2,mM	.68-.02	.68-.02	.25-.06 .25-.01	.4-.04 .4-.01 .4-.003	.25-.06 .25-.01	.20-.05 .20-.01
C storage time,hrs	1	1	5 10	1 2.5 5	5 10	4 8

Pond depth = 20 cm, Prod.=$4gm/m^2/hr$=.83mM C/hr, K_L=0.04m/hr for outgassing

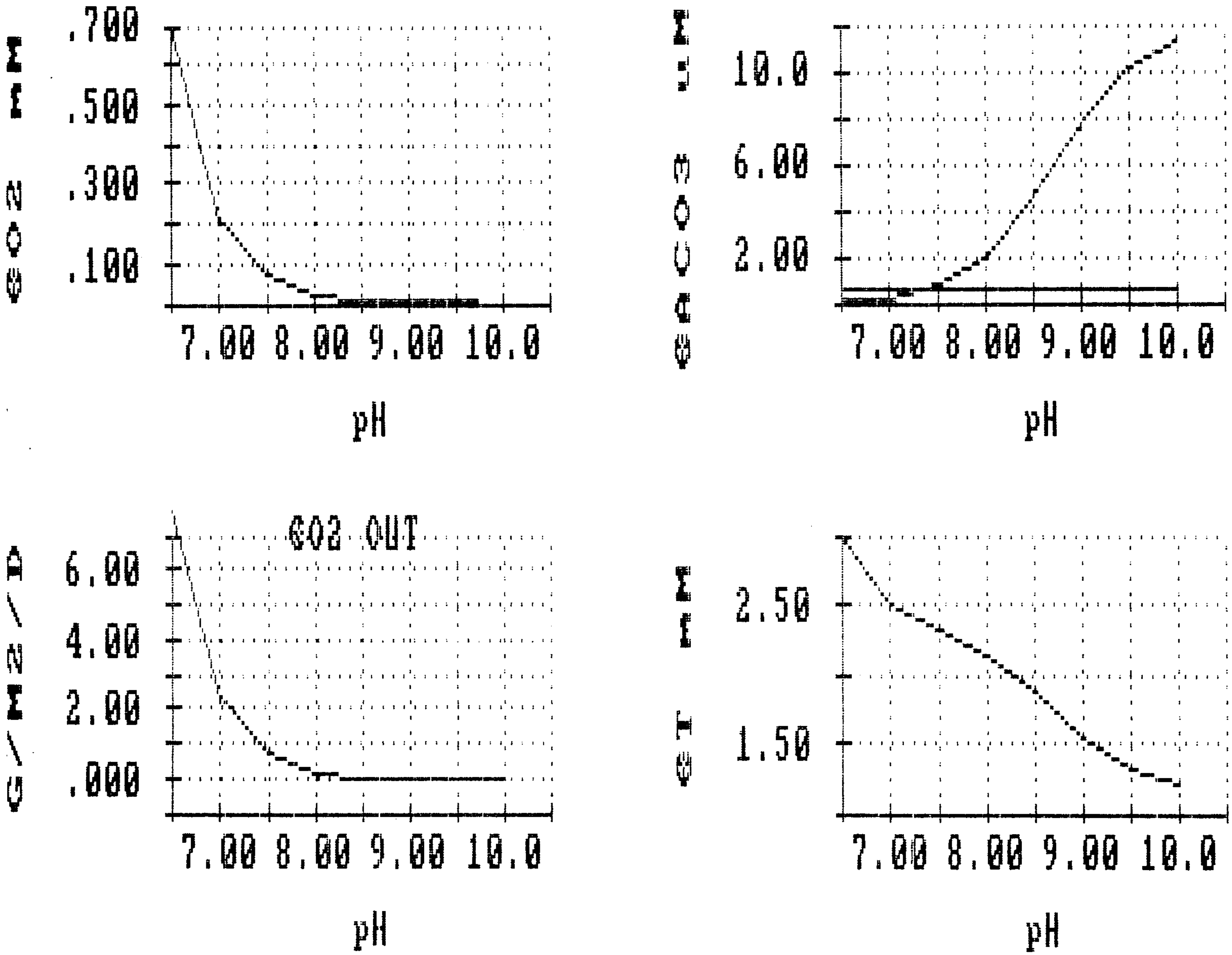


Figure 2-1. Carbonate Chemistry of Seawater (or partially conditioned Salton Sea)

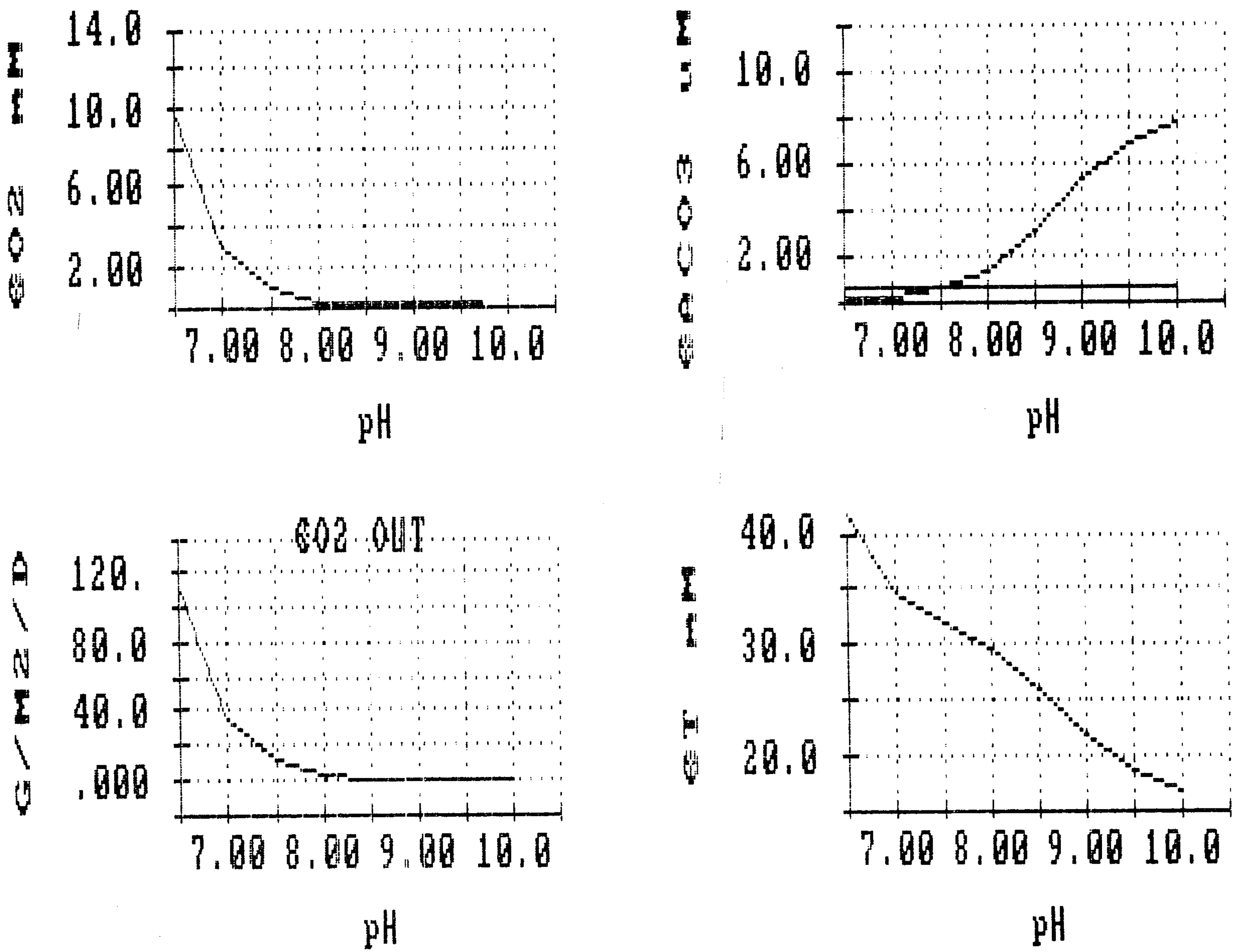


Figure 2-2. Carbonate Chemistry of Conditioned Type II Water: Alkalinity = 32 meq/L
Blowdown = .125

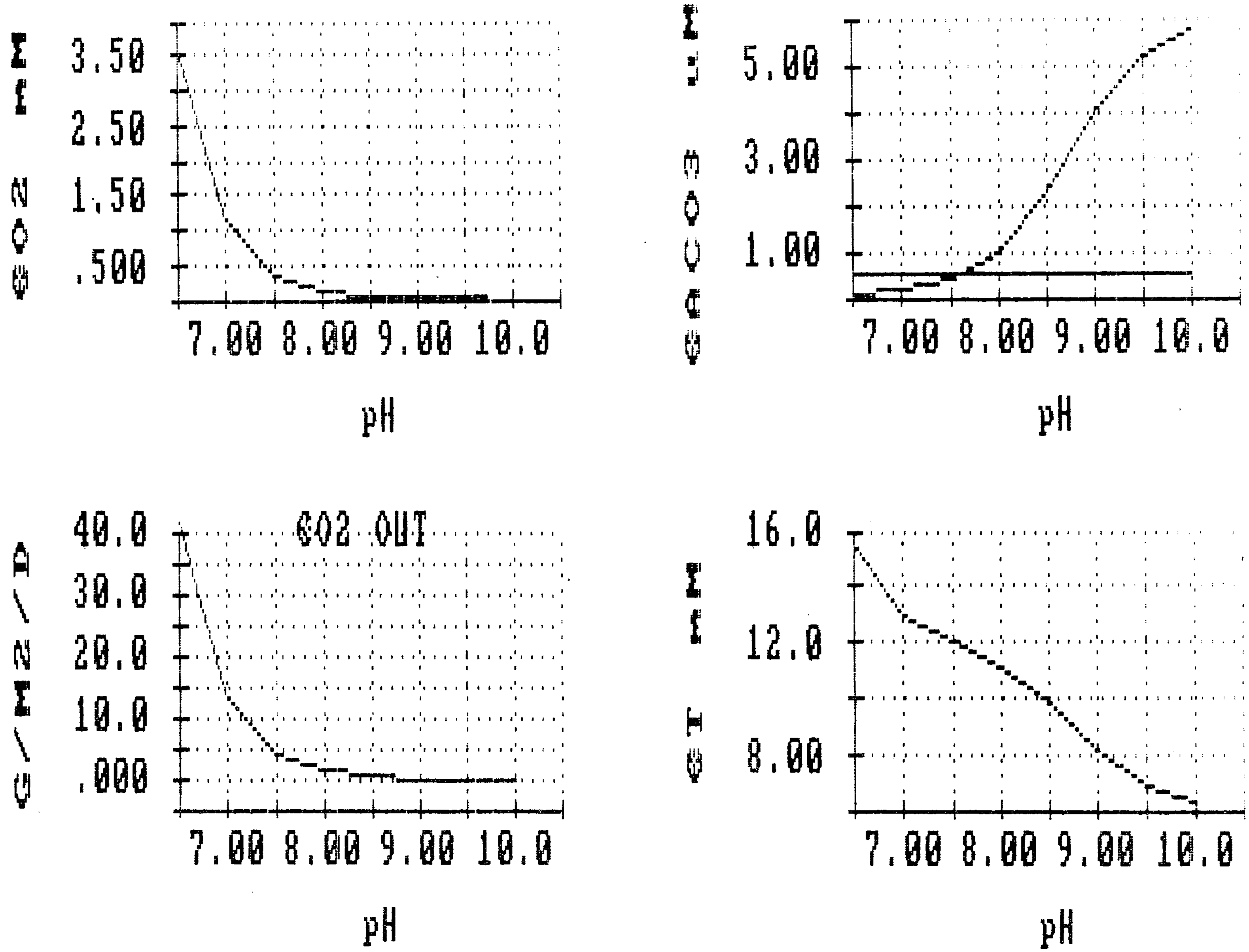


Figure 2-3. Carbonate Chemistry of Type II Water: Alkalinity = 12 meq/L
Blowdown ratio = .25

Six water resources are characterized in Table 2-1. The first two represent high TDS waters which cannot bear much concentration due to evaporation without severely limiting the number of species which will grow productively. Thus seawater is a resource that requires substantial blowdown. The Salton Sea water can be used at the proposed experimental site in southern California to simulate seawater as a once through resource, after conditioning to remove more than half of the calcium. (Salton Sea water has 2.5 times as much calcium as regular seawater). For seawater the upper limit on pH is rather strictly set by calcium carbonate precipitation because the alkalinity is so low, as discussed above. The lower limit is set, by CO_2 outgassing, at pH about 6.5. Due to the low alkalinity, only about one hours worth of carbon can be stored in the water. If pond depth were halved only a half hours worth of carbon could be stored.

The saline groundwaters examined by SERI were evaluated for their usefulness in open ponds systems. Generally speaking, the major water resource cited, the Type I Low is unsuitable for this application. It has too much calcium, requiring $2400/acre/yr for conditioning in chemical cost alone. Although the Type I moderate has a more favorable ratio of alkalinity to calcium, it ends up requiring more sodium carbonate because it must be blown down more. The low waters can bear an eight fold concentration whereas the moderate can bear only a twofold concentration. The Type I high has a favorable alkalinity to calcium ratio meaning that the calcium can be removed by equilibration with air in a settling pond to allow precipitation of calcium carbonate, followed by sedimentation. However the water is so high in TDS that it cannot be concentrated further and thus too much of it would be required. Not only does it become very expensive to pump so much water, but there is no evidence that enough of this resource exists. So for the purposes of analysis the Type I waters are unsuitable. If similar water, but lower in calcium were found it could be utilizable. At 50 tons/acre/yr algal productivity, and a blowdown ratio of .125, the chemical cost of calcium removal is about a half cent per kg dry wt. of biomass per 100 ppm calcium.

A similar analysis reveals that the Type II water of TDS = 4ppt can be used despite its high calcium. The much higher alkalinity results in precipitation of most of the calcium as calcium carbonate upon equilibration with the air. The low Type II water can be concentrated many fold in terms of TDS. It is not clear whether high TDS or high alkalinity would first limit the extent to which this water type could be concentrated. Whether magnesium would precipitate as the water concentrates is an issue that must be addressed by experimentation. If not there would be no water conditioning costs beyond a settling pond and possibly some flocculant to remove calcium carbonate precipitates. If magnesium needed to be removed, lime would be used. A complete water conditioning analysis would be needed. However, preliminary analysis indicates that about 6-8 mM lime would reduce the magnesium content 60%, the alkalinity 70%, leave little residual calcium, and increase the pH to about 9-9.5. This would cost about one cent per kg biomass. The uncertainty in the water conditioning status results in an uncertainty in the alkalinity of the treated water. Starting with an excess of alkalinity over calcium of 10 mM, an eight fold concentration by evaporation in the ponds would lead to a "steady state" alkalinity of 80mM. This would require a pH of close to 9 to avoid considerable outgassing. It might be difficult to find organisms which grow well at such a high alkalinty and pH (although the CO_2 concentration would still be quite high).

2.2. LARGE SCALE DESIGN WATER

The water source selected for the large scale system design is Type II saline groundwater with an initial TDS of 4 ppt. Since removal of calcium, some magnesium, and concommitant alkalinity is likely, the design includes a water holding pond to be used to sediment precipitates. After conditioning and evaporative concentration with a .125 blowdown ratio the steady state salinity is 32 ppt with an alkalinity of 32-60 mM. It is assumed that calcium and magnesium have been lowered to levels which do not cause precipitate problems. Of course, it is also assumed that an adequate resource base exists for this type water.

Given this water resource, with an alkalinity of greater or equal to 32 mM, the upper limit of pH is dictated by the growth response of the algae to CO_2 concentration. If the organism grows well at or below air levels of CO_2 then the pH could go up to 9 or higher. If it requires higher CO_2 then a pH of 8.5 is more suitable. The low limit is set by outgassing, which increases steeply below pH 8 due to the high alkalinity as shown in Figure 2-3. But, also due to the high alkalinity, large amounts of carbon can be stored with a small change in pH. With just one carbonation station per pond, a pond of 40 acres could be operated. For a smaller pond, the pH range could be narrowed, further lowering the outgassing loss (by increasing the low pH limit) or increasing the minimum CO_2 concentration (by decreasing the high pH limit).

2.3. EXPERIMENTAL SYSTEM WATER RESOURCES AND GROWTH MEDIA

The proposed experimental site is located in Brawley, California in the Imperial Valley. The site borders the southwest side of the Salton Sea. Water from the sea is available by pipeline or by drilling a shallow well. An existing 1700 gpm well provides brackish (2.5 ppt) groundwater. As discussed above, saline groundwaters can be expected to contain high levels of alkalinity after concentration. On the other hand waters that are originally highly saline, like seawater, contain little alkalinity. The water resources available on the proposed site can be mixed in various proportions, after conditioning, to obtain waters that are representative of both high and low alkalinity resources. The primary experimental alkalinity will be 24meq/l. In some cases, however it will be convenient to operate at lower alkalinities. The basic formula for the experimental water resource for the demonstration system is a mixture of Salton Sea water (after removal of all of the Ca) and well water in a ratio of about 19:1 achieving a TDS of 4 ppt and an alkalinity of about 3 meq/l. After evaporation, the TDS is 32 ppt and the alkalinity is 24 meq/l. Any starting TDS is achievable by using a different ratio of Salton Sea water to well water. This allows variation of the concentration of major ions (although not independently) and blow down ratio. Since only the Salton Sea water needs conditioning, the cost per acre per year is about $300. The conditioning can be achieved by addition of lime and soda ash if both calcium removal and magnesium removal is required. If only calcium is to be removed, soda ash alone is added. The practical methods of water conditioning that would be used at the large scale will be tested in small scale holding ponds.

SECTION 3.0

ANALYSIS OF CARBONATION SYSTEMS

3.1 INTRODUCTION

In the last section, the carbon storage capacity of several waters was analyzed. This, along with the constraints of minimizing dissolved CO_2 level at the lowest pH and maximizing it at the highest pH defined both the pH range of operation and the transit time allowed between carbonation stations. This transit time was based on the summertime daily demand averaged over 12 hours, which is equal to the production of 4 $gm/m^2/hr$ of algal biomass. When demand is lower, the pH range can be narrowed and/or the carbon inflow rate diminished. In this section the means with which this maximal rate is supplied from the gas stream to the pond liquid is analyzed. The basic chemical engineering of gas-liquid transfer is presented in two appendices; AI and AII. The results for two different types of carbonators, a bubble cover and an in-pond sump, are discussed below. Along with the parameters relevant to each manner of transfer, the consequences on the mixing requirements of the ponds are analyzed.

3.2 COVERED AREA CARBONATORS--BUBBLE COVERS

By covering a small percentage of the pond area with a membrane-covered structure submerged at its edges, gas transfer can occur passively through the pond surface under the cover. The cover serves to trap a gas volume with a high concentration of CO_2. The flow of gas under the cover can be turbulent or laminar, depending on the placement of the injection sites. Laminar flow allows somewhat higher stripping efficiencies since the gas bled off will contain the lowest CO_2 concentration. A gas bleed is necessary since desorption of oxygen and nitrogen will occur under the cover. If the percent coverage is low the bleed rate needed is even a lower percentage of the inflow rate, so it has little effect on the carbon transfer efficiency.

The percent coverage of the pond depends on the transfer rate of CO_2 under the cover. As detailed in Appendix AI, this depends on the hydaulics under the cover, i.e., the rate of turnover of an element of water surface. Several techniques can be used to increase this surface renewal rate. Anything which increases the hydraulic slope under the cover, like increasing the flow velocity by lowering the depth, increases the renewal rate. However, since the area of coverage is not negligible, the effect of this increased mixing on the overall power and head loss must be considered. Alternatively, the roughness coefficient under the cover can be increased. The means proposed for doing this is to increase the roughness of the underside of the cover and place the cover partially in contact with the water. This "rippled" cover design can be effected by using a corrugated material for the cover. In this manner the average surface element lifetime for a 20 cm deep pond mixed at 20 cm/s, could be decreased from 50 s to less than 4 s (corrugation spacing of 5 cm). This, too, will increase the head loss and mixing power required but only about 5-10%, much less so than if the same low turnover time were achieved by lowering the depth to 10 cm and increasing the velocity to 40 cm/s under the cover. As shown below this latter design increases the overall mixing head and power required by 50%.

The results of the analysis are shown in Table 3-1. Three basic cases are detailed: a pond 30 cm deep mixed at 15 cm/s, a pond 20 cm deep mixed at 20 cm/s, and a shallow pond only 10 cm deep mixed fast at 30 cm/s. Several simplifying assumptions were made in constructing the table; the major ones being that temperature is constant at 30° C, that the water can store the carbon transferred, and that the dilution by O_2 outgassing is negligible. It can be seen from the results that a low concentration CO_2 gas phase (like flue gas) would require much to high a percent coverage in all cases. In order for the coverage to be less than 5%, a turnover time of less than 10 s is required, which is theoretically equivalent to a mass transfer coefficient of .04-.05 m/hr. The coefficients used in the construction of Table 3-1 were obtained from empirical correlation of stream flow data (see Appendix AI). The coefficients measured by Weissman and Goebel [1] in high rate ponds 20 cm deep mixed at 20 cm/s were higher by a factor of 3.4 than those calculated using the correlation. If the higher, measured values are used then each of the values for % coverage in Table 3-1 could be reduced by multiplying by .3. The percent coverage for cases 2a and 3 then become 8% and 4% respectively. If in addition, the rippled cover is used instead of the smooth one, then the covered area for case 2a is reduced from 8% to 2%. In case 3, this reduction is only from 4% to 2% since the turnover time was initially shorter.

The assumption here is that a covered area carbonator is feasible if the percent coverage is low (about 2%). This may be achievable with a smooth cover in shallow, fast mixed ponds, in 20 cm deep ponds if the area under the cover is made shallow, or in these moderately deep, moderately mixed ponds if a rippled cover is used. The effects on mixing head and power can be expected to be significant when a portion of the pond is made shallow, but small when the roughness and wetted perimeter are increased by a rippled cover.

3.3 IN-POND CARBONATION SUMPS

An alternative way of transferring CO_2 to the pond water is to inject the gas at or near the bottom of a sump which spans all or part of the channel. As in the case of the covered carbonator, theoretical results must be taken as only approximate since the actual process involved is complex. In all of these gas transfer problems, the real answer is only obtained from the operation of pilots. Nonetheless, the theoretical treatments do give a sense of what is possible and economically feasible. The chemical engineering treatment of sump carbonation is detailed in Appendix II. The major results of that analysis and the major uncertainties involved will be summarized first.

Table 3-1. Covered Carbonator - % Areal Coverage

Case	1a	1b	2a	2b	2c	3a	3b
Depth cm	30	10	20	10	20	10	10
Vel. cm/s	15	45	20	40	20	30	30
K_L m/hr	.013	.070	.022	.061	.085	.047	.085
C in*	380	2200	640	1900	2465	1400	2465
C req.*	166	166	166	166	166	166	166
% area	44	8	26	10	6.7	12	6.7
head % 1a	100	170	300	450	---	1700	---
power % 1a	100	170	400	600	---	3500	---

Carbon supply (C in) and carbon demand (C req.) are in mmoles/m^2/hr
Assumes pure CO_2 equilibrates to 29mM and back pressure is negligible.
For flue gas (15% CO_2) multiply coverage by 6.7.
The depths and velocities listed are those under the cover. Case a) of each group gives the depth and velocity for the pond since no changes occur under the cover in these instances.
Cases 2c and 3b refer to the rippled cover, which determines K_L.
K_L = (diffusion coefficient/turnover time)$^{1/2}$
The values in the table are based on an empirical correlationgiven in the literature. Actual values have been measured to be over three times faster, see text.

Gas transfer is qualitatively different in sumps. Under bubble covers the hydraulics of the flow under the cover determines the mixing that is needed to disrupt the thin film which effects transfer and mixes the dissolved gas with the bulk liquid. In sumps it is the movement of the bubble swarm relative to the flowing liquid which "renews" the transfer surface. Thus the behavior of bubble swarms determines transfer rates. This behavior is quite complicated and the models used thus only appoximate the actual process.

The major uncertainties in the modeling are the bubble size distribution, the consequent average rise velocity relative to the water, and thus the average rate of gas-liquid transfer. In the treatment in Appendix II, an average bubble rise velocity, relative to the water, of 30 cm/s was assumed. It was further assumed that the mass transfer rate distribution is very narrow. These assumptions have empirical justification, in that it has been found that after about .5-1.0 meter of rise the bubbles coalesce to a minimum equilibrium size with these characteristics. However it has been noted [6,7] and measured [1,6,7] that in shallow sumps the transfer rates exceed, by a large factor, those predicted on the basis of this equilibrium. Consequently, we will analyze the sensitivity to assumed transfer rate. Given a transfer rate, one can estimate the average bubble diameter and then the rate of rise. With this information, the basic parameters of sump transfer can be calculated. These parameters are the depth of the sump required to strip a specified percentage of CO_2 and the gas hold-up which determines (for a given sump depth) the head gained by the water column in co-current flow or the head opposing the water flow in counter-current flow. One more case of interest is when the sump does not contain a baffle to direct the water flow. In this case of "lateral" flow the gas has little effect (except causing some extra turbulence) on water head.

The average hold-up,e, is given by the following formula for co-current flow:

$$e = Q_g C_{pe} / (Q_L (1 + V_B/V_L) + Q_g C_{pe})$$

Q_g is the gas flow rate, Q_L is the water flow rate, V_B is the average rise velocity of the bubbles relative to the water, V_L is the water velocity, and C_{pe} is the average pressure correction factor for the compression of the gas at subsurface depths. Appendix Table AII-1 gives values for this correction factor as a function of depth. For pure CO_2 a further correction factor is required since the volume of gas decreases with time as gas is absorbed. This factor is the log mean of the flow rate at the entrance to the sump and at the exit. There is also a pressure factor to account for the fact that most of the gas volume is at the lower depths. This is somewhat arbitrarily assigned the value of $p_{atm}/p_{2/3}$, but is only of the order of 10% for shallow sumps. For counter-current flow the term inside the parentheses in the denominator becomes $V_B/V_L - 1$ where $V_B > V_L$ and both are considered absolute (positive) values. The hold-up is related to the lift or head by:

$$h_o - h_i = h_o e - K V_L^2/2g$$

where h_0 is the level of outflow above the sump sparger, h_i is the distance between the inflow liquid level and the sparger level, g is the acceleration of gravity, and K is a friction factor for losses upon entry, transit through, and exiting the sump. These losses are small at low liquid velocities.

As discussed in Appendix II, the stripping rate of oxygen has been found to be .7%/s or, at a bubble rise velocity of 30cm/s in still water, .7%/foot. The transfer for CO_2 is related to that of oxygen by the ratio of the diffusion coefficients to the two-thirds power times the ratio of the solubilities. At 20^o C this ratio is about 18, yielding a stripping rate for CO_2 of 12.5%/s. Table 3-2 shows the sump depth required for stripping 95% of the inflowing gas from the bubbles given several different stripping rates per second. This stripping rate has to be considered an unknown parameter since measured values have been found to depend on sump depth, some have been found to be much greater, and the bubble size is known to depend on sparger type and flow rate as well[6,7]. For low flow rates, spargers which produce fine bubbles (and hence are more expensive) can be used. These will likely provide transfer rates higher than those calculated based on the oxygenation experiences, since it may take a considerable fraction of the sump height for an equilibrium bubble size to be reached. As flow rate increases, due either to the use of recycle or in the flue gas utilization case, not only does initial bubble size increase in a device specific manner, but the type of device may need to be changed to one which can handle the higher flow rates at moderate cost. Ususally this means that the device produces larger bubbles to begin with. To add to the uncertainty, the salinity of the medium affects the equilibrium bubble size and the rate of attainment of equilibrium due to the inhibitory action of charged species on bubble coalescence. Significant enhancement of transfer rates has been seen at TDS as low as 5 ppt [8].

The lowest rate shown in Table 3-2 corresponds to the rate found for oxygenation (adjusted for CO_2), the next lowest is considered the minimum value that would exist in a shallow sump when purified CO_2 is used, and the last is a value which is similar to that found by Weissman and Goebel [1] in a water column of 20 cm. The table shows the sump depths for co-current, counter-current, and lateral (no baffle to direct water flow) flow. In addition, the bubble size and bubble rise velocities calculated from the analysis of Aiba et al[7] which would yield the given stripping rate, are shown. These values must be taken as highly approximate, but do aid in choosing sparger type. As bubbles get above .2-.25 cm, they change from being rigid spheres to having deformable walls. One consequence of this is that the rate of gas transfer from the bubbles becomes somewhat insensitive to bubble size because the volume to surface area ratio decreases with deformation. At even moderate gas flow rates, the turbulence prevents the bubble size from getting very much larger than .5-1 cm. In the end, only experimentation will yield reliable results and design specifications.

The cost of construction of a sump depends on sump depth. Not only does the material requirement increase with depth, but in addition, deep sumps require more expensive construction methods. For this reason, the sump depth is to be kept below 1.5 m or so. Costs of construction in this range of depth are relatively low.

A method that can be used to keep the sump shallow is to cover it with a membrane, capture the gas, and recycle it via a blower back down to the sparger. Given a sump depth of 1.5 meters, the recycle ratio required to achieve 90-95% (using pure CO_2) or 80% (using flue gas) overall stripping of carbon is shown in Table 3-3 for several different conditions. A lower removal efficiency is used for flue gas because the back pressure from dissolved CO_2 becomes significant as carbon is stripped from the dilute gas. Since the depth of the sumps cannot be predicted precisely, recycle adds flexibility to the ways in which a given sump depth can be made adequate.

Table 3-2. Effect of Assumed Transfer Rate on Sump Depth

Transfer Rate,% Removal/s	13	25	50
Bubble Diameter, cm[1]	>.2	.125	.08
Rise Velocity,cm/s[2]	30	15	9
Distance Traveled/s, cm			
Co-current	50	35	29
Lateral	30	15	9
Counter-current	10	--	--
Time Required, 95% Overall Removal, s	23	12	6
Sump Depth, 95% Overall Removal, m			
Co-current	11.5	4.2	1.7
Lateral	7	1.8	0.5
Counter-current	2.3	---	---

Liquid velocity = 20cm/s

[1] d Rate$^{-2/3}$ [7]

[2] [7] and Stoke's Law

Table 3-3. Sump With Gas Recycle

	Transfer Rate % Removal/s	Overall Removal,%	Recycle Flow / Inflow
Pure CO_2			
Co-current	25	90	.37
	25	95	.44
	13	90	1.8
	13	95	1.9
Lateral	25	90	.04
	13	90	.9
	13	95	1.0
Counter-current	13	90	.05
	13	95	.1
15% CO_2			
Co-current	25	80	.96
	13	80	6.5
Lateral	13	80	3.0
Counter-current	13	86	.01

Sump Depth = 1.5 meter
Liquid velocity = 20 cm/s

Table 3-4 gives the average hold-up as a function of CO_2 demand for co-current flow in a sump of depth 1.5 m. This gives a range of head lift to be expected during carbonation. The assumed stripping rate may be too high for flue gas relative to pure CO_2 since the gas flow rates, and consequently the hold-up are so high. The lift for the flue gas case in which inflow of gas is provided 24 hours a day is just sufficient to mix the 8 hectare pond. However, the CO_2 injection efficiency is quite low.

Table 3-5 gives results for counter-current flow. Here the water velocity can be lowered by widening the sump beyond the value of the pond depth to prevent wash-through of the average bubble, and to reduce the head opposing the channel flow. As can be seen from the table, the opposing head is already significant when pure CO_2 is used and prohibitive when flue gas is used. To a good approximation, the hold-up is inversely proportional to sump width. An equation is given in the Table 3-5 relating hold-up to transfer rate, sump volume, and gas flow for the case in which the liquid velocity is equal to the average bubble rise velocity but opposed to it. In this case hold-up depends on the steady state ratio of gas to liquid volumes. It can be decreased by deepening the sump.

The values of hold-up and head shown in Tables 3-4 and 3-5 were calculated for an eight hectare pond, which was the size chosen for the large scale design. If a pond size of four hectares was chosen, halving the flow of CO_2 into the sump, no adjustment is required in hold-up and head if the length to width ratio is changed so that channel width is also halved (thus halving the liquid flowrate). By maintaining the same length to width ratio of 20, the liquid flowrate only decreases by a factor of 1.4. Thus the ratio of gas to liquid flowrates in the sump decreases by a factor of 1.4, decreasing hold-up and head by the same factor at low flowrate ratios, and somewhat less at high ratios. Head attained with gas lift pumps is discussed more fully in Section 4.0.

Table 3-4. Average Hold-up and Lift in 1.5 m Deep Carbonation Sumps

	Co-current, transfer rate = 25%/s, V_B = 15cm/s			
CO_2 req.,gm/m^2/hr	8.6	8.6	4.3	4.3
Gas Composition, % CO_2	100	15	100	15
No Recycle, Overall Removal = 71%				
Input Gas Flowrate, l/s	167	1129	83	561
Gas flowrate/Liq.flowrate	.056	.65	.028	.32
Hold-up, %	2.9	26	1.5	14.7
Head, cm	4.4	39	2.2	22
With Recycle				
Overall Removal, %	95	80	95	80
Input Gas Flowrate, l/s	116	925	58	462
Gas flowrate/Liq.flowrate	.06	1.04	.03	.52
Hold-up, %	2.9	36	1.5	21
Head, cm	4.4	54	2.2	33

CO_2 req. = (1.2)(1.8) x algal productivity (either 4 or 2 gm/m^2/hr, includes 20% for increased C content of high lipid producing algae.
Pond area = 8 hectare
Input gas flowrate at 1 atm, 30^o C.
Liquid flowrate through sump = 1725 l/s (43.1 m wide channel, 20cm deep, V = 20cm/s, sump spans channel, sump width equals pond depth)
For pure CO_2 cases, gas flowrate (for ratio) is log mean of gas flow rate at sump bottom and sump top, pressure corrected.
See Table 3-3 for recycle ratios.

Table 3-5. Average Hold-up and Lift in Counter-Current Sumps

Counter-current, transfer rate = 13%/s, V_B = 30cm/s				
CO_2 req.,gm/m^2/hr	8.6	8.6	4.3	4.3
Gas Composition, % CO_2	100	15	100	15
No Recycle, Overall Removal = 86%, Sump Depth = 1.5 m				
Input Gas Flowrate, l/s	128	860	64	430
Gas flowrate/Liq.flowrate	.03	.50	.015	.25
Hold-up, %	5.3	48	2.7	32
Head, cm	8.0	72	4.1	48
With Recycle, Sump Depth = 1.0 m				
Overall Removal, %	95	85	95	85
Recycle Ratio	.31	.17	.31	.17
Input Gas Flowrate, l/s	116	870	58	435
Gas flowrate/Liq.flowrate	.047	.59	.023	.30
Hold-up, %	8.1	52	4.2	36
Head, cm	8.1	52	4.2	36

Counter-current, $V_B = V_L$

Pure CO_2:
  Holdup = Input gas flowrate/transfer rate/sump depth/sump Xsec. area
Flue gas: hold-up gets very large as unabsorbed gas builds up, until channeling occers.

For conditions see Table 3-4.

3.4 RECYCLE OF NON-LIPID CARBON FROM EXTRACTION RESIDUES

Since carbon is one of the most expensive inputs to the system, and since only part of the carbon is extracted in lipid products, recycle of the carbon remaining in the unextracted residues must be considered. An effective method of processing the residues is to anaerobically digest them. A covered lagoon is specified in the large scale design for this. With 60% of the algal biomass as lipid and 90% of this extracted, the residues contain 46% of the algal carbon. 65% of this will degrade in the lagoon to methane and CO_2 in a ratio of 6:4. The other 35% will be split between the two liquid compartments of the lagoon: the sludge and the lagoon water. Both the gaseous commponents and the lagoon water (effluent) can be recovered. The digester gas is combusted to operate an engine generator. The resultant flue gas is about 14% CO_2, and contains 31% of the original algal carbon. Mixing this with the pure CO_2 input, in a ratio (by volume at STP) of 7:1, yields a gas containing 35% CO_2. If this is then injected into the ponds with a 90-95% efficiency, and the dissolved carbon in the digester effluent is recycled back to the ponds, the result is a decrease in overall CO_2 input from 2.4 kg/kg algae to 1.6 kg/kg (see Table 6-3).

The 35% CO_2 gas phase can be introduced into the ponds just as any other gas phase. Now, however, an overall stripping efficiency of 90% is assumed as opposed to 95% with pure CO_2 since backpressure reduces the driving force near the top of a sump. Again, the transfer rate is not certain. With 90% overall removal, the sump would have to be 5.3 m deep if the transfer rate is 13%/s and 1.3 m deep if the rate is 25%/s, for lateral flow of the pond relative to the bubbles. If the gas is injected co-currently, more head can be gained than in the pure CO_2 case due to the larger gas volume. For the two transfer rates, the sump depths required are 7-9 m and 2.3-3 m respectively. The lifts obtained are 28-53 cm amd 17-27 cm respectively. Thus the lift gained by co-current injection of a mixture of pure and recycled digester flue gas matches the mixing requirements of the system when moderate transfer rates are assumed.

The recycle of non-lipid carbon has perhaps the greatest potential for reducing the cost of the lipids produced of any single factor. This will be more fully discussed in several ensuing chapters.

3.5 OXYGEN DESORPTION

The desorption of oxygen, and nitrogen to a lesser extent, affects the efficiency of CO_2 transfer in both sumps and under covers. The ratio of oxygen mass transfer to CO_2 transfer depends approximately on the ratio of driving forces, since diffusion coefficients are similar. Assuming no CO_2 back pressure, which is very nearly true in the pure CO_2 case even with 95% absorption, and a dissolved oxygen concentration that is 500% of saturation, the dilution of CO_2 by O_2 under a covered area carbonator is only about

10%. That is, at steady state the gas phase under the cover would be over 90% CO_2 and the bleed rate would be about 10%. If CO_2 moved by plug flow across the "ripple" covered channel, then at the end the transfer rate would be slowed somewhat by oxygen dilution, but the gas bled off would have a higher oxygen concentration. Similarly, in sumps the CO_2 transfer rate would decrease near the top of the sump where the dilution of the gas phase with oxygen is greatest. This dilution would not affect transfer significantly in the pure CO_2 case, because ample driving force of CO_2 would remain until very near the top of the sump. Since flue gas is already dilute in CO_2, further dilution would also not be significant except at the tail end of transfer. It should be noted, however, that the back pressure of dissolved CO_2 may be significant when carbonating moderately alkaline waters to near neutral pH when flue gas is used, with or without further dilution by oxygen.

Potentially more serious is the desorption of supersaturated oxygen in a sump due to nucleated bubble formation. Using the mass transfer coefficient for outgassing through the surface of a 20 cm deep pond mixed at 20 cm/s and the supersaturation level of 500% found by Weissman and Goebel [1], about 40% of the oxygen produced outgassed due to mass transfer through the surface of the pond. Since these conditions existed for prolonged periods of time, 60% can be estimated to have desorbed due to nucleated bubble formation. If oxygen loss in a sump, for example, reduced steady state levels to 200% supersaturation, the mass transfer loss through the pond surface would be only 25% of the oxygen produced. At the lower DO nucleated bubble formation would be substantially reduced, implying that about 50% of the oxygen produced would need to be removed in the sump to maintain such a low DO. The rest of the oxygen desorbs, due to nucleated bubble formation, through the pond surface. The rate of desorption through the surface by bubble formation (as distinguished from mass transfer) calculated from the above condition (500%) is .007 mM/s. In order for the DO to be dropped from 500% to 200% by sparging air in a sump with gas residence time of 4 s, the desorption rate would have to be 25 times as great. At this point no data exists concerning the desorption of supersaturated oxygen due to carbonation or aeration. However if such a reduction was attained it would dilute the gas volume of pure CO_2 by 30-80%, depending on the ratio of input gas flow rate to the flow rate of pond liquid through the carbonator. This would have impact on sump design, i.e., sump depth required for a given transfer efficiency, especially in sumps with recycle. The desorption of supersaterated oxygen and the impact on sump design will be tested in the proposed experimental system. Of course, if very efficient removal of oxygen were attainable, a separate sump just for this purpose would be constructed ahead of the carbonation sump. In the case of flue gas carbonation, the dilution is 5-10%, which would further aggravate the backpressure problem for transfer near the top 25% of the sump.

The desoprtion in sumps by sparged gas will be used to determine the practicality of using aeration as a means of preventing the build-up of oxygen to inhibitory levels, i.e., to determine how many sumps would be required to maintain the DO within specified limits.

3.6 CHEMICAL ENHANCEMENT OF CARBON DIOXIDE TRANSFER

A detailed discussion of the chemical reactions of CO_2 with other dissolved species, and the effect of this on transfer coefficients, is given in Appendix I. Even at the highest alkalinities and pH imagined for algal ponds, the enhancement due the reaction of CO_2 and hydroxyl ions is negligible. The pH independent, first order reaction of CO_2 and water to form bicarbonate is fast enough to enhance transfer rates that have not already been enhanced hydraulically by increased surface renewal. The two processes are competitive, so that the chemical enhancement is still very small in sumps, where surface renewal rates are on the order of 100-150 per second. The covered area systems were designed so that surface renewal rates were as high as possible to increase transfer rates. The renewal rates discussed above were on the order of .25/s for 2% coverage and .016/s for 13% coverage. Chemical enhancement due to a first order reaction goes as the square root of 1 + k/s where k is the reaction rate constant and s is the surface renewal rate. k is equal to about .02/s. Thus physical transfer rates would have to be small, meaning high areal coverage, for chemical enhancement to have much effect. It may be of interest that several catalysts which have significant effect on the first order rate constant may be present in an algal system. These include carbonic anhydrase which is often secreted by algae and phosphate which is a media component. Enhancement due to these can be tested by comparing transfer in media not including these species to that in media which does contain them.

3.7 LARGE SCALE SYSTEM CARBONATOR DESIGN

The Base Case design specification of the carbonation system for the large scale system will be taken as a 1.5 m deep sump for the introduction of pure CO_2. Pure CO_2 mixed with anaerobic digester flue gas serves as a case with significant cost reduction possibilities. The use of unpurified power plant flue gas would most likely require deeper sumps. In addition, unless the power plant is actually within about 1 km from the pond system, the transport of dilute CO_2 becomes too expensive [2]. Thus even if power plant derived gas is used to serve multiple large pond systems [9], it may well have to be purified for transport.

The shallow depth of the sump is still considered adequate due to the flexibility provided by gas recycle. The introduction of the CO_2 will be controlled by pH on an on/off basis. Due to the lag time for return of the water to the carbonation station, 2.5 hours for an eight hectare pond, the low limit pH set point must allow for storage of enough CO_2 to meet the maximum demand for two transit times. At lower demand the upper pH reached in the pond will not be as great as when the demand is high. Thus the flow rate of CO_2 must be able to meet the expected average hourly demand, which may change over long time periods (seasons). An alternative way of controlling CO_2 input is to control the flow rate of gas so that a specified amount of carbon is stored. This could be accomplished with a microcomputer controller, programmed to calculate or determine from a table, carbon storage as a function of pH change. One would want to use such a system in situations where the alkalinity is low, requiring many carbonation stations per unit area. Such a scheme would preclude installing more stations than needed to meet the maximum hourly demand, i.e., only enough carbon is needed for one as opposed to two transit times.

3.8 EXPERIMENTAL SYSTEM DESIGNS AND TESTS

The experimental system is designed to allow determination of the carbonation parameters and design specifications needed to remove most of the uncertainty that now exists. In addition, several of the alternative types of carbonation systems described above will be constructed and tested. Yet the primary method is still considered to be shallow sumps with pure CO_2. Tests will include measurement of overall transfer efficiency, in both uninoculated medium and algal suspensions, as a function of sparger type, gas flow rate, gas to liquid flowrate ratio, sump depth, and pond DO. DO removal will also be measured. Covered carbonators with smooth and rippled surfaces will be tested for transfer efficiency as a function of covered area, hydraulic slope under the cover, and bleed rate. If necessary gas recycle will be tested in sumps as well.

SECTION 4.0

ANALYSIS OF MIXING SYSTEM

4.1 OPEN CHANNEL FLOW

Some form of mixing is required to maintain cells in suspension, to prevent thermal stratification, and to disperse nutrients. The most widely used formula for open channel flow is the Manning's formula:

$$V = \frac{R^{.67} S^{.5}}{n}$$

where: V = velocity, meters/sec
R_h = hydraulic radius, meters (= depth for wide channels)
S = hydraulic slope = head loss/unit length
n = Manning's roughness, sec meters$^{-0.3}$

For a given velocity and channel length, the equation can be rearranged to solve for the head loss:

$$h_L = \frac{LV^2 n^2}{R^{1.33}}$$

where: h_L = head loss, meters
L = channel length, meters

Typical values of n are 0.010 for very smooth surfaces, 0.014 for unfinished concrete, 0.017-0.025 for smooth earth (canals and ditches), 0.029 for gravel. A value of 0.018 was used in the head loss calculations for the large scale ponds. These ponds will be lined with graded crushed rock, rolled to a smooth finish. The mixing power requirement is given by:

$$P = \frac{9810\ AV^3 n^2}{d^{.3} e}$$

where: P = power, watts
A = pond area, meters^2
e = overall mixing system efficiency

Note that the power varies as the cube of the mixing speed, and actually increases slightly at lower depths. The efficiency term includes both the hydraulic efficiency of the pumping device and that of the drive. Overall efficiencies can be a high as 70% for large centrifugal pumps, whereas 30-40% is more typical for paddle wheels or airlift pumps.

4.2 PADDLE WHEELS

Paddle wheels have emerged as a preferred method for mixing high-rate ponds for the following reasons: (1) They are well matched to the pumping requirements of high-rate ponds in that they are high volume, low head devices (i.e. high specific speed); (2) Their gentle mixing action minimizes damage to colonial or flocculated algae, which improves harvestability; (3) They are mechanically simple, requiring a minimum of maintanence; (4) Their drive train can easily be designed to accomodate a wide range of speeds (high turn-down ratio) without drastic changes in efficiency; (5) They do not require an intake sump, but simply a shallow depression for maximum efficiency. Some of the disadvantages are: (1) The paddle wheel itself must be custom designed; (2) They are large relative to other types of mixers, especially at higher heads; (3) They are fairly expensive, though not particularly so for a low shear type pump. (4) For practical purposes, the maximum head is limited to 0.5 meters. This would only be a constraint in very large (>20 hectares) ponds, or at high (>30 cm/sec) velocities in moderate sized ponds.

The two primary competitors for pond mixing are air lift pumps and Archemedes screw pumps. The air lift pump has a lower initial cost, but is generally considered to be a low efficiency device. Actual efficiencies appear to be very design specific, although theoretical efficiencies (see Section 4.3) are comparable to paddle wheels. In practice, the potential efficiency of paddle wheels are probably greater. About 10% of the losses in paddle wheel systems can be attributed to the variable speed unit, which may not be necessary in the long run. Air lift pumps probably produce higher shear forces than paddle wheels, which could interfere with autofloccualtion. The Archemedes screw pump has been successfully used for high rate pond mixing, but has a higher first cost and requires an intake sump. At high heads (>0.5 meter), the screw pump may be more economical than paddle wheels.

Although paddle wheels are chosen in the large scale system design, the experimental system will include one 50 m^2 pond mixed with an air lift. This will allow studies of combined mixing and carbonation as outlined in the next section. The optimization of mixing systems, although of interest, is not as critical a technical issue as harvesting or species control.

The actual design of the paffle wheels is covered in Section 7.3.

4.3 GAS LIFT MIXING

4.3.1 Introduction

Gas lift mixing is an alternative to paddlewheel mixing which doesn't require large, custom fabricated mechanical parts. It is an interesting option in systems for which carbon is supplied via an in-pond sump due to the potential for combining the carbonation and mixing systems. In this section, gas lift is analyzed in its own right, but is also related to the carbon demand of the system. When flue gas is used for carbonation the lift may be derivable from the carbonation alone for part of the time, but most likely an air supply system would need to be available to substitute for the flue gas when carbon demand is low and during the nighttime. When pure CO_2 is used, supplementary air would always be required.

4.3.2 Gas Lift Mixing Without Carbonation

The basic relations needed to analyze gas lift mixing are derived in Appendix II. Two of them are used in the following discussion and are thus reproduced. The efficiency of the gas lift is given by, for adiabatic compression of air:

$$E = E_d E_{mc} \cdot \frac{C_{pe}\, g\, h_o}{nRT\left(\frac{p_2}{p_1}^{\frac{k-1}{k}} - 1\right)\left(\frac{K}{K-1}\right)} \cdot \frac{V_L}{V_B + V_L} \cdot \frac{H_g}{H_g + \frac{K_F V_L^2}{2}} \qquad (4\text{-}1)$$

Cpe is an average correction factor for compression of air with depth
g is the acceleration of gravity, 980 cm/s/s
h_o is the depth of the sump in cm
V_L is the liquid velocity in the sump in cm/s
V_B is the average rise of the gas bubbles relative to the liquid in cm/s
H is the lift in cm
$K_F V_L{}^2/2$ is the total friction loss for liquid entering, traversing and leaving the sump
E_{mc} is the compressor motor efficiency

E_d is the dynamic efficiency which covers losses due to the irreversible nature of the real compressor and is taken equal to .7

$nRT\left(\frac{p_2}{p_1}^{\frac{K-1}{K}} - 1\right)\left(\frac{K}{K-1}\right)$ is the theoretical adiabatic work of compression of the gas/cm^3

$K = C_p/C_v = 1.4$ for air

nTR $=(1.218\times10^{-2}$ mol $^oK)(8.31\times10^7$ erg/mol $^oK) = 1.012\times10^6$ for 1 cm^3 gas entering the compressor at 1 atm at any temperature.

The lift is equal to the gas hold-up times the sump depth minus fluid flow losses:

$$H = e\,h_o - \frac{k_F V^2}{2g} \qquad e = \frac{Q_g/Q_L}{\left(1+\frac{V_B}{V_L}\right)/C_{pe} + Q_g/Q_L} \tag{4-2}$$

For shallow sumps, <10 meter or so, Table 4-1 shows that the efficiency doesn't change with sump depth for constant ratio of gas and liquid velocities. For moderate liquid velocities, all terms in equation 4-1 are nearly equal to unity except for the efficiency factors and the velocity ratio term. Thus the efficiency is governed by the velocity ratio. Table 4-2 shows the overall efficiency as a function of liquid velocity. Gas bubble rise velocity was taken as 30 cm/s which is consistent for a situation where the flow rate of gas is not expected to be very low. The efficiency is only 25% at the design liquid velocity of 20 cm/s. Increasing this to 40 cm/s, by narrowing the sump to baffle distance to 10 cm, increases efficiency to 34%. Further increases in efficiency require even higher liquid velocities (relative to gas bubble rise rate) which could be attained by directing pond channel flow into a shallow sump leading into a set of draft tube entrances. The liquid velocity increases linearly with the ratio of channel cross sectional area to total draft tube cross sectional area. If this option is chosen then the efficiency can be much higher. Usually, however, large velocities are used for systems with very large lift, because the velocity head, which is significant at high velocity, is easily lost upon exit from the draft tube.

The lift is determined by the ratio of gas and liquid flow rates through the sump as well as the ratio of the velocities. The two dependencies on velocity ratio act together in that decreasing the ratio of gas to liquid velocities, increases both the efficiency and the lift. Table 4-3 gives the efficiency in terms of liquid velocity and lift for a system with three draft tubes. The fluid losses are significant when the liquid velocity exceeds 100 cm/s.

Table 4-1. Effect of Sump Depth on Air Lift Efficiency

h_o (m)	C_{pe}	p_2/p_1	$\dfrac{C_{pe} h_o g \times 3.82\times10^{-7}}{((p_2/p_1)^{.286} - 1)}$
1	.954	1.1	.986
1.5	.934	1.15	.980
3.0	.966	1.29	.966
5.0	.816	1.48	.944
8.0	.740	1.77	.919
10.0	.699	1.97	.905
50.0	.365	5.84	.769
100.0	.245	10.68	.698

Table 4-2. Gas Lift Efficiency as a Function of Liquid Velocity

V_L	$V_L/(V_L + 30)$	Efficiency*
10	.25	.15
15	.33	.20
20	.40	.24
30	.50	.30
40	.57	.34
50	.62	.38
60	.67	.40
80	.73	.44
100	.77	.46

* All other efficiency factors assumed equal to .9x.7..95 for shallow sumps. Velocity head losses neglected.

Table 4-3. Lift and Efficiency of a Draft Tube Air Lift System

Area ratio	2.5	5	10	25
Pipe D, cm	120	85	60	38
V_L, cm/s	50	100	200	500
$V_L/(V_L+V_B)$	.62	.77	.87	.94
Hold-up	.15	.17	.20	.22
Lift, cm	20	20	20	20
Tube rise length, cm	130	120	100	95
$K_F V_L{}^2/2g/\text{lift}$	.02-.08	.07-.32	.27-1.3	1.7-8.0
Efficiency. %	38-36	45-36	43-24	20-6

For three draft tubes
Ratio of gas to liquid flowrate = .3
$E_d E_{mc}$ = .7x.9 = .63

$$K_F = \{K_{ent} + (K_{bend} + L/D)f\}V^2/2g + (0-1)V^2/2g$$
$$= \{.04 + (15 + 200/D)f\}V^2/2g + (0-1)V^2/2g$$

This assumes a flared entrance, one segmented 90° bend, smooth pipe, and a total draft tube length of about 200 cm.

The range in loss and in efficiency corresponds to 0% or 100% loss of the velocity head.

4.3.3 Combined Gas Lift Mixing and Carbonation

The combination of gas lift mixing and carbonation requires matching the mixing system performance to the constraints set by the carbonation. As shown above, the air lift efficiency increases with increases in liquid velocity. However, the injection efficiency of CO_2 increases only with gas bubble rise time which decreases as liquid velocity increases unless the height of rise is made greater. As an example consider narrow, long draft tubes into which the channel flow has been directed. The liquid flow rate increases in proportion to the area contraction ratio. This increase leads to a shortening of the time a gas bubble stays in contact with the liquid and thus a decrease in the overall transfer efficiency, unless the tube length is appropriatley lengthened. The degree of difficulty encountered in trying to optimize a combined gas lift and carbonation system, depends on the distance which must be bridged between the initial configurations and requirements of each. With pure CO_2 the lift attainable was much lower than required to mix the pond while with flue the lift was much greater than necessary. When carbon is recycled from digester flue gases mixed with pure CO_2, a good match is achieved between transfer and mixing needs during times of maximal carbon demand.

When purified CO_2 is used as the gas for mixing, the total flow rate of gas is restricted to that required to carbonate the pond. It was concluded in Section 3.0 that even for an 8 hectare pond with 43m wide channels, the gas flow rate would be low enough to attain a transfer rate of 25%/s and a gas to liquid flowrate ratio of about .02 for sumps without recycle and .056 for shallower sumps with recycle. The former sumps were 4.2 m deep the latter only 1.5 m deep. The lift provided by these configurations was only 4 cm. Only by increasing the pond size dramatically (and also increasing the carbon storage capacity of the medium) can the lift attainable be increased faster than the lift required. That is, if the pond area is increased X fold, maintaining the same length to width ratio, the ratio of gas flow to liquid flow increases by $X^{1/2}$. This increases the lift. The lift is also increased by increasing the liquid velocity, accomplished by adjusting the baffle placement. Many workable configurations are possible. For example, a 40 hectare hectare pond could be mixed at 20 cm/s by pure CO_2 sparged into a 5.5 m deep sump, when CO_2 demand is highest. At other times, air would be needed to dilute the CO_2. Since the efficiency of such a setup would be 35-40% and air would be required about 75% of the time, the overall power requirement would be about the same as with a paddlewheel mixing system. For smaller pond sizes the mismatch between lift needed and pure CO_2 required for carbonation is too great. If long draft tubes are used, the rise time of the bubbles still needs to be 12 s for 95% stripping efficiency. The tubes would need to be 20 m deep (long) and have an area of one tenth of the channel cross section to provide a lift of 20cm.

With flue gas, as with dilution of the CO_2 by air, the lift increases due to the increased flow. It can be assumed that the increase in gas flow rate would lead to an increase in average bubble size and hence an increase in the depth needed to strip CO_2 efficiently. For flue gas, at a maximal average carbon demand of 8 gm/m^2/hr, the lift attainable in a 6.2 m deep sump (80% injection efficiency) is over 74 cm. The gas flow could be lowered by carbonating 18-24 hours a day into a medium containing sufficient alkalinity to minimize pH fluctuations in the upward direction during the day and downward during the night. This lowers the lift to 39 cm, about twice that needed. The most promising possibility for combined processes would be to supply the carbon as flue gas over 24 hours during the high productivity months but only over 12 hours when productivity is lower. During these times air would substitute for the flue gas at night, when mixing speeds could be lower, reducing the power input. Since the cost of transporting the flue gas to the site includes the cost of pressurization sufficient for injection the overall power costs would be low. In order to reduce the lift enough, the baffling would need to be arranged to lower the liquid velocity in the sump to, say, 10 cm/s. Then, an eight hectare pond could be mixed at 20 cm/s by flue gas in a 5 m deep sump when demand for CO_2 is (averaged over 24 hours) 2 gm/m^2/hr. The airlift efficiency would be low, about 15-20% and the deep sump required would be a substantial capital cost and in increased maintenance cost, but no additional cost would be incurred for mixing power. Of course the power plant needs to be very close to the pond system to transport it unpurified.

Given the uncertainties in the parameters for both carbonation and gas lift taken alone, design and cost estimation of a combined system must await the results of the experiments proposed. A configuration which allows economic use of CO_2 for gas lift may be possible, but it must be kept in mind that carbon is the most expensive input into an algal biomass production system. Any system for combining air lift and carbonation must use the CO_2 with the highest efficiency, essentially as high an efficiency as carbonation alone would achieve.

4.4 MIXING SYSTEM SPECIFICATION

For the 1000 acre system, both paddlewheel and airlift mixing systems were designed. The former requires twice the capital to construct as the latter, but there is more experience on which to base its efficiency. As discussed in Section 11.0, the impact of this cost difference on production costs is small. The airlift system is comprised of three draft tubes, each 2 m long, as described in Table 4-3. The calculated efficiency for this configuration was the same as for the paddlewheel system, about 40%. However, confirmation of this estimate, from the experimantal system, is necessary.

No specific design is specified for a combined gas lift mixing- carbonation system. The gas transfer parameters are not known well enough to formulate a meaningful design at this time. Again, the impact of capital cost savings is minimal.

SECTION 5.0

ANALYSIS OF HARVESTING SYSTEMS

5.1 INTRODUCTION

Harvesting algal biomass from dilute suspension (.05-.1%) is ususally a two step process: a primary concentration step which reduces the volume of water 20-50 fold and a secondary step, usually centrifugation, which further reduces the volume 5-10 fold. The capital and operating costs of industrial centrifuges are so high that it precludes their use without the initial concentrating step. In special circumstances, as in the harvesting of _Spirulina_ with screens, only one step is necessary. This is due to the high value of the algal biomass in this case, as well as to the large size of the filaments and the manner in which they pack on screens which allows high flow rates of water. An inclined screen or vibrating screen is used. Generally, however, a harvesting process must be able to concentrate varied types of cells, especially when the harvester effluents are to be returned to the growth ponds. Any organisms which the harvesting process fails to remove from the recycle stream, will be given a potentially significant competitive advantage in the growth pond [11]. Only in rare circumstances is the growth medium so selective for a chosen organism that this advantage will not result in dominance of an unwanted species. Thus, the harvesting process must not only remove the desired organism, but also any potential contaminants.

Four primary harvesting methods were examined: microstraining, belt filtering, flotation with float collection, and sedimentation in two stages. Each of these methods discriminates on a size and/or density basis in performing the biomass separation. The strainer and filter are, of course, size discriminators, with performance characteristics that improve as the algal particle size increases. However, the filter is also more efficient for oraganisms which form a mat, thus decreasing the effective pore size. Sedimentation and flotation devices collect biomass on the basis of both size and density. Since the operation of each of these devices becomes tenuous for small cells (<6-8 microns), and since the harvester must, in a strict sense be universal in its removal capabilitiy, some universally effective method is required to transform all cells into a form which can be removed by the chosen collection device. As will be shown, the major cost of a harvesting system lies in this transformation process.

5.2 CAPITAL COSTS OF PRIMARY HARVESTING DEVICES

The four types of primary harvesting devices vary considerably in their installed capital costs. Conventional dissolved gas flotation units cost about .25 million dollars per million gallons per day of suspension. The costs of microstrainers, belt filters, and settling ponds, on the same basis are .09, .12, and .05 million dollars. Nevertheless, if no pretreatment of the cells were required to 1) make each device work efficiently and 2) make each work universally, the capital cost differences would not translate to large differences in annualized production cost. The most expensive device, the dissolved gas flotation unit, would add 50% to the capital costs of the system, but these only contribute 10-25% of the annual production. The important point is that none of the collectors functions either well or, more importantly, universally on its own. A rough calculation shows that if conditioner is added to the suspension prior to primary harvesting, for each cent of conditioner added per kg of biomass harvested, the production cost increases by about 10%. The conclusion is that while cost differences do exist among various primary harvesting devices, the impact of these differences is small compared to the potential impact of the cost of pretreating the biomass so that any or all of the devices are, in fact, useable.

5.3 PRETREATMENT OPTIONS

There is ample experience with chemical flocculation of suspended solids to improve the separation characteristics. The thrust of recent work with chemical flocculants is to manufacture new chemicals, or devise new combinations of known chemicals, which reduce the dose and cost of the flocculation step. This subject will not be reviewed here, except to say that work on the use of flocculants is advancing due to the recently acquired capability of manufacturing very high molecular weight polymers. If enough flocculant is added, the desired characteristics can be induced. At present, the use of standard chemicals, like iron chloride, would require about .75-1mM per gm biomass, or about 4-6 cents per kg biomass. Thus an alternative to conventional flocculants is required.

One alternative is to condition the algae without adding flocculants. Nitrogen deprivation is generally recognized to enhance the sedimenting characteristics of algae, although data in the literature is not applicable to the situation here. Sedimentation rates were generally not measured and the algae were usually starved for prolonged periods of times, well past any active growth. A report, by one of the authors, on the other hand presents data supporting the hypothesis that normally non-settling, unicellular, flagellated organisms (*Dunaliella*) can be made to settle at rates high enough for collection in settling ponds, by short term nitrogen deprivation [4]. This process was found to work with all but the smallest organisms (2-5 micron) and even with an *Isochrysis sp.* containing 60% lipid. But the question of universality of the process still poses a problem.

Another alternative is to utilize the supersaturated oxygen to float the cells to the surface of the pond, to a foam collector. The advantage of this over sedimentation lies in the fast rate of rise due to the dissolved gas coming out of solution. However, the universality of this approach is, again, open to question.

Preliminary experiments have been performed, at Microbial Products, Inc., using high charge denstiy, high molecular weight polymers to flocculate a test organism in saline medium. A Chlorella sp. was used to demonstrate that even small cells (3-5u) can be clarified by simple sedimentation when first pretreated with the polymer. The media used to grow the alga, to a density of 500 ppm organic dry weight, contained 32 ppt TDS, mostly as NaCl. Alkalinity was 30 mM, pH was 8.45, and hardness was 5mM. The cells were nitrogen sufficient. The dose of polymer necessary to effect complete clarification with an adequate sedimentation rate of the flocs (>50 cm/hr) was 1ppm polymer per 500 ppm biomass. This is the basis for the design assumption used in this report. The experiments were of a very preliminary nature, and not necessarily optimized. However, nor have they been validated on a large scale, having been performed in jar tests with 250 ml of suspension. The universality of the approach has also not been demonstrated, but the type of organism used provided a stringent test. It is the opinion of the authors that such an approach is highly promising, but use on a large scale must be accompanied by continual monitoring of the effectiveness of the flocculation as pond conditions change. The type and mixtures of polymer used may have to be changed on an ongoing basis to avoid selecting for organisms which elude entrapment by a particular flocculant.

The primary method of harvesting the biomass produced in the large scale and experimental systems is specified as pretreatment with polymer followed by sedimentation in a deep pond. The nitrogen depleted condition of the biomass should, in the experience of the authors, promote flocculation and sedimentation. The slurry from this pond is then further concentrated in thickening tanks of much reduced volume.

The budget for the proposed experiment has funds specifically for testing and monitoring polymer induced harvesting.

SECTION 6.0

SUMMARY OF ANALYSIS AND IMPLICATIONS FOR SYSTEM DESIGN

6.1 WATER CHEMISTRY

The characteristics which are desirable in a water resource are low hardness, an excess of alkalinity over hardness, and low TDS (but still too salty for conventional agriculture). Any hardness which cannot be removed by the already present alkalinity (by equilibration with the atmosphere) costs $.005 per kg algae produced per 100 ppm hardness to be removed. Lower TDS results in lower net water usage, although the difference becomes less as the blowdown ratio (= recipricol of salts concentration factor) decreases. Considering a TDS of 40 ppt as the practical upper limit for productive cultivation, a seawater system uses 4.5 times as much water as a system with an input water TDS of 4 ppt, for the same rate of evaporation.

The water chemistry determines the carbonation system specifications and the pH range of operations. The lower limit of pH is determined by outgassing of CO_2. Since CO_2 is the single most costly input to the system, outgassing must be minimized. With low to moderate mixing velocities (15-20cm/s) and proper specification of the low pH limit, overall loss of carbon due to outgassing can be kept to less than 5% of the carbon incorporated into biomass. The upper limit of pH, for properly conditioned waters, is determined by the physiological response of the algae to dissolved CO_2 concentration. This is unknown, but a safe level (one that should not lead to substantial loss of productivity) is 3-6 times air saturation or 30-60uM. The carbon storage capacity of the medium depends on the alkalinity and the pH range of operation. It determines the number of carbonation stations required for a pond of specified size, depth, and liquid velocity, or alternatively, the maximum pond size serviceable with one CO_2 station. The details of interaction between alkalinity, pH, and CO_2 concentration are discussed in Section 2.0.

All of these considerations combine to determine water resource requirements and pH of operation. The combination used for the 1000 acre design are the following: water resource is Type II at 4 ppt TDS with excess alkalinity of 4 meq/l, no water conditioning chemical costs, and eight fold evaporative concentration resulting in a medium with TDS = 32 ppt, alkalinity of 32 meq/l, and a pH range for operation not to exceed 8.0-9.0. Water usage calculates to (assuming an average evaporation rate of .01m/m^2/day) 11.5 liters /m^2/day or 16.5 x 10^6 m^3/400 ha/yr. When this water is obtained from wells of 50 m pumping depth, the electrical energy cost ($.065/KW-hr) is $220,000 or $16/acre-ft.

In the operation of the experimental system, the large scale water specification will be simulated by a mixture of the two types of waters that are plentiful on the proposed site: 5% Salton Sea water, preconditioned by adding sodium carbonate at a cost of $300 per acre per year or $450/acre-ft used and 95% well water available at 1700 gpm. The composition of these water resources is tabulated in Section 10.0.

6.2 CARBONATION

The CO_2 required to produce algal biomass with 50% lipid is 2.2kg/kg biomass. Even at the lowest extrapolated cost of CO_2 [9], the cost of this input comprises over 50% of the annual production cost. Thus the most important aspect of the design is the efficient use of carbon. In this subsection, the methods of transferring CO_2 into the ponds are summarized. Recycle of carbon in particular, and all nutrients in general, is discussed in subsection 6.3 below.

Two types of carbonators were analyzed: covered area carbonators and carbonation sumps. The former type may be feasible when pure CO_2 is used, the latter is feasible for both pure CO_2 and flue gases. The practical efficiency approaches 95% for stripping a pure CO_2 input gas and 80% for a flue gas input.

Gas-liquid mass transfer was analyzed from first principles and evaluated in light of the large body of biochemical engineering data that has been accumulated. Nonetheless, a great uncertainty remains in the transfer rate that is to be expected in large scale algal systems. In the sump carbonation problem, the transfer rates were shown to depend on the average rate of rise of bubble swarms, which depends on the distribution of bubble sizes. This is determined by sparger orifice, gas flow rate, and bubble coalescence factors (salinity, surfactants, turbulence). The rate of removal per unit length of sump depth can be much faster than would be expected if the bubble swarms travel through a considerable per cent of the sump depth before the equilibrium size distribution is attained and/or if the equilibrium bubble size is small due to anti-coalescence factors. Although the empirical data from CO_2 transfer into algal ponds is sparse, and based on small systems, it has been found generally that transfer is faster than an equilibrium analysis would predict in the absence of some of the accelerating factors. The covered area carbonator was analyzed in terms of surface renewal theories of gas-liquid transfer with parameters determined using empirical correlations developed for stream and channel aeration. Again, however, experiments on transfer through the surface of algal ponds, yielded rates significantly faster than were predicted. Thus with either carbonation method, the design specifications cannot be fixed with great precision until results from carefully designed experiments are obtained. At present the following options are specified for the 1000 acre system.

1. "Ripple" covered area carbonator covering 2% of the pond area for pure CO_2, 90-95% efficiency.
2. 1.5 m deep carbonation sump for pure CO_2, 95% efficiency.
3. 2-3 m deep sump for flue gas, with carbonator gas recycle flow of 100-200%, 80% efficiency.
4. 2-3 m deep sump for the mixture of pure CO_2 with digester flue gas, no recycle, 90% efficiency.

Option 2 is used in the Base Case economic analysis, but option 4 has the most potential for reducing lipid production costs.

The experimental system will be used to determine the carbonation parameters. Both covered and sump carbonators will be tested. In addition, the effect of oxygen desorption, during carbonation, on the efficiency of CO_2 input will be determined. These tests will also serve as the basis for the decision as to whether is ts feasible to remove dissolved oxygen from solution, and thereby relieve its potentially inhibitory affect on photosynthesis.

6.3 NUTRIENT RECYCLE

Since carbon is such a significant input into the system, recycle of the carbon left in the residue from the lipid extraction process was considered. It was concluded that anaerobic digestion of these residues in a covered lagoon would allow recycling of 34% of the algal carbon, reducing carbon input from 2.2kg/kg algal biomass produced to 1.45kg/kg. This recycled carbon is considered available at no cost, in terms of operations, due to the electrical energy derived from the methane gas produced which is used to pressurize the recovered CO_2 for injection into the ponds. The capital cost of the digestor-generator system is included in the economic evaluation of this option. The flue gas obtained from the combustion of the digester gases is 14% CO_2. Wnen combined in the required volumetric proportion with purified CO_2, the resultant mixture is 35% CO_2. This will be more expensive than pure CO_2 to inject into the ponds but less expensive than flue gas. The extra expense comes as a capital cost of constructing deeper carbonation sumps, and possibly use of recycle. Since operating costs, especially the cost of carbon, will be shown to dominate annual costs, these increased capital expenditures have little impact on production costs. The 30% savings in CO_2 input has significant impact.

In addition to carbon, much of the nitrogen and phosphorus remaining in the extract residues can be recycled. The potential for recycle is even greater than for carbon. Table 6-1 summarizes the elemental balance for C, N, and P in the anaerobic digestion process. The partition of the elements in the liquid portions of the digester can only be roughly estimated at the present time, but it is evident that the potential exists for these elements to be recovered to a significant extent.

Table 6-1. Nutrient Partitioning in an Anaerobic Lagoon

Digester Compartment	% of Total C	% of Total N	% of Total P
Gas Phase	65	0	0
Liquid Effluent	18	75	50
Sludge	17	25	50

In constructing Table 6-1, it is assumed that 65% of the carbon entering the digester is gasified (39% to methane,26% to CO_2), that the nitrogen is predominantly solubolized to ammonium, and that the phosphorus partitions evenly between the liquid and sludge fractions. The gas and liquid effluent fractions are recyclable. Table 6-2 gives the recycled nutrients as a per cent of the total, using the assumptions that the algae is 60% carbon, lipid is 70% carbon, soluble nutrients are recycled with 100% efficiency and the gaseous carbon is utilized with 90% efficiency due to losses upon injection into the pond.

Table 6-2. Recycle of Nutrients

Nutrient	Kg Recovered/Kg alg.	Kg Lost/Kg alg.	Net Input Kg/kg
Carbon	(.25x.18)+(.25x.65x.9)=.19	(.25x.17+.35)/.9=.44	.44
Nitrogen	.04x.75=.03	.04x.25=.01	.01
Phosphorus	.005x.50=.0025	.005x.50=.0025	.0025

Recycle thus lowers the nutrient demand, per kg algal biomass produced, to 1.6 kg for carbon (from 2.4 kg), to .01 kg for nitrogen (from .04 kg), and to .0025 kg for phosphurus (from .005 kg).

A covered lagoon digester and engine generator is specified for the 1000 acre system. If budget constraints allow, a test lagoon will be constructed for the experimental system, but without combustion of digestor gas, as engine generators are disproportionately expensive at such a small scale. The recycle of N and P will be determined and the use of digester flue gas can be simulated by diluting pure CO_2 with air.

6.4 NUTRIENT LOSSES

The loss of nutrients from the system is a potentially significant problem unless care is taken in the design and operation of the system to prevent such losses. The loss of carbon, primarily as CO_2 to the atmosphere, has been discussed extensively in Sections 2.0 and 3.0. In the Base Case design, to be specified in detail in Section 10.0, the pond depth is 20 cm and the mixing velocity is 20 cm/s. Shallower depth and higher mixing velocity both act to aggravate CO_2 outgassing, which becomes about 2.5 times greater if depth is halved and mixing velocity increased to 30 cm/s. In addition, carbon storage in the medium, on an areal basis, decreases with decreases in depth. For media of the same alkalinity and the same range of dissolved CO_2 concentration, the shallower, faster mixed pond must be operated in a narrower pH range to achieve the same efficiency of use of carbon. This means that more carbonation stations are required per unit area of growth pond.

Ammonia is another volatile species when present in unionized form. In fresh water, the pK of the ammonium-ammonia acid-base system is 9.5. Activity coefficients decrease as ionic of the medium iuncreases. In the mixed acidity constant convention (protons expressed in activity, other species expressed in concentration) the pK for the ionization of ammonium increases. In the approximation that the activity coefficient for ammonium is the same as for bicarbonate, the pK in a medium with the ionic strength of seawater is 9.85. If all of the nitrogen demand of the algae is introduced as ammonia at the beginnning of each batch cycle (a worst case scenario) then the concentration of total ammonia-nitrogen would be about 2.5mM. At pH 8.5 and 9.0, the fraction of this total that is present as ammonia is, respectively .04 and .12. With a mass transfer coefficiet of .06 m/hr for transfer through the surface, the loss of ammonia is, at the two pH's, .03mM/L/hr and .09mM/L/hr. This is low compared to the rate of nitrogen uptake by algae. However, if again the pond depth is halved and the mixing velocity is increased to 30 cm/s, the rate of outgassing is five times as high (the volumetric rate of uptake by algae is also twice as great). So the outgassing losses of carbon at low pH combined with the outgassing losses of ammonia at high pH act to squeeze the pH of operation down, and more severly so with shallower, faster mixed pond designs. In the Base Case, nitrogen losses were specified at 30%, which should be amply conservative even when the outgassing during carbonation and airlift is included.

6.5 MIXING SYSTEM

Three options were analyzed for mixing the ponds: paddlewheels, airlift, and combined carbonation and mixing in sumps. The overall efficiency of the three methods are, respectively 40%,40%, and 20%. The lower efficiency of the combined system is due to the need to optimize for efficient transfer of CO_2. However, the net power consumed in this option is only 35% higher than the other two because the CO_2 cost includes power for compression, and air use is only supplementary. The paddlewheel system is twice as capital intensive as the airlift system. The combined carbonation-mixing system costs almost as much as the paddlwwheel system because the sump used must be deep to insure efficient carbon utilization.

The decision on which system, paddlewheels or airlift, is best suited for mixing algal growth ponds, must await the results of experiments in which an optimal design for each has been tested. They are not actually being compared on an equal basis in this report. The practical efficiency of the paddlewheel system is much better known, i.e., the dynamic losses are more easily factored into the analysis. Thus the 40% efficiency is most likely achievable. The airlift system, on the other hand, was evaluated from first principles, with an ac hoc factor inserted for dynamic losses. No airlift has been designed, to the authors' knowledge, which both fits the application here and is scalable to large systems. The engineering experience with airlifts is based on much higher lifts used in conjunction with difficult to handle materials or in air lift fermenters. The efficiency achievable in a ponding apllication is yet to be determined.

Based on the literature, consultant input, and previous experience, a paddlewheel system thought to be structurally sound and efficient has been designed. Several configurations of airlift, and combined gaslift-carbonators, are presented in Section 4.0. These configurations are merely plausible at this point, since the behavior of bubble swarms in relatively shallow sumps in saline water which may contain surfactants, needs to be empirically determined. Thus the proposed experimental system has been designed for the measurement of the relevant parameters.

A novel type of mixing system, applicable to very large systems, has been analyzed based on the design specifications of a hydraulic consultant. The fan pump has potentially very high efficiency when pushing large flows, at low velocity against relatively low head (a meter or more). Although this alternative has not been included in the 1000 acre design, nor the experimental system, it does open up the possibility of operating much larger ponds than has been considered practical up to now.

6.6 MIXING VELOCITY AND DEPTH

The mixing velocity and depth specified for the 1000 acre design are 20 cm/s and 20 cm. These are considered sufficient to serve the respective purposes of keeping algal solids suspended and providing carbon storage in the medium. Although there is some evidence for enhanced productivity at increased mixing velocity and lower depth when devices are installed in the ponds to induce organized mixing regimes [3], the evidence is not conclusive. This specification increases capital costs, stemming from manufacturing the airfoils used to establish the organized mixing and from the increase in both mixing system costs (due to the higher mixing velocity) and carbonation station requirements (due to the lower areal carbon storage capacity). In addition, the operating costs increase due to the increased power input for mixing, which increases as the velocity to the 2.5-3 power and depth to the negative 1/3 power. CO_2 utilization efficiency also can be expected to decrease as the mass transfer coefficient for outgassing increases with mixing velocity increments and depth decrements (the surface turnover rate increases). The potential impracticality of installing airfoils at reasonable cost makes basing the entire system design on this tenuous.

The experimental system is also designed for 20 cm depth and 20 cm/s mixing velocity, although the system will be constructed with the flexibility to vary both.

6.7 POND LINING

Some form of pond lining is required to eliminate loss of water due to percolation. In the large scale design, the lining specified is crushed rock over a clay sealer. This combination is inexpensive when the materials are not too distant form the site. The major questions to be investigated during the proposed experiment are the degree of sealing that is achievable with this lining and the effect on mixing power input and CO_2 outgassed. Both are expected to be somewhat greater than with a plastic liner due to the increased roughness of the pond bottom. Since the rock is graded and rolled, the anticipated roughness coefficient is .018 compared to very smoothly applied plastic lining of roughness .012.

Plastic liners were considered but not chosen as a primary design specification due to initial cost, installation difficulty, maintenance problems, and questions as to whether linings would perform as well as expected. The least expensive plastic lining would cost between $1.5-2.0 per m^2 installed. However, these liners would need to be constantly under water to avoid photodegradation. In addition, the very small thickness (10 mil) predisposes such a lining to developing small holes which leak. Rodents may also wreak havoc by chewing through the liner. The high quality plastic liners cost $5.5 per m^2. Although performance goals would be much easier to achieve with expensive liners, problems with rodent damage could be serious. The high capital cost of these liners, $23 million per 1000 acres, makes the economics of their use marginal. It doubles the depreciable capital investment, but DCI only contributes 10-20% to the annualized production costs. If a plastic liner were required, few experts recommend the cheaper liner over the expensive one.

High quality plastic liners are specified for all ponds in the proposed experimental system except for the largest ponds. In one of these .4 hectare ponds, clay sealer with a crushed rock overburden will be tested. The other will be plastic lined (hypalon). Thus the performance of both options will be evaluated.

6.8 HARVESTING SYSTEM

Primary harvesting was evaluated in terms of devices for concentrating the algal biomass and in terms of the reliability and universality of each method. Microstraining and air plus DO flotation proved to be more expensive than using a belt filter, which was analyzed as more expensive than simple sedimentation. It was noted, however, that the differences in costs, both capital and operating, would not translate into a large difference in the overall cost of producing algal biomass(see Section 11.0). The most important conclusion is that none of the devices, in and of themselves, could be expected to be very reliable in the longterm. Each discriminates on the basis of size and/or density, which would eventually lead to a situation where organisms that eluded the primary collection would be given a competitive advantage in the growth ponds.

It is necessary to develop a means of pretreating the biomass, with flocculants, so that all biomass is collected by whatever device is used. Recent work with very high molecular weight, highly charged polymers appears to be a promising method of accomplishing this. A given polymer flocculates a wide variety of organisms from a given medium. In addition, a repetoire of similar polymer can be tailored, on site, to deal with changes in pond flora or simply changes in conditions that result in different flocculant needs. The conclusion is that harvesting microalgae appears feasible but will require constant monitoring. There does not appear to be one single answer, in terms of chemical additives, unless pond conditions are extremely constant.

Final concentration of the biomass is accomplished by centrifugation. The cost of this step does depend on the device used in the primary concentration. Flows from microstrainers are expected to be five times those from settling ponds and flotation collectors. A belt filter concentrates, or is expected to concentrate somewhat more than the other methods.

6.9 PRODUCTIVITY ENHANCEMENT AND SPECIES CONTROL

Dissolved oxygen accumulates to high concentration in moderate to large ponds during active growth of algae. Recent experiments [1] as well as a wealth of laboratory and limnological data implicate oxygen as an inhibitor of photosynthesis. The desorption of oxygen is a problem which is not amenable to analytic solution in all but highly specified cases. It is a non-equilibrium, nucleation process which is specific to both the chemical and hydraulic characteristics of each particular situation. Thus the feasibility of intentionally restricting DO levels to within prescribed limits will be tested empirically in the proposed experiment. The results will be evaluated so as to suggest large scale approaches. It will also be necessary to measure biomass productivity as a function of DO in order to determine the need for DO removal and the cost effectiveness of lowering it to a given level.

Another possible productivity enhancement method involves creating conditions of organized mixing in the ponds. Several means of accomplishing this were analyzed, including contouring the pond bottom to include flow obstacles, laying rows of obstacles on the bottom, and using airfoils. None were determined to be attractive in terms of effectiveness of promoting the desired flow patterns and/or in terms of cost. Until evidence to the contrary is firmly established, the assumption used in this report is that productivity is determined by climate, medium, and especially strain cultivated. Certainly the affect of recycling the medium on longterm productivity will be monitored in the experiment.

Longterm species control is related to productivity. Strains used must be highly productive and competitive as well. Up to now, except in extreme conditions of medium composition (which results in low productivity) natural strains have dominated all outdoor ponding systems. One advantage of producing carbohydrates instead of lipids is that all of these dominant organisms observed to date, accumulate the former under nitrogen limiting conditions. A comprehensive survey of strains which accumulate lipid and are highly competitive as well is, however, just underway.

6.10 EFFLUENT DISPOSAL

Of the many technical, resource, and logistic obstacles which must be overcome for the proposed process to be successful, one has not received enough attention. The blow down water represents a significant disposal problem. With the specified blowdown ratio of 1/8 of the inflow, or 1/7 of the evaporative losses, the volume of water per hectare per year is over 5000 m^3. The mass of salts (predominantly NaCl) is over 150 Mt, or about 1.5 Mt salt per Mt dry biomass. This water mass can be evaporated in ponds with an area of .14 hectare per hectare of growth ponds. These ponds need not be lined, since the precipitates will quickly seal them. Even so they must be dredged, forming salt "mountains" in short order. Either this salt must be stored as such indefinitely, or it must be transported to a disposal site, i.e., the nearest gulf or ocean. Thus an operating cost should be included for the transport of this material. The only way to reduce the amount of salt produced, from open systems, is to utilize less saline water resources and/or to decrease the blowdown ratio. Assuming that seepage and percolation losses are not persistent, the lower limit on blow down is set by the secondary concentration. If 10-20% solids is the final concentration before processing, and .005-.1% solids are in the pond effluent the minimum blow down is 2% of evaporation. Unless fresh water is used, this minimum is unrealistic. For practical purposes, if a water resource containing 2 ppt TDS is used and the blowdown ratio is .04, then the volume of water and mass of salt coming out of the system per hectare per year become 1600 m^3 and 80 Mt or .75 Mt/Mt biomass. These are still large numbers. A cost of $.0067 per kg of salt is assumed for the large scale system to transport the salts to a disposal site.

The organic matter that is not either a product stream or recycled back to the growth ponds, must be dredged from the bottom of the anaerobic lagoon periodically. This sludge can be dried and disposed of on site.

SECTION 7.0

POND CONSTRUCTION

This section examines the basis for choosing a particular pond design, and presents specific details relating to its construction. Section 7.1 discusses the various factors which influence growth pond size and geometry. Section 7.2 specifies the particular configuration chosen, and lists the design parameters of importance to pond construction. Section 7.3 presents the specific design options and their costs for each major element of pond construction. The high-rate pond concept, consisting of open, shallow channels in which the culture circulates, is used as the basis of pond design in this report. This chioce is based on both experience (of the authors and others), and on an analysis of alternative designs within the context of fuels production, where the cost constraints are extreme. It should be noted that within the basic high-rate pond framework, there is still ample room for innovation and cost optimization, as the analysis in this section will show.

7.1 POND SIZE AND GEOMETRY

The basic geometric parameters in high-rate pond design are pond size, number of channels, and length to width (L/W) ratio. The latter is defined in this report as the lenth of the center divider wall (i.e. the length of a single channel without bends) divided by the single channel width. In general, a single loop (two channels) is preferred from a hydraulic standpoint. If the pond is large enough, this configuration can take full advantage of a paddle wheel mixer [12]. The choice of pond size and shape is driven by economic factors and the effect on other system elements, such as mixing and carbonation. A simple geometric optimization indicates that a large pond with a low L/W ratio gives the most pond area for the least wall length, as illustrated in Figure 7-1, for the case of twelve ponds in a side-by-side (shared walls) configuration. Since the walls are a significant cost item, and since other pond elements also show economies of scale, ponds need to be made as large as practical. The choice of L/W ratio must take into account the trade-off between minimizing wall length and other factors which affect costs. At low L/W ratios, the channel width increases, as does the cost of those elements related to channel width, such as mixing stations, carbonation sumps, etc. Once a tentative size and L/W ratio is chosen, the head loss (which depends also on velocity and roughness) can be calculated by Manning's formula, and evaluated relative to the proposed mixing system. The channel length may then need to be adjusted to fully utilize the capabilities of the mixing system.

Figure 7-2 shows the cost of the growth ponds as a funtion of pond size for the range of 2-10 hectares. For each point, the number of ponds is adjusted to maintain a total system area of about 200 hectares, so the economies of scale indicated in the figure are a result of geometric factors rather than declining unit costs. (One exception is for laser grading, which has a slight scale economy built into unit cost function, since the cost depends somewhat on the channel width). The figure shows a marked increase in cost below 4 hectares, with diminshing returns above 10 hectares. Both the cost figures and shape of the cost curve are unique to the specific design presented in this section, and may not apply to other designs. A further point with regard to Figure 7-2 is that it includes only the cost of pond construction, mixing, and carbonation, and does not include harvesting, water or carbon supply, engineering, etc.

An alternative to the conventional "racetrack" configuration described above was suggested by Dr. James Harder of the U.C. Berkeley, as a method of generating circular flow patterns (secondary flows), for cycling the culture in and out of the light with a minimum of head loss. A low mound with sloping sides would be constructed using conventional earthmoving equipment. The channel would then be built in a spiral pattern on the mound. This would produce a continuously sloped bottom, as required for shallow, mixed ponds. (Each channel would be later be graded flat across its width if necessary). The inner and outer channels would be joined by an inverted siphon, with the mixing head provided by an large diameter, low speed axial flow pump ("fan pump"), custom designed for the head requirements of the pond. The spiral configuration may have distinct advantages in the construction of very large ponds (>20 hectares), although the secondary flows are minimal at large radii of curvature. It replaces the abrupt channel bends with a continuous gradual bend, eliminating the need for flow deflectors, and providing a more uniform mixing regime. Since its shape is more compact than a long narrow pond, it may better conform to a given terrain and reduce the initial rough grading costs. (However, the fine grading will likely be more difficult than with straight channels). Although the spiral configuration was not employed in the baseline design, it does deserve consideration, especially in the context of very large ponds.

7.2 POND AND SYSTEM CONFIGURATION

For the large-scale system a pond size of 8 hectares (19.8 acres), with a L/W ratio of 20/1 was chosen. Key design parameters are listed in Table 7.1 for the individual pond, and in Table 7.2 for the 192 hectare pond system, consisting of 24 ponds as shown in Figure 7-3. Figure 7-4 shows a plan view of a single pond. The time required to complete one loop around the pond at 20 cm/sec is about 2.6 hours. A single carbonation station will satisfy the carbon storage requirements, given the assumptions about water chemistry outlined in Section 2. Referring back to Figure 7-3, the 8 hectare pond takes full advantage of the economies of scale, without pushing the size into a region of dimishing returns, some of which are not readily quantifiable (e.g. loss of operational flexibility, problems of solids deposition, etc).

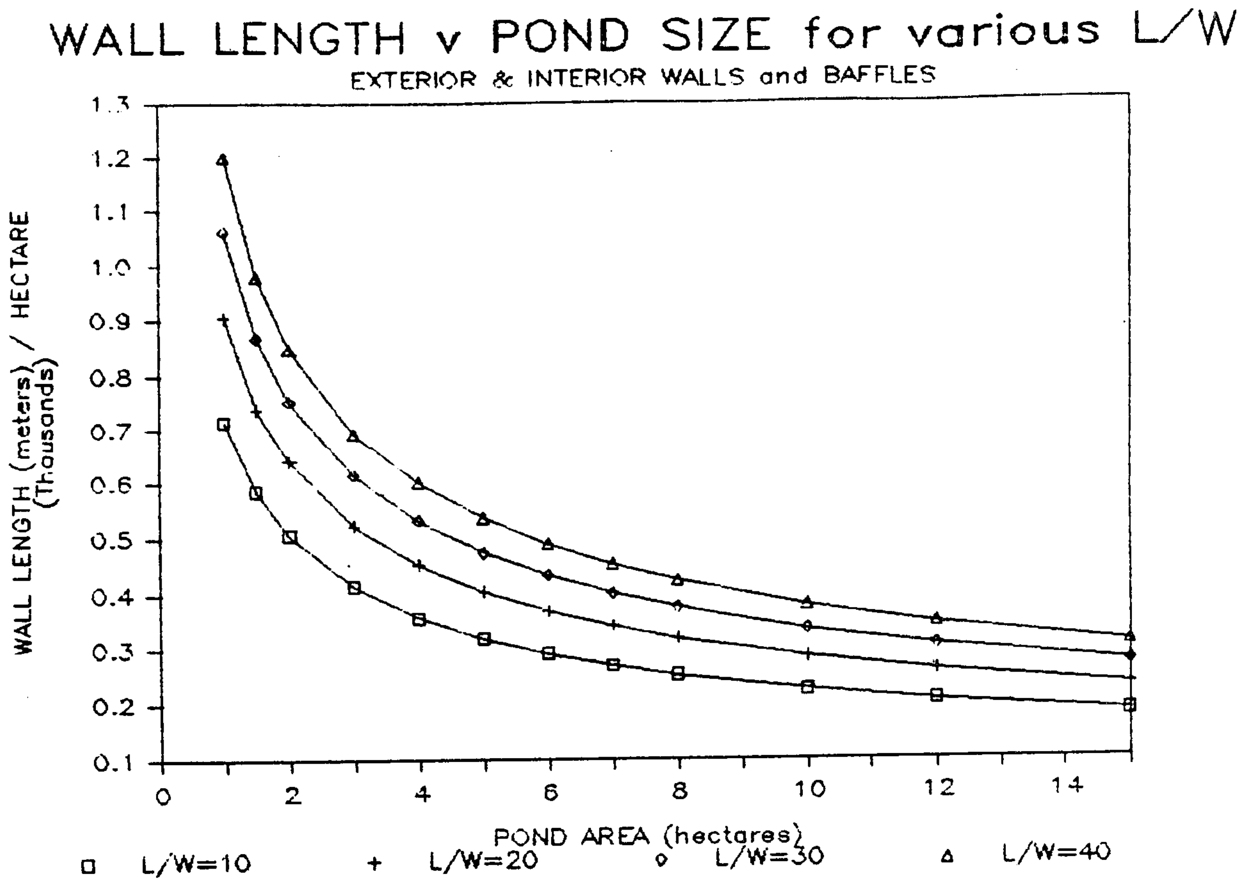


Figure 7-1

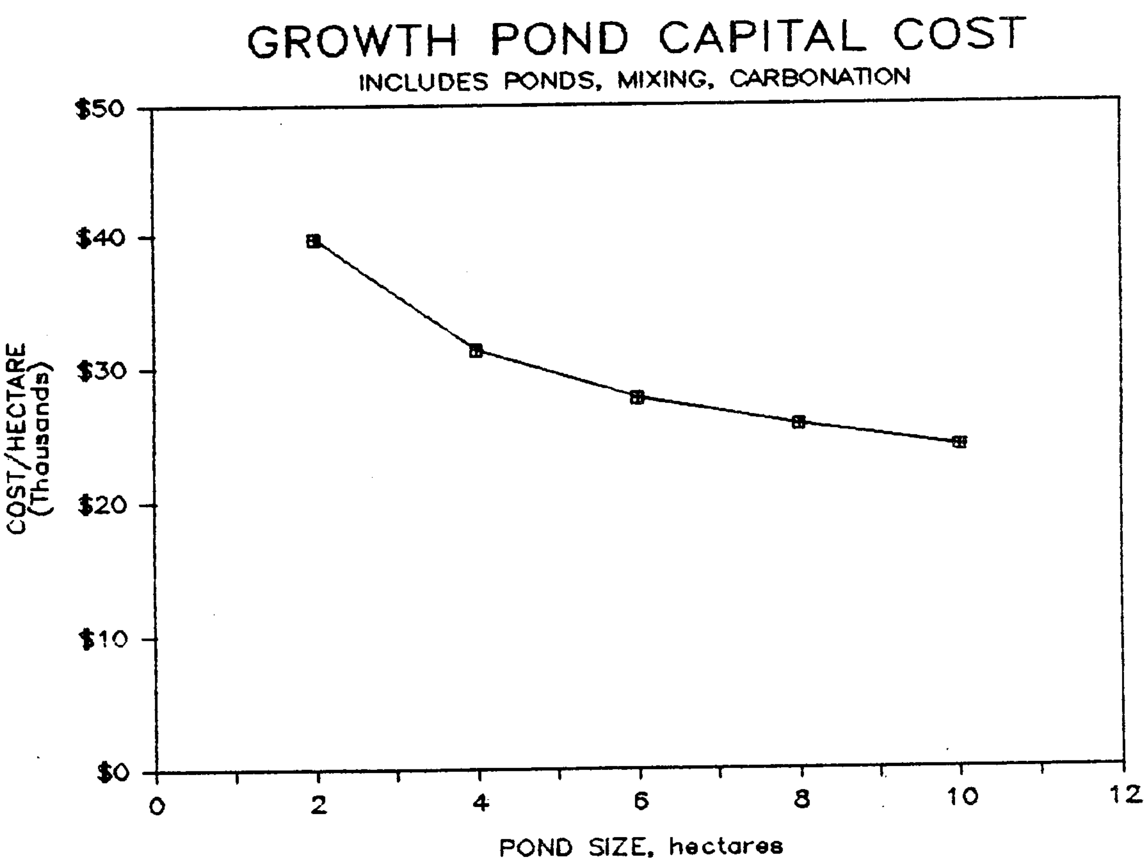


Figure 7-2

Table 7-1 Pond Design Calculations - Single Pond

INPUTS

DESCRIPTION	NAME	VALUE
Pond Area	AREA	8.0 hectares
# of Channels	#CHAN	2 -
L/W Ratio	L/W	20 -
Depth	DEPTH	20 cm
Channel Velocity	VEL	20 cm/sec
Mannings 'n'	MANNING	0.018 sec/m^0.33
Paddle Eff.	PW.EFF	0.6 -
Drive Eff.	DR.EFF	0.7 -
PW Width/Chan Width	PAD/CHAN	0.75 -
Evaporative Rate (max)	EVAP	1.50 cm/day
Blowdown Rate (max)	BLOWDOWN	0.21 cm/day
Detention Time	DET.TIME	4.00 days
Wall Ht. (above grade)	WALL.HT.A	40 cm
Wall Ht. (below grade)	WALL.HT.B	10 cm
Sump Depth	SUMP.DEPTH	1.5 meters

OUTPUTS

DESCRIPTION	NAME	S.I. UNITS	ENGLISH UNITS
Channel Width	CHAN.WID	43.06 meters	141.3 feet
Paddle Width	PAD.WID	32.30 meters	106.0 feet
Centerwall Length	CW.LEN	861.25 meters	2825.6 feet
Single Channel Length	CHAN.LEN	928.89 meters	3047.5 feet
Total Channel Length	TOT.CHAN.LEN	1857.78 meters	6095.0 feet
Total Wall Height	TOT.WALL.HT	0.50 meters	
Slope	SLOPE	1.107 x 10^-4	1.107 x 10^-4
Total Head Loss	HEAD.LOSS	20.98 cm	8.26 inches
Total Efficiency	TOT.EFF	42%	42%
Hydraulic Power	HYD.PWR	3477 watts	4.66 hp
Total Power	TOT.PWR	8278 watts	11.10 hp
Total Unit Power	TOT.UNIT.PWR	0.10 watts/sq m	0.10 w/sq m
Velocity at Paddle	VEL.PAD	26.7 cm/sec	0.87 ft/sec
Pond Volume	VOLUME	16000 cu meters	565035 cu ft
Evap. Flowrate	Q.EVAP	1200 cu m/day	42378 cu ft/day
(alternate units)		833 liters/min	220 gpm
Blowdown Flowrate	Q.BLOWDOWN	168 cu m/day	5933 cu ft/day
(alternate units)		117 liters/min	31 gpm
Harvest Flowrate	Q.HARVEST	4000 cu m/day	141259 cu ft/day
(alternate units)		2778 liters/min	734 gpm

Table 7-2 Pond Design Calculations - Pond System

Uses input from table above, as well as the following:

INPUTS

DESCRIPTION	NAME	VALUE
# of Adjacent Ponds	#ADJ.PONDS	12
Sets of Adj. Ponds	#SETS	2

OUTPUTS All outputs refer to the entire pond system

DESCRIPTION	NAME	S.I. UNITS	ENGLISH UNITS
Total # of Ponds	#PONDS	24	
Total Pond System Area	S.AREA	192 hectares	474.4 acres
(alternate units)		1920000 sq meters	20666304 sq ft
Center Wall Length	S.CW.LEN	20670 meters	67814 feet
Tot.Straight Wall Length	S.STR.W.LEN	43062 meters	141279 feet
Curved Wall Length	S.CUR.W.LEN	6494 meters	21304 feet
Total Wall Length	S.TOT.W.LEN	49556 meters	162583 feet
Total Wall Area	S.TOT.W.AREA	48695 sq meters	524149 sq ft
System Pond Volume	S.VOLUME	384000 cu meters	13560845 cu ft
Evap. Flowrate	S.Q.EVAP	28800 cu m/day	1016928 cu ft/day
(alternate units)		19999 liters/min	5284 gpm
Blowdown Flowrate	S.Q.BLOWDOWN	4032 cu m/day	142370 cu ft/day
(alternate units)		2800 liters/min	740 gpm
Harvest Flowrate	S.Q.HARVEST	96000 cu m/day	3389760 cu ft/day
(alternate units)		66662 liters/min	17612 gpm

Table 7-3 Growth Pond Cost Summary - 192 Hectare System

INSTALLED COST FOR 192 HECTARE SYSTEM

24 PONDS 8 HECTARES EACH

DESCRIPTION	QUAN	UNITS	UNIT $	TOT $	$/HECTARE
GROWTH PONDS					
Earthworks					
Rough Grading	474.4	acre	500	$237,216	$1,236
Lazar Levelling	474.4	acre	1,000	474,432	2,471
Finish Grading*	474.4	acre	2,500	1,186,080	6,178
Sump Excavation	19293	cu yd	2.5	48,232	251
Walls & Structural					
Straight Walls	231761	sq ft	4.0	927,044	4,828
Curved walls	34949	sq ft	5.0	174,744	910
Flow Deflectors	10485	sq ft	5.0	52,424	273
Sump bottom	1956	cu yd	100	195,635	1,019
Sump ends	104	cu yd	200	20,729	108
Rails & Piers	13559	ft	12.0	162,713	847
Solids Removers				61,000	318
Mixing System					
Paddle Wheels	24	# PW	35,790	858,960	4,474
P.W. Structural	24	# ponds	1,500	36,000	188
P.W. Depression	331	cu yd	150	49,577	258
Carbonation System	24	# Ponds	14,640	351,360	1,830
Instrumentation	24	# Ponds	4,000	96,000	500
GROWTH PONDS TOTAL				$4,932,147	$25,688

* Includes crushed rock liner

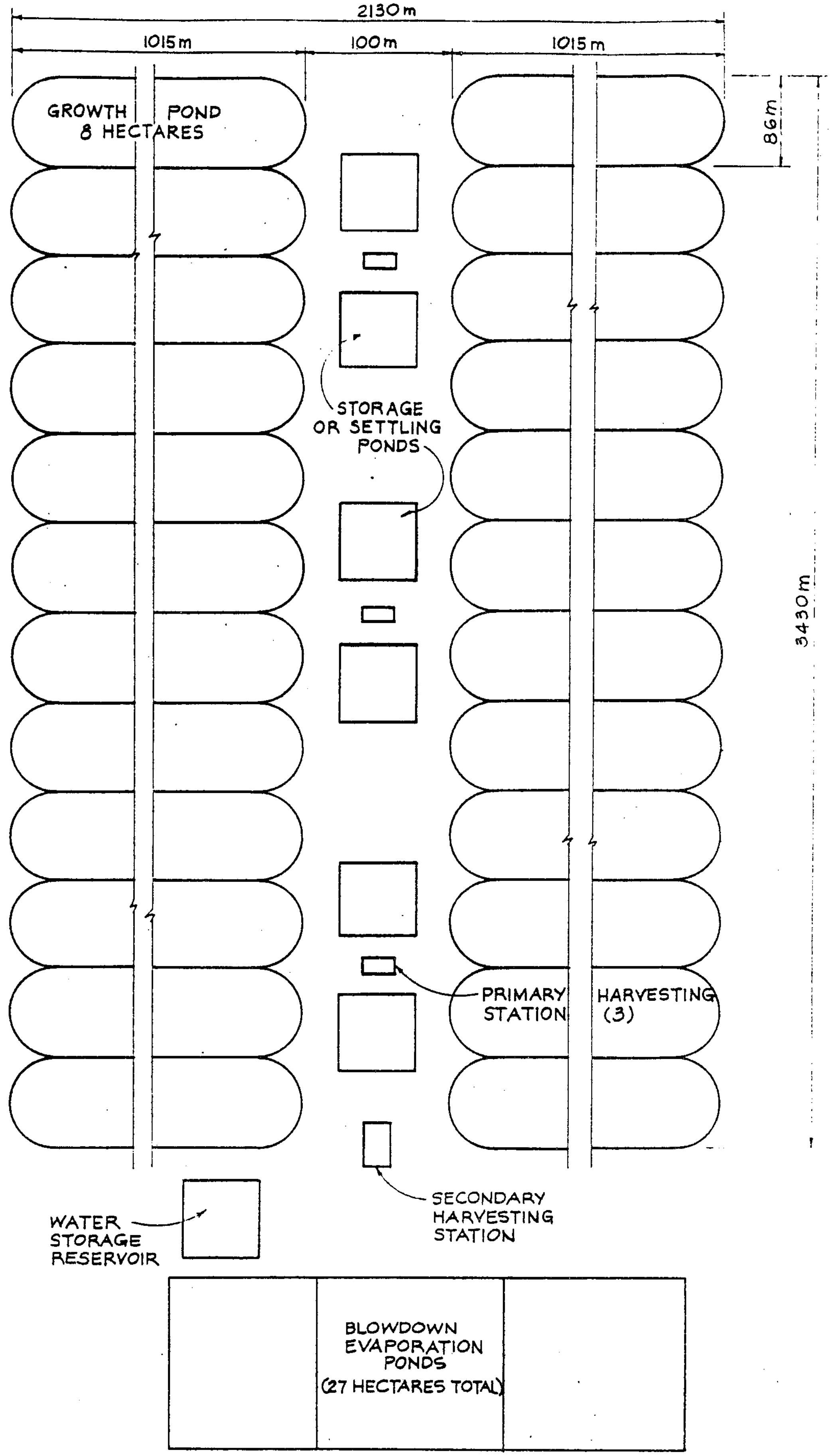


Figure 7-3 Pond System Layout

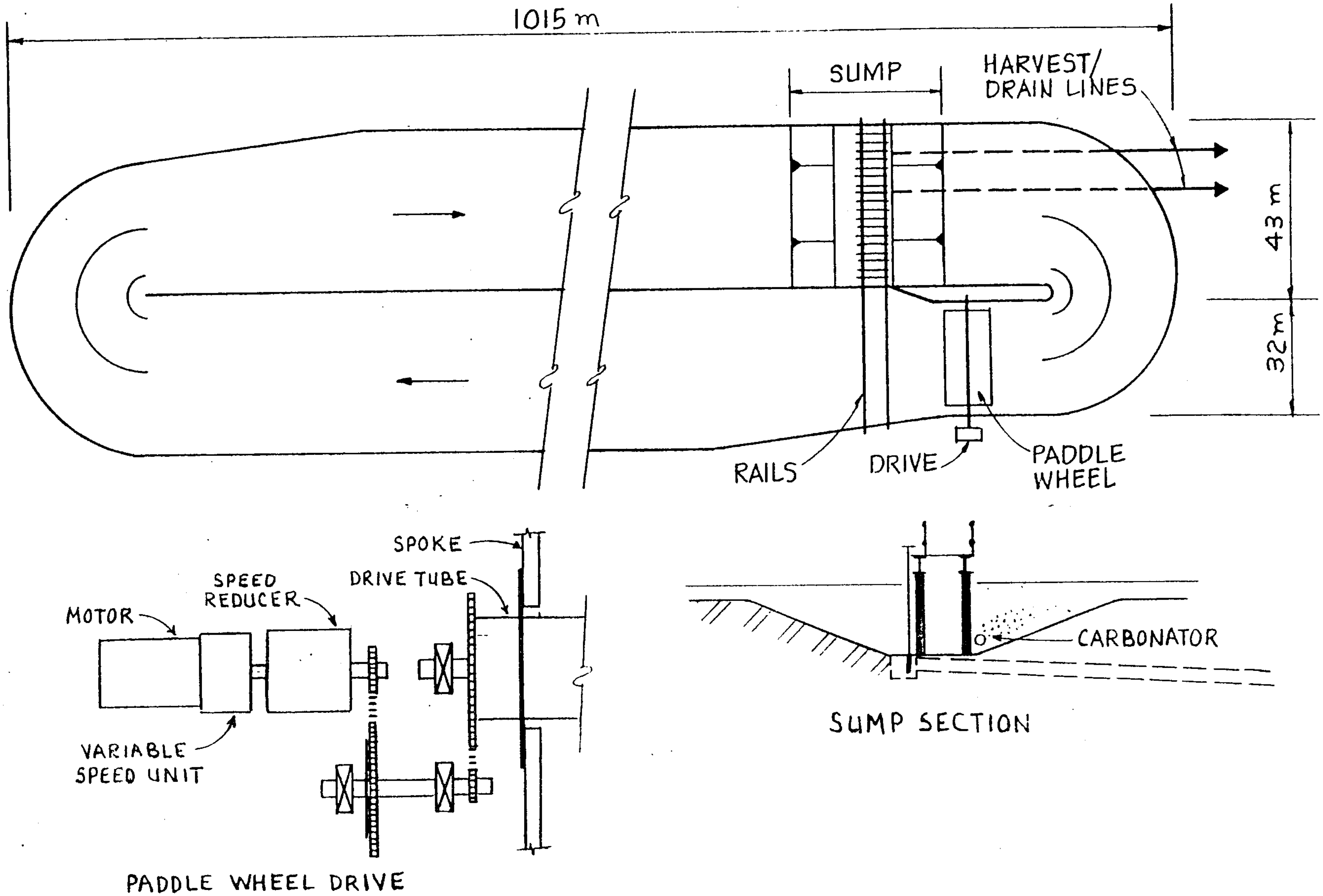


Figure 7-4 Single Pond Layout

7.3 POND DESIGN ALTERNATIVES AND COSTS

7.3.1 Introduction

This section presents the actual designs of the various growth pond elements, along with an estimate of their costs. In some cases, a brief analysis is included, and alternative designs are explored. Sections 7.3.1 through 7.3.8 treat earthworks, walls and dividers, sumps, solids removal, carbonation, and mixing respectively. In section 7.3.9 a summary of costs, based on the chosen alternatives is presented. Harvesting is treated in Section 8.

7.3.2 Earthworks

Rough Grading

As a preparation for laser levelling, the pond grade is established roughly with conventional earthmoving equipment. The amount of rough grading will vary from almost nil on flat terrain to major quantities of cut and fill where the terrain is rough or sloping, requiring terracing to establish the pond grade. Compaction of fills is essential to prevent settlement and can be a significant cost. Presence of rocks, roots, and other material can add appreciably to rough grading costs. Dikes or drains to control surface drainage may also be required. Earthworks for access roads, storage reservoirs, etc., is not included. For favorable terrain, rough grading costs can vary from 0 to $4000/hectare. A value of $1200/hectare is used in the cost estimate.

Laser Levelling

Laser levelling is needed to achieve the tolerances required for shallow pond operation, and to establish the very flat channel slopes to meet hydraulic requirements. It has become a widely used technique in agriculture, where the cost can be as low as $900/hectare for large fields. For pond construction, the cost will be greater, since the slope must be set separately for each channel, and tolerances will be quite rigorous. A value of $2500/hectare is used in the cost estimate.

Finished Grading and Crushed Rock Liner

The cost constraints of fuel production may prohibit the use of synthetic pond liners, which alone would cost $38,000-58,000/hectare depending on the quality of liner used. In the cost estimate, it is assumed that the ponds will be built on tight soil, which is sufficiently impermeable to control seepage. Seepage rates generally decline with time as a result of clogging by organic matter and inorganic precipitates. In some cases, a clay liner (15-30 cm) may be necessary. Often, the clay is available on-site, as in the case of the experimental facility described in Section 12. If not, finished grading costs could be substantially higher, if long haul distances were involved. However, even a doubling or tripling of finished grading costs would have a minor affect on overall pond construction costs. Finish grading is usually required on clay soils due to loss of grade from construction activites such as trenching, etc. The soil will be overlain with a minimum thickness blanket (3-5 cm) of crushed rock or other coarse granular material to control suspension of fines. The particle size distribution and degree of compaction of this rock overburden will determine the channel roughness. Transport costs for imported granular material can be appreciable if local material is unavailable. The total cost for finish grading and crushed rock liner will likely cost between $4000-8,000/hectare. A design value of $6200/hectare is used in the cost estimate.

7.3.3 Walls and Dividers

A variety of wall construction techniques are conceivable, ranging from earthen berms to slipform concrete. This section will examine five options, cite their pros and cons, and give cost estimates for each. Four of these options are illustrated in Figure 7-5. In general, walls need to be about 0.4 meters high, except in areas where wave action would require more freeboard (e.g. downstream of paddle wheels), or where water accumulates under quiescent conditions. Very large ponds would probably require higher walls since larger wind induced waves could occur. In general, the center wall which divides the two channels can be of identical construction as the exterior walls. In small ponds with low head losses it may be possible to use a less expensive method for the dividers (e.g. membrane dividers).

Earthen Berms

Earth berms must be constructed of compacted fill by over-building, trimmed to give smooth side slopes, and then protected against erosion. A minimum top width before trimming is about 2 meters, for conventional compaction equipment. Slope protection material would have to be several times the thickness of the bottom blanket, because of more difficult construction and greater erosion potential. Even so, there is likely to be high maintanence costs associated with slope protection. Extra protection, e.g. gunnite, will be required where wave action tends to concentrate. Weed growth on the top and sides of earth berms is also a potentially costly maitanence problem. More land is required as compared to narrow walls. Although it may be possible to construct berms for as little as $16/linear meter with special equipment, the O&M disadvantages weigh heavily against earth berms. Earth berms may be appropriate for those walls which form the perimeter of the pond system, since an access road will be needed around the perimeter, and drainage considerations may require it to be elevated.

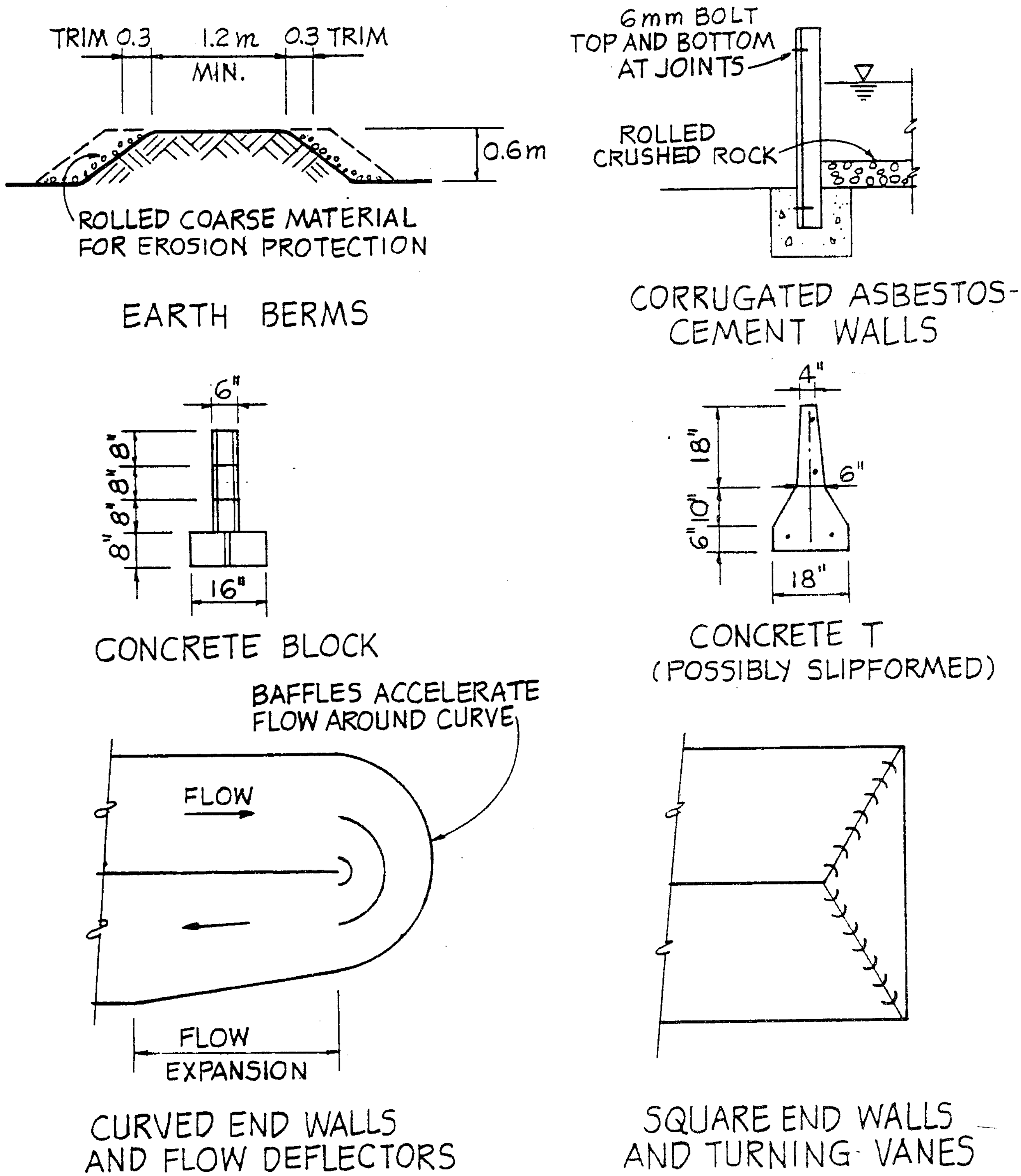


FIGURE 7–5 POND WALL AND FLOW DEFLECTOR OPTIONS

Corrugated Walls

A novel technique, used by Dodd in Singapore, employs corrugated asbestos-cement roofing sheets embedded in a shallow concrete-filled trench. The sheets are cut and drilled off-site, then installed with the corrugations running vertically. Adjacent pieces are caulked and bolted together, forming a free standing, waterproof wall which can expand and contract with changes in temperature. Material costs are about $9/meter, with total installed costs estimated at about $24.50/meter. O&M costs are likely to me quite low.

Membrane Walls

The membrane wall consists of a Hypalon membrane supported on a framework of 2"x2" redwood posts spaced 1.2 meters (4 feet) on center. The posts are set in 15 cm (6 inch) diameter holes filled with concrete for anchoring and stability. The top of the membrane is anchored between two railings again made of 2x2 redwood. The bottom of the membrane is anchored in a 30 cm (12 inch) deep trench backfilled with earth which also prevents seepage under the wall. This estimated cost of this method is $18.07/meter ($5.51/ft, see Appendix III). Although this makes it one of the lowest cost options, it is also the least durable, so that O&M costs would probably offset initial cost savings. However, experimental ponds are often membrane-lined, so this method, or some variation, would be appropriate. The experimental ponds at Vacaville, CA used solid wood walls and a membrane liner.

Concrete Block

The concrete block wall consists of a poured concrete footing 15 cm x 30 cm (6" x 12") with 2 courses of 4"x8"x16" blocks set on top. The block cells would be filled with pea gravel concrete and a single #4 reinforcing bar set in the top block to give the wall adequate longitudinal strength. The footing would be set below grade for stability and to prevent seepage under the wall. The estimated cost of this option is $23.35/m ($7.12/ft).

Poured Concrete

The poured concrete divider wall is set on a 15 cm x 30 cm footing, is 10 cm (4") thick and 40 cm (16") high. Custom made reusuable forms are proposed that are free-standing to reduce installation and relocation labor. One #4 reinforcing bar would be set along the upper part of the wall to hold the wall together and minimize cracking. Cracking would be controlled by joints every 6.1 meters (20 ft) and sealed with silicon to minimize seepage through cracks in the control joints. The cost of this option is about $17.50/meter ($5.34/ft). Given the large quantity of walls to be constructed for the large scale system, it should be possible to employ slip-forming techniques at about the same cost. The key concern with concrete walls is seepage. Shrinkage cracks will definitely occur, but there was a lack of agreement among consultants as to their effect. One felt that the control joints plus the tendency for cracks to clog would mitigate the problem, while another felt that the need to make control joints waterproof would greatly increase the cost of concrete construction. The author's experience with experimental (1000 m^2) high rate ponds tends to support the former, but the problem is certain to be more pronounced in large ponds. Although most seepage will be contained within the pond system by virtue of the side-by-side layout, excessive seepage would result in short circuiting within the pond. If seepage is not a serious problem, then the concrete walls are the preferred method of construction, possibly in combination with corrugated walls for the curved portions. Such a combination was used in the cost estimate. The difference in cost between the two alternatives ($24.50/m for the corrugated vs $17.50/m for the concrete) is not enough to have a major impact on the system capital cost.

Flow Deflectors

The curved flow deflectors aid in minimizing the extent of eddy formation and the associated stagnant zone downstream of the bend. Again, these can normally be built in the same manner as the straight walls, although some of the wall options outlined above are more easily adapted to curves (e.g. corrugated walls). The situation is improved if the baffles and end walls are placed eccentrically, and the channel width decreased through the bend, as shown in Figure 7-5. The acceleration of flow is sufficient to reduce eddy formation and eliminate most of the solids deposition. The channel constriction technique is also used at the other end of the pond, which allows access to the paddle wheel [12]. An alternative to curved deflectors are turning vanes, also shown in Figure 7-5. Turning vanes are widely used in analogous fluid systems, (e.g. air ducts), although Dainippon Ink & Chemical, Inc. apparently holds an application patent for their use in algae production ponds. If closely spaced, (i.e. at intervals equal to the pond depth), the vanes are quite efficient, and virtually eliminate problems of eddy formation and solids deposition. They also allow the ends of the pond to be square, which simplifies construction. The quantity of material required for both curved deflectors and vanes is about the same, but the latter require more elabarate support and would be more expensive to install. Curved deflectors were used in the design, priced at $25/meter ($273/hectare). Vanes would cost about twice this amount.

7.3.4 Sumps

Sumps are necessary in order to provide a deepened area for CO_2 additions so that high absorbtion is acheived, and to provide a collection point for draining the pond. The sump may also provide an area of reduced velocity where inert solids and settleable organic matter can accumulate and be removed. The analysis of CO_2 absorption presented in Section 3.0 concluded that for an for an 8 hectare pond, a sump depth of about 1.5 meters will result in 95% absorption of CO_2, if the diffusers produce bubbles about 1 mm in diameter. This depth is also sufficient to prevent vortexing and air entrainment in the drain pipe. The sump cross-section is shown in Figure 7-4. In order to facilitate drainage, insure uniform carbonation, and expedite excavation, a single sump is constructed across the entire width of the channel. The transition into the sump is gradual to minimize turbulence and avoid the need for concrete formwork. The length (in the direction of flow) is related to the issue of solids removal, as discussed in the following section. In the baseline design, the length is kept to a minimum (1 meter). The distribution pipe for CO_2 is located at the downstream end of the sump, so as not to interfere with any solids deposition that may occur. CO_2 absorption actually takes place in the downstream transition section. The harvesting/drain piping originate from small "sub-sumps" located in the bottom of the sump, so that a submergence of 1.5 meters is maintained.

7.3.5 Solids Removal

The accumulation of both inert and organic solids, which by virtue of their density or tendancy to agglomerate into flocs and settle, is inevitable after a period of pond operation. This material tends to settle in "dead spots" (i.e. stagnant areas) within the pond, and will eventually lead to operational problems. Since a sump is to be included for the reasons outlined above, the question arises whether it could also serve as a sedimentation basin for removal of settleable solids. If, for instance, the sump were made 5 meters long (bottom length), the detention time of fluid passing through the sump would be about 300 seconds. A geometric analysis indicates that only very rapidly settling particles ($V_p \geq 30$ cm/min) will be removed completely, with the remainder being removed in proportion to their settling rate divided by 30 cm/min. Since most of the solids settle at rates well below this, the efficiency of the sump for solids removal is likely to be quite low. Although the efficiency could be improved by increasing the length of the sump (V_p varies as 1/L), the marginal increase in performance would not justify the expense. The problem is that both low velocities and long detention times are necessary. Thus it may be necessary to design stagnant zones, similar to those created by an unbaffled turn, in which solids can accumulate and be removed, but in such a way so as not to disrupt the mixing regime over a large area of the pond. It may be possible to do this in a portion of the carbonation sump, or perhaps in a smaller sump right after the bend before the paddle wheel. The whole issue of solids accumulation needs to be assessed at the pilot scale, since most of the current information comes from observations of waste treatment ponds, which are subjected to high organic loadings, and in which a large accumulation of solids would be expected. A method for removing solids from the main sump, which uses a suction device mounted on a travelling bridge, has been developed [12], and is described in Appendix III. For an 8 hectare pond, this method would cost about $1300/hectare, most of which is for the piers and rails on which the bridge travels. Regardless of whether the sump will accumulate sufficient solids to justify the expense of the travelling bridge, the \$1300/hectare has been included in the cost estimate, since it will allow access to the sump for servicing the carbonation system and effluent drain valves. Also described in Appendix III is a mechanical method for dredging the pond channels, which is estimated to cost \$310/hectare. The final decision on the best approach to solids removal must await pilot scale studies which will define the severity of the problem and the effectiveness of the variuos alternatives.

7.3.6 Carbonation

The analysis in Section 3 concluded that a single carbonation station per pond would be sufficient to maintain an adequate level of dissolved carbon in the culture. The carbonation system is designed to supply the peak demand of 7.2 gm $CO_2/m^2/hr$, which corresponds to about 4800 liters/min (170 cfm) at the assumed delivery temperature of 27°C (80°F). A 100 mm (4") distribution pipe spans the channel on the downstream side of the sump, with spargers spaced along its length. The effect of sump depth and bubble size on CO_2 transfer efficiency was also described in Section 3. The analysis concluded that very fine bubbles (1 mm) were necessary for lateral flow carbonation in shallow sumps, and that the sump depth increased rapidly for larger bubbles. While it is a fairly simple matter to generate bubbles of 2-3 mm diameter with porous diffusers, to produce 1 mm bubbles the flowrate per diffuser must be quite low (bubble size is more sensitive to flowrate than to pore size). The manufacturers contacted could provide only cautious estimates of flowrates for this application, so that the actual type and number of diffusers will have to be determined by pilot scale tests. For purposes of cost estimation, a relatively conservative approach was taken, based on information supplied by Cardox Co., a manufacturer of porous tube diffusers. The total cost of \$18,800 (\$2350/hectare) includes the pH controller, control valve, CO_2 flowmeter, and associated piping. Since experience has shown that the rates of CO_2 absorption are often better than those predicted by the methods of Section 3, significant cost reductions are possible in this category.

7.3.7 Mixing

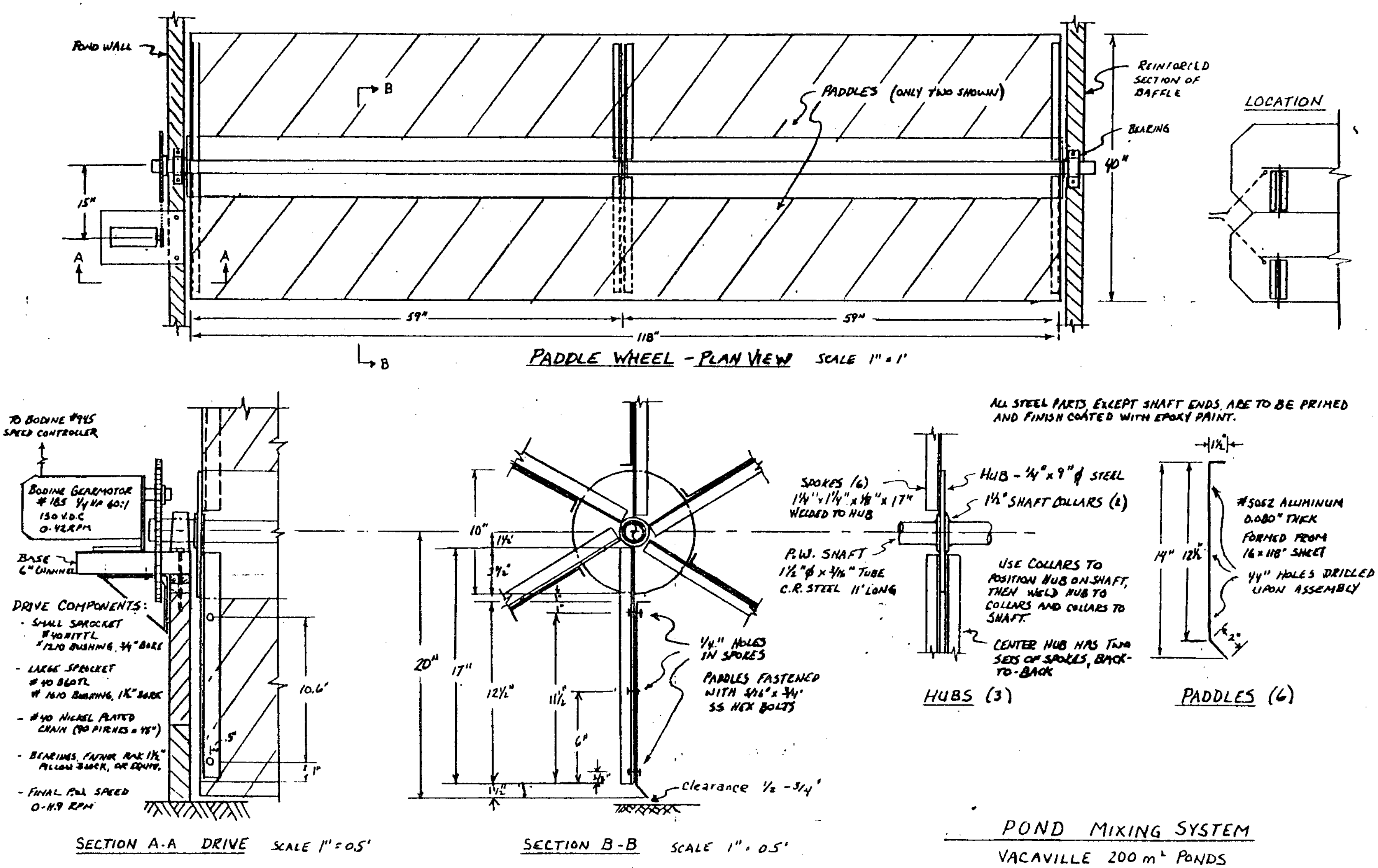
For an 8 hectare pond mixed at 20 cm/sec, the Manning's formula predicts a head loss of 21 cm (8.3"), for an assumed roughness factor of n=0.018. The corresponding hydraulic power is 3480 watts (4.67 hp). Paddle wheels are chosen for mixing, for reasons outlined in section 4-2. The constriction in channel width (see Section 7-3-3) allows the paddle wheel to be 25% smaller than the channel width, and allows access to the drive. Even so, the width is 32.3 meters (106 ft), so the paddle wheel is broken into three sections, separated by flexible couplings that allow some vertical and angular misallignment. Design of the paddle wheels is based on a stress and deflection analysis of the paddles, the drive tube, and of each paddle wheel section. Although the low rotational speed means that torques are very high, it is the deflection of the paddle wheel under its own weight (and that of attached growth, which can be significant) that governs design of the tube. The design is similar to the small paddle wheels used in the Vacaville, CA experimental ponds shown in Figure 7-6, with some notable changes. Many of these changes are based on improvements made by Dodd in the course of designing several "generations" of paddle wheels. The number of blades is increased from six to eight, and adjacent sections are offset by 22.5°, both of which reduce pulsations on the drive train and in the flow of the water. A solid baffle between sections also blocks flow across the channel. A small clearance between the paddle tip and the shallow (5cm) curved depression below it makes the paddle wheel in effect a large positive displacement pump, with a minimum of backflow. The drive shaft is a steel pipe with solid stainless steel stub shafts at the bearings. The drive consists of a 15 hp electric motor, a variable speed unit, an in-line speed reducer, and a two-stage chain and sprocket reduction. The final driven sprocket is inboard of the bearing, so that the stub shaft merely supports the paddle wheel weight, and does not transmit the drive torque. The various components (and their cost) are listed in Appendix III. The drive tube and hubs are epoxy coated steel, while the spokes and paddles are fiberglass, for corrosion resistance. In order to have the necessary rigidity in bending, the paddles must either be supported with a structural member (if flat), or be formed into a structural shape themselves. The large quantity of paddles required for the large scale system allows the use of custom designed shapes, at very little cost over that of flat sheets. Thus a curved shape paddle was designed that has sufficient rigidity to span the 2.43 meters (8 ft) between hubs without structural backing. The paddle is 0.64 cm thick, 57 cm high, and has a 61 cm radius of curvature. (The paddle height must accomodate the nominal depth, lift, depression, and wave action generated by the paddle wheel itself). The curved shape also sheds water efficiently as it leaves the water. With each pair of ponds laid out as a mirror image, the two drives are close together, minimizing wiring costs.

The paddle wheels are a major cost component of the growth pond system, each one costing about $37,500, including structural supports, concrete depression, and installation. This is almost 20% of the growth pond cost.

7.3.8 Instrumentation

Instrumentation to measure temperature and dissolved oxygen will be provided for each pond, and this information, along with pH, CO_2 flowrate, and make-up water flowrate will be transmitted to a microcomputer-based data acquisition unit which records the data and prepares summary reports. Given the relatively low price of such systems, one will be installed at each harvesting station, (i.e. for each eight growth ponds), eliminating the long data transmission lines that would be necessary in a totally centralized system. Allowing $2000/pond for the measuring and transmitting equipment,

Figure 7-6 Paddle Wheel Design (typical)



and $16,000 per harvesting station for the data acquisition system (installed costs), the overall cost is $96,000, or $500/hectare. A system of this type could later be expanded to automate harvesting operations, an option not provided for in the cost estimates.

7.3.9 Pond Cost Summary

A summary of costs for the 24 eight hectare ponds is given in Table 7-3. The table was developed on a spreadsheet, so that the sensitivity to changes in design or costs could be examined with relative ease. The quantity of material in each category is an equation based on the Pond Design Calculations, samples of which were listed in Tables 7-1 & 7-2. In most cases the unit costs were fixed, although in some cases (e.g. carbonation, mixing), they were represented by an equation, or reference to a look-up table. For this reason, the range of the program is limited to ponds of about 2-10 hectares, although the quantities column is accurate over a much wider range. The mixing system costs were generated from a separate program.

SECTION 8.0

HARVESTING SYSTEM DESIGN

This section presents three alternative designs for harvesting, inluding (1) Microstraining; (2) Belt Filtration; (3) Settling. (Air/DO floatation, mentioned in Sections 5 and 12, is not analyzed in this section). In each of the three cases, further concentration is acheived through centrifugation, in order to bring the solids content of the slurry up to 10% VS. However, the costs of centrifugation vary widely according to the level of concentration achieved in the primary harvesting step. The harvesting operations are based on the analysis of Section 6, which concluded that an induction phase would be necessary to achieve high lipid levels in the cells. This in turn implies batch, as opposed to continuous, harvesting. The harvest flowrates are based on a four day (system average) detention time,although the growth/induction cycle will most likely be longer, especially in the winter months.

8.1 MICROSTRAINING

Microstrainers are an attractive harvesting method because of their mechanical simplicity and availability in large unit sizes. The recent availability of very fine mesh polyester screens has revived interest in their use for microalgae harvesting. Fabrics as fine as one micron (nominal) have been successfully employed for algae removal from lagoon effluents. However, attempts to harvest small unicellular algae with the one micron fabric have yeilded disappointing results, in spite of the fact that the cells were larger than one micron in size [13]. (The lack of success in straining small algae with the one micron weave can be attributed to the fact that the actual openings in this fabric are rectangular, with dimensions of approximately one by six microns). Subsequent studies concluded that it would be necessary to flocculate the cells prior to microstraining. Current research is aimed at determining the types and amounts of flocculant necessary for various species. The use of flocculants in harvesting can have a significant impact on operating costs, depending on the dose required. Under certain conditions, it may be possible to use microstrainers without adding flocculants, as in the case of large unicellular (e.g. Platymonas) or colonial algae, or with culturing techniques that induce "autoflocculation."

The design assumptions for the microstrainer based harvesting system are listed in Table 8-1, and are discussed below. The critical design parameters are hydraulic loading (flowrate/net effective submerged area), solids removal efficiency ((influent TSS-effluent TSS)/influent TSS), and concentration ratio (concentrate TSS/influent TSS). Although accurate prediction of these parameters requires pilot scale tests with the particular species, some reasonable estimates can be made from existing information. The hydraulic loading determines the size and number of microstrainers required, and so has

TABLE 8-1. Microstrainer-Based Harvesting System

Design Parameters

Single growth pond volume	16,000 m^3
System pond volume	384,000 m^3
Detention Time[1]	4 days
System harvest flow (alternate units)[2]	96,000 m^3/day 66,700 liters/min
Microstrainer influent density	750 mg/l
Microstrainer hydraulic loading	325 lpm/m^2
Microstrainer size[3]	3.7 m dia X 4.9 m long
Net effective submerged area/unit	22.6 m^2
Flow per unit	7345 liters/min
Number of units required	9
Concentration Ratio	10:1
Centrifuge influent density	7500 mg/l
Flowrate to centrifuge	6700 liters/min
Centrifuge design loading	15000-3800 liters/min
# of units required	2
Concentration ratio	13:1
Centrifuge effluent density	100 g/l (10% Total solids)

[1]See text

[2]24 hr/day basis

[3]Envirex Model

a strong impact on capital costs. Its value depends primarily upon the influent solids concentration, the size of the screening media, and the characteristic size of the algae. The "one micron" screen imposes too severe a penalty on hydraulic loading to be a practical alternative. Since this analysis presupposes large, colonial, or flocculated algae, the required mesh size will fall in the range of six to twenty microns. The hydraulic loading is assumed to be 325 lpm/m^2 (8 gpm/ft^2), a reasonable value for the mesh range specified. The importance of solids removal efficiency depends on the mode of operation, ranging from critical for "once-through" systems (e.g. seawater), to moderate for recycle systems, wherein unharvested cells are returned to the ponds in the harvest return stream. On the other hand, the microstrainer must produce a significant increase in solids density to justify its relatively high cost. For lagoon effluent polishing, a concentration ratio of 6:1 is typical. However, these systems are operated to produce the lowest possible effluent density, not to concentrate algae. When optimized for the latter, a tenfold concentration should be attainable. This corresponds to a tenfold reduction in flow to the secondary harvester. Since centrifuge capacity is largly flowrate dependent, the concentration ratio is of major importance.

In order to reduce large diameter piping runs, the primary harvesting is decentralized into three harvesting stations, each containing three microstrainers, and each serving eight growth ponds. An alternative would be to use slightly smaller microstrainers (increasing the number to twelve) and building six stations, each serving four ponds. This would further reduce piping runs, but the savings would be offset by higher microstrainer and facilities costs. This trade-off is not considered in the cost analysis. A schematic of the harvesting system is shown in Figure 8-1. An entire pond volume is harvested on the fifth day of the growth/induction cycle, so the system average detention time is actually five rather than four days, and harvesting could be spread over 1.2 days per pond. However, a one day per pond timetable is more practical and allows some flexibility in O&M.

An unfortunate consequence of batch harvesting is the need to store the clarified effluent until harvesting is completed. In a continuous process, it is returned immediately to the pond, but in a batch process this would dilute the culture along an exponential curve. Since the microstrainers are likely to be limited by solids, rather than hydraulic loadings, it may be possible to return the effluent, while gradually increasing the harvest flowrate to compensate. Obviously there are practical limits to this, as flowrates would have to increase indefinitely in order to harvest all of the cells, so in fact some algae would always go unharvested. With careful controls, the correct amount of seed algae could be left in the pond for the next batch. However, given the uncertainties involved, it is prudent to include storage ponds in the design, which hold the clarified effluent until harvesting is complete. If located entirely above grade, the effluents can be returned to the growth pond by gravity. If harvesting begins after dark on the fourth day and is finished 24 hrs later, the entire night is available to drain the effluents from the storage pond back into the growth pond.

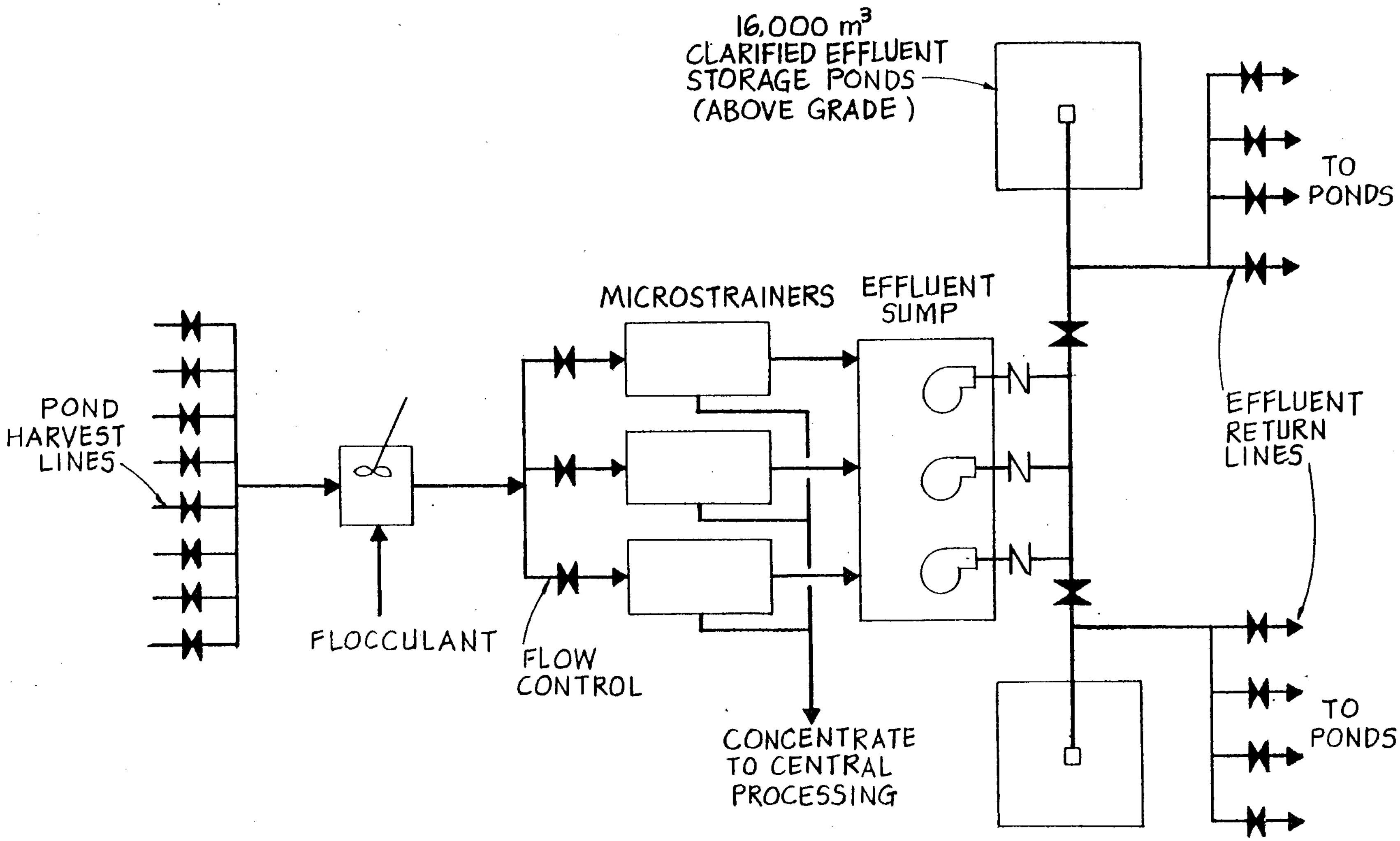


Figure 8-1 Microstrainer-Based Harvesting System

The concentrated harvest stream from all microstrainers is pumped to a centralized secondary harvesting facility. The total flowrate to this facility, assuming 24 hr/day operation, is 6670 l/min (1760 gpm). Three large horizontal solid bowl decanter centrifuges, constructed with corrosion resistant materials, are required. These are rated at 1510-3960 lpm (400-1000 gpm) each, depending on feed characteristics, flocculant dosage, and degree of concentration required. Although the use of disk nozzle centrifuges might reduce capital costs, the savings would most likely be offset by higher operating costs [12]. The centrifuges concentrate the solids by a factor of 10-13, producing the required 10% slurry. The solids are discharged continuously by means of an internal screw conveyor. The clarified effluent from the centrifuges is returned to the ponds via the makeup water network, eliminating the need for separate return piping.

Capital costs for the microstrainer-centrifuge harvesting option are shown in Table 8-2. The total harvesting cost is $21,800/hectare. If the centrifuges could be operated at 3700 lpm (very close to their maximum rated capacity), then only two units are required, and the total harvesting costs are reduced to $18,700/hectare. Estimated annual operating costs are shown in Table 8-6, for an assumed flocculant cost of $.01/kg dry algae. The sensitivity of operating costs to flocculant dose is discussed in Section 11. Although one would intuitively expect a trade-off between the quantity of flocculant used and the microstrainer design parameters (which would effect capital costs), there is not sufficient data at this time to evaluate this trade-off, other than to say that certain species simply cannot be harvested without flocculants.

8.2 BELT FILTRATION

A vaccuum belt filter for algae harvesting, developed by Dodd in 1972, used a fine mesh stainless steel screen with a paper precoat. The concept was later refined, with the sreen and precoat being replaced by a polyester fabric similar to those used in microstrainers. A pilot scale unit operated in Singapore was very successful in harvesting Micractinium and other large species, somewhat less so with small algae such as Chlorella. In general, the floccuant dose should be less than in the case of microstrainers, since the algae mat aids in filtration. Although the belt filter has a relatively high capital cost, it can concentrate solids much more effectively than a microstrainer, by a factor of about 70:1 as opposed to 10:1. Since the product is more concentrated, flow to the secondary harvester is reduced substantially, resulting in lower secondary harvesting cost and reduced energy consumption. For the entire 192 hectare system, the secondary harvester flow is about 950 lpm (250 gpm). The layout of a belt filter system would be very similar to the microstrainer system shown in Figure 8-1. The capital cost of the belt filter option is shown in Table 8-3, based on estimates supplied by Dodd. (The belt filter was assumed to cost 50% more than a microstrainer of equal capacity). Storage ponds are again necessary.

TABLE 8-2. MICROSTRAINER BASED HARVESTING OPTION

CAPITAL COSTS

PRIMARY HARVESTING		
9	Microstrainers (3.7 m dia x 4.9 m) stainless steel construction, includes motor & controls @ $160,000 ea.	$1,440,000
3	Housings, with concrete pit & sumps @ $50,000	150,000
3	Pumps stations, including all valves & Piping	300,000
6	Storage ponds	300,000
TOTAL		$2,190,000
TOTAL/hectare		$ 11,400
SECONDARY HARVESTING		
2	Sharples PM95000 centrifuges, complete with motor & control @ $650,000	$1,300,000
1	Housing & concrete pad (4000 ft)	100,000
1	Pumps & piping	150,000
TOTAL		$1,550,000
TOTAL/hectare		$ 8,070
TOTAL HARVESTING		$3,740,000
TOTAL/HECTARE		$ 19,480

TABLE 8-3. BELT FILTER BASED HARVESTING OPTION

CAPITAL COSTS

PRIMARY HARVESTING	
Belt Filters (1.5 x Microstrainer Costs)	$2,160,000
3 Housing, with concrete pad & sumps	150,000
6 Effluent Storage Ponds	300,000
3 Pump Stations, including a-1 harvest/return piping, valves, etc.	300,000
Total	$2,910,000
Total/hectare	$ 15,200
SECONDARY HARVESTING	
2 Sharples PM60000 Centrifuges @ $275,000	$ 550,000
1 Housing + concrete pad (2000 ft^2)	50,000
1 Pump/piping system	50,000
Total	$ 650,000
Total/hectare	$ 3,390
TOTAL HARVESTING	$3,560,000
TOTAL/HECTARE	$ 18,600

The centrifuge cost is based on two units, each smaller than those used with the microstrainers. The choice of two smaller, rather than a single larger unit reflects the O&M realities of the need for more than a single centrifuge. Since the solids concentration going into the centrifuge is so much higher than in the case of the microstrainers, a more conservative design flowrate was used. The overall cost is $17,800/hectare. Harvesting cost is not nearly as sensitive to centrifuge flowrates as in the cases of microstrainers. Operating cost estimates for the belt filter option are shown in Table 8-6. Note the significant reduction in energy cost associated with this option.

A comparison of Tables 8-2 and 8-3 shows that the two options presented thus far have capital costs very nearly the same, if the favorable assumption regarding centrifuge flowrate is made for the microstrainer option. Operating costs for the belt filter option are significantly less. However, any comparison should be made with some caution, since large microstrainers are commercially available and their costs are known, whereas the belt filter represents a new technology, for which costs are necessarily more uncertain.

8.3 SETTLING

The use of settling for microalgae harvesting is based on the observation that nitrogen-starved cultures will settle at rates of 30 cm/hr or greater without adding flocculants. (The addition of small amounts of polymer can increase this rate and produce a more compact concentrate). The same phenomenon has been observed in non-nitrogen starved cultures, but in a less consistant, and more species and media dependent fashion.

A settling rate of 30 cm/hr is still relatively low, and requires long detention times in the settling device, whether it be batch or continuous. The latter would not apply in this case, given that the growth/induction process is inherently batch. The major engineering problem in a large scale batch settling process is to devise an effective, inexpensive solids removal scheme. The major drawback of batch settling is the large piping systems that are needed to transfer the fluids in relatively short periods of time, so that more time is available for quescient settling. This is partially overcome by the use of shared settling ponds and pump stations, which service a number of ponds in sucession. Other questions raised in connection with large scale settling ponds include the effect of wind mixing on settling rates and the possible degredation of the lipid fraction during the process. The wind mixing problem can be mitigated through the use of berms. Degredation is not likely to be a problem in the time period (approximately 24 hours) involved in the proposed design.

The layout of the settling pond system is shown in Figure 8.2. A two stage process is proposed, consisting of a large (0.5 hectare) settling pond, which concentrates the algae by a factor of 20, followed by a second gravity thickening stage, which concentrates by an additional factor of 2.5, for an

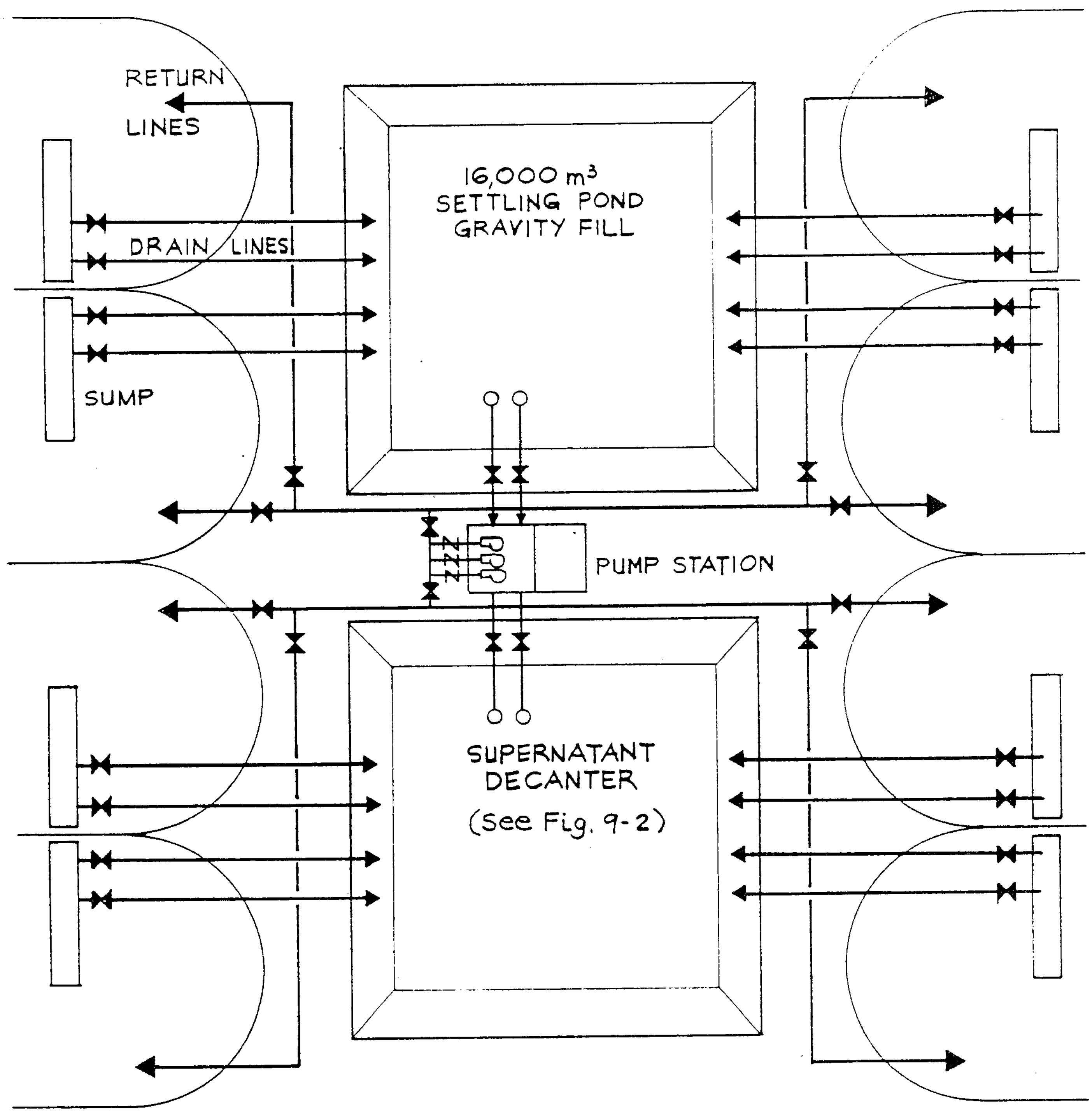


Figure 8-2 Settling Pond System Layout

overall concentration factor of 50. The choice of a two stage process is based on experience in operating the 40,000 liter batch settling pond in Vacaville, CA. This approach greatly reduces the need for precise discrimination between the concentrated bottom fraction and the less concentrated layer that forms above it, albiet at a greater cost. The large settling pond is sized to hold one pond volume (16,000 meter^3) in a 3 meter depth. The working volume is all below grade, so the ponds can be drained by gravity. It is a conventional sloped-sidewall basin, with 2:1 sideslopes. The dimensions are shown in Figure 8.3, along with other relevent details. The pond is lined with Hypalon[1] or CPE[2]. The pond bottom slopes towards a central alley on a 2% grade. Both supernatant and concentrate piping go to an external pumping station that is shared by two settling ponds. The following summarizes the harvest sequence: (1) The contents of the growth pond are drained to the settling pond. The drain lines are sized to complete this in four hours. (2) The contents settle for a period of eight hours. (3) The supernatant is decanted off over a period of ten hours, and is returned to the growth pond (see Figure 8-3 for draw-off details). Since liquid is removed from the top of the pond, settling continues in the lower layers. When the liquid level reaches the break in the slope, decanting is complete. Approximately 5% of the contents remain in the settling pond. (4) The remaining concentrate is pumped to the secondary thickener. At various intervals, the concentrate stream (or stored secondary supernatant) is diverted into the flushing system, washing down solids that have have settled on the bottom of the pond into the collection alley, which leads to the concentrate sump. The concentrate removal process takes about two hours. The settling pond is now available for the next growth pond. (5) The concentrate remains in the thickener for about eight hours. The secondary supernatant is returned to the growth ponds, while the secondary concentrate is pumped to a centralized facility for centrifugation to 10% solids, as in the other harvesting schemes.

The sequence as described above is based on a four day cycle time, which is the shortest conceivable period for growth and induction. Actual cycle times will probably be five days or longer, depending on the time of the year, so that settling and decanting times may actually be longer than described above. The flowrates of the various fractions are listed in Table 8-4, for the four day cycle. Additional harvesting details are given below.

The pump station is a wet pit design, and is divided into two sections for supernatant and concentrate. Three vertical mixed flow pumps, each sized at 50% of the required flowrate, are provided. The pump station operates

[1] A synthetic rubber liner manufactured by DuPont
[2] Chlorinated polyethylene liner

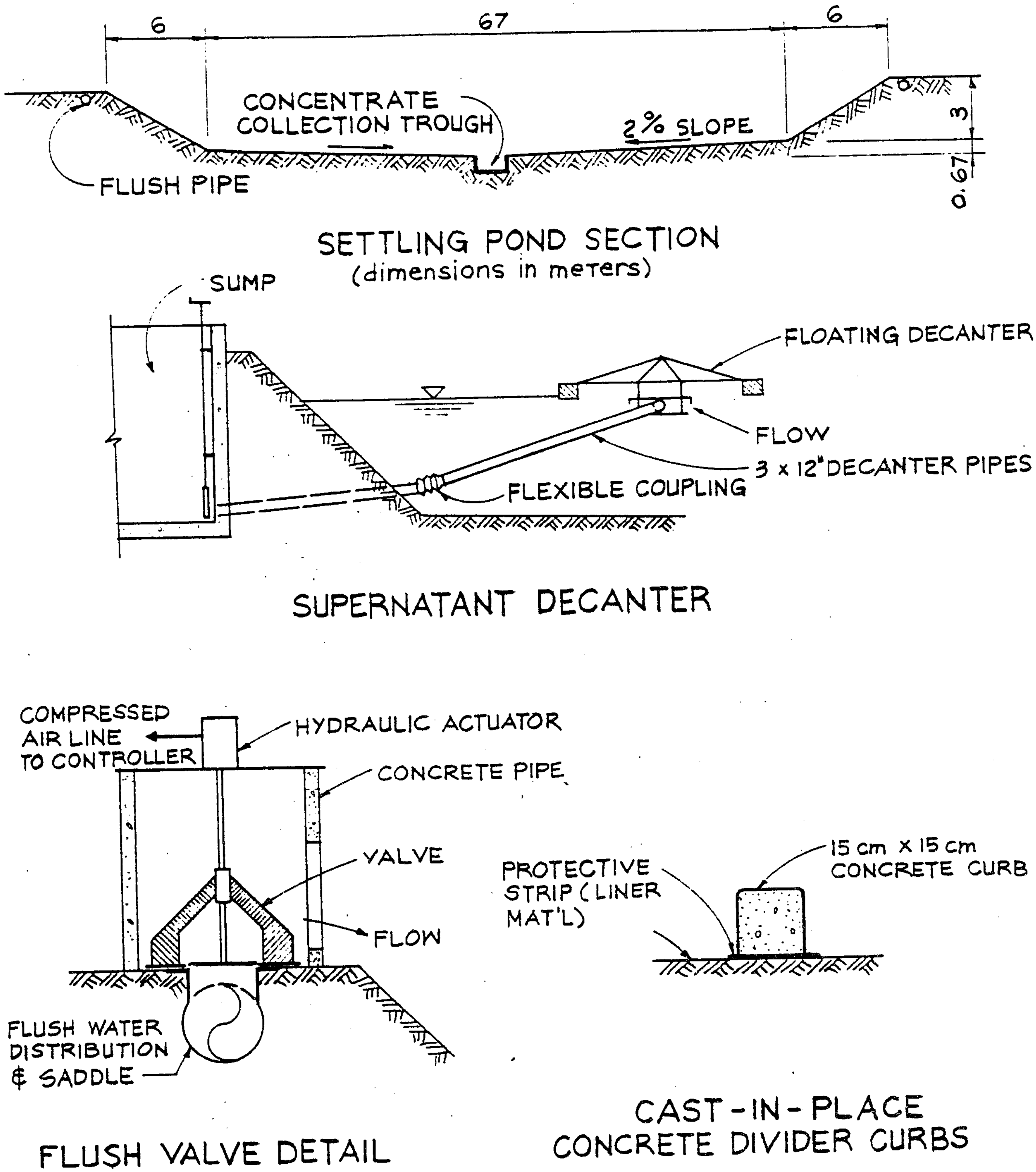


Figure 8-3 Settling Pond Details

almost continuously when the cycle time is four days. Two screw pumps are used to pump the primary concentrate into the secondary thickener. (Low shear solids handling pumps are required for handling all concentrate fractions). The flush system is very similar to those used in dairies for manure collection, and is designed to provide flushing velocities of about 1 meters/sec (3 ft/sec), which should be sufficient to resuspend and transport solids accumulated on the bottom of the pond. The channels are divided by 0.10 meter (4 inch) concrete curbs which are either poured directly on the liner or precast and set in place. In the figure, the concentrate itself is diverted into the flush network, consisting of a distribution pipe which sequentially feeds the flush water into 16 channels, 8 on either side of a concrete collection alley (see details in Figure 8-4). It may be necessary to divert some of the initial concentrate (which will have the least amount of solids) into a small storage pond, and use this liquid for flushing, or build two smaller secondary thickeners and flush with recyled secondary supernatant. Since good distribution of the flush water over the width of the channel is essential, optimization of the geometry may also be necessary, (i.e. a larger number of channels, or a change in pond geometry to longer, narrow channels). None of these options would have a major impact on costs. These and other issues, such as the optimal duration and frequency of flushing, must be resolved in pilot studies.

After settling in the secondary thickener for about 8 hours, the secondary supernatant is decanted in the same fashion as before, and pumped into the return flow network. Secondary concentrate is pumped to the central facility for centrifugation. The average flowrate is 40% greater than in

Table 8-4 Settling Pond System Flowrates

SINGLE POND:

Fraction	Volume m^3	Volume gallons	Time hrs.	Flowrate lpm	Flowrate gpm
Pond Volume	16,000	$4.23(10^6)$	4	66700	17612
1^o Supernatant	15,200	$4.02(10^6)$	10	25300	6690
1^o Concentrate	800	$2.11(10^5)$	1.2*	11400	3000
2^o Supernatant	480	$1.27(10^5)$	2	4000	1057
2^o Concentrate	320	$8.45(10^4)$	2	2670	705

SYSTEM (centrifuge flowrate):

2^o Concentrate	$320\ m^3/4$ days x 24 ponds =	1330	352

* Total removal time is longer because of recyle to flush system.

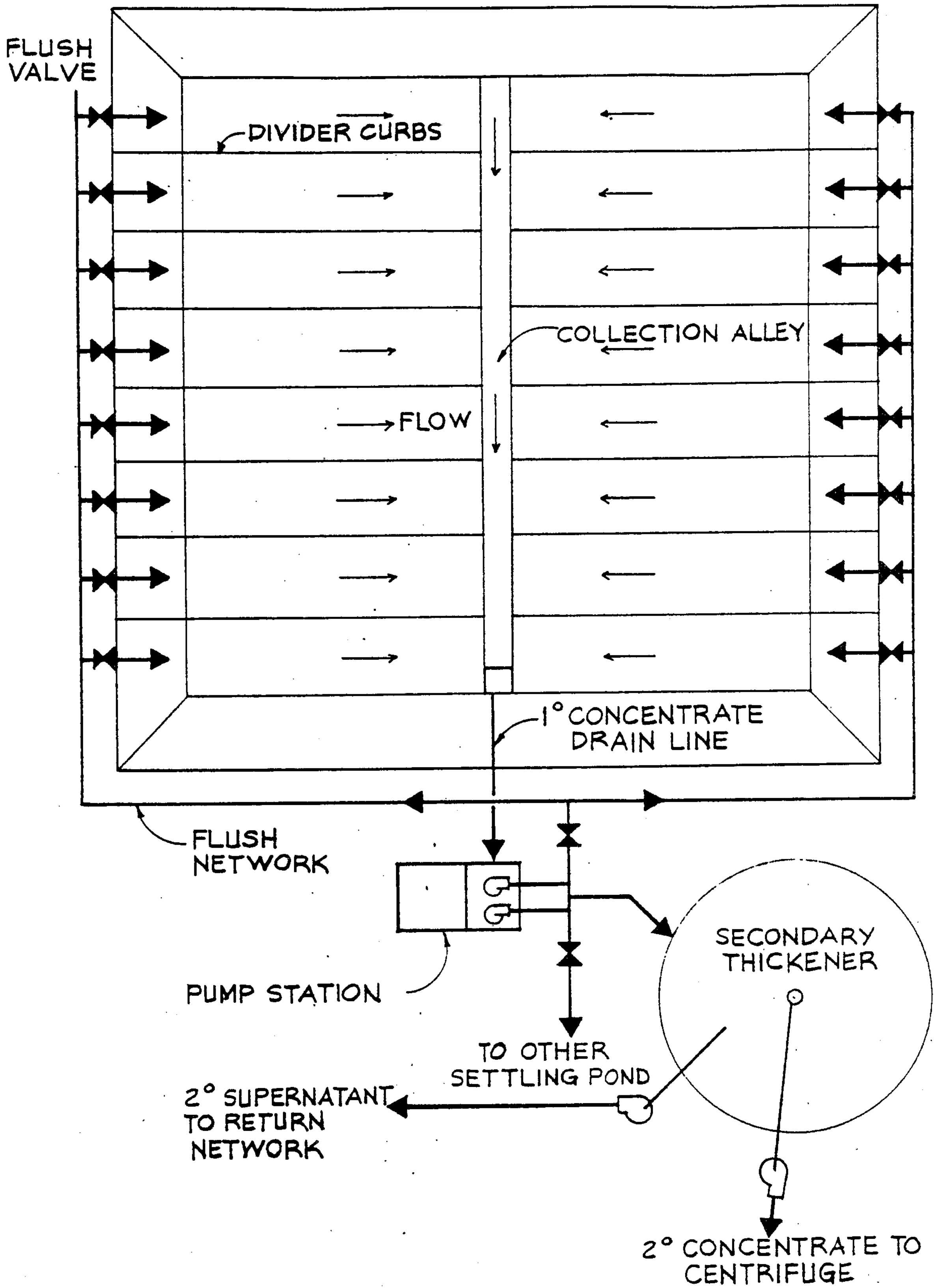


Figure 8-4 Settling Pond Flush System and Secondary Thickening

the case of the belt filter, but a factor of five less than in the case of microstrainers. Capital costs for the settling option are shown in Table 8-5. At $11,260/hectare, cost of this option is substantially less than for either microstrainers or belt filters. Estimated operating costs are shown in Table 8-6.

Table 8.5 Settling Pond Harvesting Option

Capital Costs

PRIMARY HARVESTING (2 stage)

Each harvesting station consists of 2 settling ponds (below grade), one pump station, and 1 secondary thickener. Each station serves 8 growth ponds.

	Station	System
Excavation & shaping: 25,000 yd^3 @ $2.50/$yd^3$ x 2 =	$125,000	$375,000
Liners: 71630 ft^2 @ $0.60/$ft^2$ x 2 =	86,000	258,000
Piping, valves, sumps, pumps (see Table AIV-1)	156,000	468,000
Secondary thickening (see Table AIV-2)	107,000	321,000
Primary Harvesting Subtotal	474,000	1,422,000
Primary Harvesting Subtotal/hectare		$7,360
SECONDARY HARVESTING:		
2 Sharples PM75000 centrifuges @$330,000		660,000
Housing & concrete pad		50,000
Pumps & piping		50,000
Secondary Harvesting Subtotal		750,000
Secondary Harvesting Subtotal/hectare		$3,900
TOTAL HARVESTING		$2,172,000
TOTAL HARVESTING/HECTARE		$11,300

Table 8-6 Estimated Operating Costs of Harvest Options

	MICROSTRAINER	BELT FILTER	SETTLING PONDS
ELECTRICAL			
Primary, kwh/yr	174,000	348,000	343,000
Secondary, kwh/yr	2,578,000	967,000	1,096,000
Total, kwh/yr	2,752,000	1,315,000	1,439,000
Total, $/yr	$178,900	$85,500	$93,500
FLOCCULANTS	$216,000	$216,000	$216,000
MAINTANENCE	$112,200	$106,800	$81,000
TOTAL	$507,100	$407,500	$390,500

Notes:
Costs are for 192 hectare system
Electrical cost based on 6.5c/kwh
Flocculant cost based on 1.0c/kg dry algae
Maintanence cost @ 3% of capital cost includes materials and outside labor only, not operating labor.

Section 9

SYSTEM-WIDE COSTS

System-wide costs include the water, carbon, nutrient, power, and blowdown systems, as well as roads and buildings. In the case of carbon and power, the emphasis is on the distribution rather than on the source, since SERI has has examined the resource issues in far greater depth than would be possible in this report. An exception is the issue of nutrient recycle within the system, which is examined in Section 6. Table 9-1 lists the system wide costs, and also summarizes the total cost for the 192 hectare system. Table 9-2 lists the electrical and maintanence costs for the system. Both tables are based on the use of settling ponds for harvesting. Total operating costs are presented in Section 10.

9.1 WATER SUPPLY AND DISTRIBUTION

The capacity of the water supply system is based on a maximum evaporative rate of 1.5 cm/day and blowdown rate of 0.2 cm/day. Maximum, rather than the average rates must be used because such rates are likely to persist for a period of months during the summer. Additional water is required for initial filling of ponds, but this is assumed to take place during the non-summer months, when excess capacity is available. The required volumes and flowrates were listed in Table 7-2. On a continuous basis, the combined flowrate is 22,700 lpm (6000 gpm). Both the supply and distribution system will be sized on this basis, although pumping costs are based on yearly averages of 15,200 lpm (4000 gpm), which reflects both reduced evaparation and some contribution from rainfall. The authors believe that the SERI estimates for water consumption were too low, while estimates for the cost/unit delivered were high. This analysis assumes a separate well field for each 192 hectare module, which differs from the approach taken in the SERI resource studies.

The well field is assumed to be off-site, to avoid O&M interference and possible subsidence problems. Eight 30 cm (12") wells, each rated at 2840 lpm (750 gpm), are set in a grid pattern with a spacing of 1600 meters (0.5 mile). The wells are assumed to be 122 meters (400 ft) deep, with a static lift of 46 meters (150 ft). A water storage reservoir (16,000 m^3) is included as part of the supply system. If necessary, water conditioning facilities would be located at the reservoir (see Section 2).

A cost estimate of the water supply system is presented in Table 9-3. The cost of wells is based on information supplied by well drillers in the Salton Sea area. The cost of the piping network is based on the assumption that inexpensive "100 ft head" PVC is used. While heavier, more expensive pipe is always used in municipal water systems, the 100 ft head pipe is widely used in low pressure agricultural applications, especially when large diameters are required. All large diameter pipes specified throughout this report are of this type. The unit costs in all cases includes installation, which runs

Table 9-1 192 Hectare System Capital Cost Summary

Base Case: 112 mt algae/ha/yr

	TOTAL $	$/HECTARE
GROWTH PONDS		
Earthworks*	$1,945,960	$10,135
Walls & Structural	1,594,290	8,304
Mixing System	944,537	4,919
Carbonation System	351,360	1,830
Instrumentation not included elsewhere	96,000	500
HARVESTING - Settling Pond Option		
Primary**	1,436,000	7,479
Secondary (centrifugation)	760,000	3,958
SYSTEM-WIDE COSTS		
Water Supply	660,000	3,438
Water Distribution	189,000	984
CO2 Distribution	50,000	260
Nutrient Supply System	150,000	781
Blowdown Disposal System*	160,000	833
Buildings (harv. blds. not included)	110,000	573
Roads & drainage*	100,000	521
Electrical Distribution (3% of above)	256,414	1,335
Electrical Supply	113,000	589
Machinery	80,000	417
ENGINEERING (10% of above)*	899,656	4,686
CONTINGENCY (15% of above)*	1,484,433	7,731
LAND COSTS ($1250/hectare)*	480,000	2,500
TOTAL CAPITAL COST	$11,860,650	$61,774
DEPRECIABLE PORTION	$6,415,601	$33,415
NON DEPRECIABLE PORTION	$5,445,049	$28,360

Notes:

* non-depreciable item
** patrially non-depr. ($375,000)
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table 9-2 Electrical & Maintanence Operating Cost Summary - 192 Hectare System

	Electrical* kwh/yr	Electrical* $/yr	Maintanence* % of Cap	Maintanence* $/yr
GROWTH PONDS				
Earthworks			1%	$19,460
Walls & Structural			3%	47,829
Mixing System	2060000	133,900	3%	28,336
Carbonation System			3%	10,541
			3%	2,880
HARVESTING - Settling Pond Option				
Primary	343000	22,295	3%	43,080
Secondary (centrifugation)	1096000	71,240	5%	38,000
SYSTEM-WIDE COSTS				
Water Supply	1555000	101,075	3%	19,800
Water Distribution	121000	7,865	3%	5,670
CO2 Distribution			3%	1,500
Nutrient Supply System	104000	6,760	3%	4,500
Blowdown Disposal System			3%	4,800
Buildings (harv. blds. not	200000	13,000	3%	3,300
Roads & drainage			1%	1,000
Electrical Distribution			1%	2,564
Electrical Supply			1%	1,130
Machinery			10%	8,000
TOTAL	5479000	$356,135		$242,389
TOTAL/HECTARE		$1,855		$1,262

* Assumes electricity costs $0.065 /kw-hr
** Maintanence materials only

about $10-26/meter ($3-8/ft), depending on size.

The water distribution system includes eveything beyond the storage reservoir and water conditioning system. Three vertical propeller pumps feed into a piping network that supplies makeup water to the growth ponds. The system schematic is shown in Figure 9-1, along with pipe sizes, flowrates, and pressure losses. The flows are for evaporative makeup only, the blowdown make-up is added when the pond is refilled after harvesting. In a system using settling ponds, the central corridor is fairly wide, (about 110 meters), so it is more practical two use two distribution mains which run down either side of the central corridor. A pump station is located at the storage reservoir. Pressure requirements are assumed to equal the dynamic (mainly pipe friction) losses, since some static head is available at the reservoir. The water distribution system could conceivably operate by gravity flow, but this assumption was not made.

Table 9-3 Water Supply System Cost

Pipe Size inches	Total Distance meters	feet	Installed Cost $/ft	Total Cost $
10" dia.	4024	13200	$6.00	$79,200
15" dia.	805	2640	9.00	23,760
20" dia.	805	2640	15.00	36,600
24" dia.	500	1640	21.00	34,440
TOTAL PIPING				$177,000
VALVES & FITTINGS				30,000
DRILLING & CASINGS 8 x 400 ft @ $50/ft				160,000
PUMPS, Submersible S.S. 8 x $18,000				144,000
Electrical equipment				24,000
Installation				22,000
HYDROLOGICAL STUDIES & TESTING				50,000
STORAGE RESERVOIR				60,000
TOTAL				$667,000
TOTAL/HECTARE				$3,470

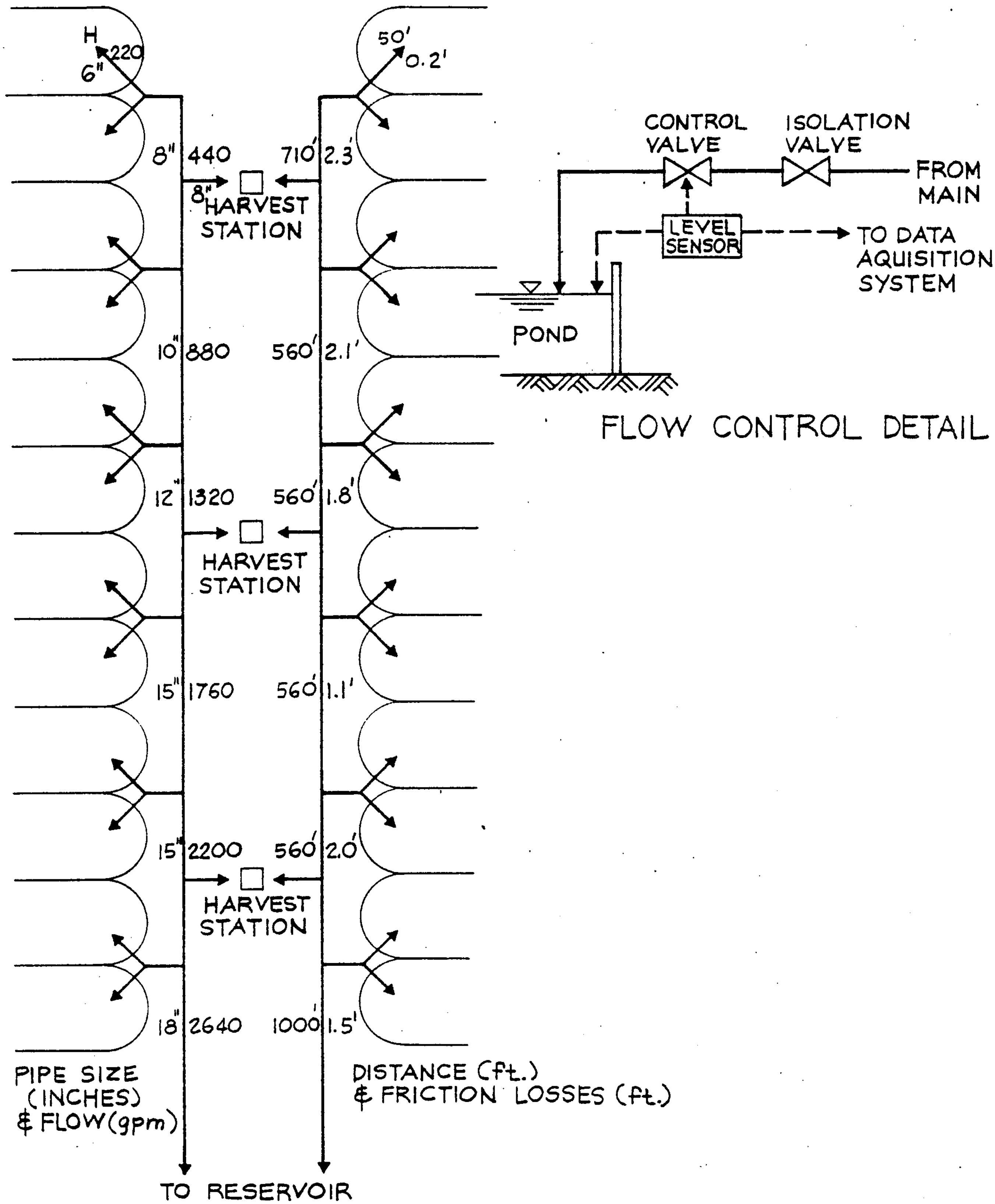


Figure 9-1 Water Distribution System

9.2 CO_2 SUPPLY AND DISTRIBUTION

The source of CO_2 for microalgae production has been examined in some detail [9]. It was concluded that CO_2 could be acquired from existing or proposed power plants and the competition from enhance oil recover (EOR) would not pose a serious problem. The cost of CO_2 derived in this fashion depended on a number of assumptions regarding scale, location, ownership of the power plant, etc., and ranged from \$0.07 - \$0.21/m^3. This report does not attempt to refine the SERI estimates. Instead, the CO_2 cost is treated as a variable, and subjected to a sensitivity analysis in Section 11.

The design of the distribution system depends upon the pressure at which CO_2 is delivered to the site, assuming that about 2.5 psi is required at the pond to satisfy the static head and sparger losses (1.5, 1 psi respectively). If this pressure is high (>50 psi), then the distribution network can consist of very small (6" and less) pipe sizes. For medium pressures (e.g. 10 psi), pipe sizes would range from 8" to 3". For very low pressures (3.5 psi, or a 1 psi drop through the distribution system), sizes would range from 12" to 4". The latter might apply to a system which recovered unpurified flue gas at atmospheric pressure. In such a case, a trade-off exists between compressing to medium pressure (lower capital cost, higher operating costs) versus low pressure (higher capital cost, lower operating cost). A high pressure distribution system is not practical, since the compressors required are very expensive compared to the high volume turboblowers that would be used for low or medium pressures. The low vs. medium pressure trade-off was not evaluated, because the impact on overall costs is very small. The medium pressure case was used to size the CO_2 distribution network. The cost of CO_2 is assumed to include delivery at 10 psi, except in the nutrient recycle scenario, where compression costs for the anaerobic lagoon flue gas are included (see Sections 6 & 11).

Since the supply costs are included in the delivered unit CO_2 cost, and the flow metering and control systems were included under growth pond carbonation (Section 7.3.6), the costs shown in Table 9-1 consist solely of the piping network, which is itemized in Table 9-4. Except for the pipe sizes, the network is identical to that of the makeup water distribution system, and therefore is not shown.

9.3 NUTRIENTS

The types of nutrient required and their source is discussed in Section 10. Liquid ammonia is used for nitrogen requirements. All other nutrients are purchased as salts. Since all of the nutrients are added at the beginning of the batch, with the possible exception of some additional ammonia during the first two days, the facilities can be located at the harvesting stations, with nutrients being added into the return flow network. If necessary, an ammonia tank can be located at each set of ponds for supplemental nitrogen. In general, a batch growth cycle greatly simplifies the process of nutrient additions.

Since the cost of adding nutrients (other than CO_2) is minor, a detailed design was not performed. An estimate of $40,000 per harvesting station plus $2500 per set of ponds, for a total of $150,000 was included in Table 9-1.

TABLE 9-4 CO_2 DISTRIBUTION NETWORK CAPITAL COSTS

Pipe Size inches	Total Distance meters	feet	Installed Cost $/ft	Total Cost $
8	1348	4420	5.00	$22,100
6	683	2240	4.00	8,960
4	372	1220	3.50	4,270
3(sch 40)	366	1200	3.50	4,200
PIPE TOTAL				39,500
VALVES, FITTINGS, MISC.				10,500
TOTAL				$50,000
TOTAL/HECTARE				$260

9.4 BLOWDOWN DISPOSAL SYSTEM

The issue of blowdown was discussed in Section 2. A blowdown rate equivalent to 14% of the evaporative rate is specified. For an average evaporation of 1 cm/day, the blowdown flowrate is 2700m^3/day (95,000 ft^3/day). The blowdown is taken from the centrifuge effluent stream, and pumped to evaporation ponds, where it is concentrated back to solid form. Ultimate disposal is by trucking the salt to a suitable landfill. The evaporation ponds cover 27 hectares (14% of the growth pond area), and consist of shallow, unlined earthen ponds. These ponds cost about $110,000 to construct, with another $50,000 budgeted for pumps, piping and misc., for a total of $160,000. A cost of 1c/kg algae (0.67c/kg salt) is included in the operating cost for ultimate salt disposal.

9.5 BUILDINGS

Buildings include an office, lab, shop, and storage sheds for nutrients and machinery. The cost of buildings for harvesting machinery was included in the harvesting cost estimates. The offices, lab, and culture rooms can be large permanent trailers. The shop has a concrete floor and utilities. The sheds have concrete floors, are open on one side, and have a minimum of utilities. A rough estimate of the building costs is given in Table 9-5.

Table 9-5 Building Costs

Offices (trailers): 1000 ft^2 @ \$30/ft^2	\$30,000
Lab, culture rooms (trailers): 1000 ft^2 @ \$30/ft^2	\$30,000
Shop: 2200 ft^2 @ \$15/ft^2	33,000
Sheds: 2000 ft^2 @ \$8.5/ft^2	17,000
TOTAL	\$110,000

9.6 ROADS AND DRAINAGE

Roads and drainage are difficult to estimate in the absence of a specific site. If the perimeter walls of the pond system are earth berms, then the cost of upgrading them to roads (unpaved) is minor, since large amounts of crushed rock is already being imported for pond lining. Drainage would involve construction of diversion channels (if necessary), placing of culverts, etc, and would be integrated into grow pond earthworks. A figure of \$100,000 is assigned to roads and drainage.

9.7 ELECTRICAL SUPPLY AND DISTRIBUTION

Electrical distribution includes the cost of all on-site power distribution and wiring, exclusive of the utility service, substation, and motor starters and controllers (the latter were included in the individual cost estimates). Power distribution systems are normally estimated in terms of total facility cost, taking both complexity and scale into account. Given the size and relative simplicity of the system, a factor of 3% is used. Considering that so much of the facilities cost is for earthworks and pond walls, which have no electrical power requirements, this estimate is quite reasonable.

In order to arrive at a figure for electrical supply, it was arbitrarily assumed that 5 miles of transmission lines would be required. A recent estimate for from the Imperial Irrigation District of \$3/ft (essentially independent of load) was used for transmission line costs (\$80,000). An additional \$43,000 (again IID estimate) was allocated for the substation cost (1000 kVA), for a total of \$113,000 for electrical supply.

9.8 MACHINERY

A budget of $80,000 is included for vehicles and machinery. Typical requirements would be two pick-up trucks, one flatbed truck, a bachhoe-loader, and misc. shop machinery.

9.9 OTHER COSTS SHOWN IN TABLE 9-1

The remaining costs shown in Table 9-1 (engineering, contingency, land), are taken from the SERI economic model. The land cost is based on a total land requirement of two times the growth pond area. Actual land requirements are quite sensitive to the well spacing, which is difficult to estimate. However, the contribution of land cost to total cost is minor, so that an error will have very little effect, especially on an annualized basis.

SECTION 10.0

LARGE SCALE SYSTEM OPERATIONS AND OPERATING COSTS

10.1 OPERATIONS OVERVIEW

The operations described below form the Base Case that serves as the reference for cost sensitity analyses in Section 11.0. Deviations from this case are so labeled.

The basic operation of the conceptual large scale system consists of batch growth followed by batch accumulation of lipids (to 50% of the ash free dry weight) induced by nitrogen limitation. Both processes are invisioned to take place in the same pond. Recent results of Laws [3] indicate that high productivity of nitrogen sufficient biomass is attainable in batch culture. The biomass productivity after nitrogen depletion from the medium, in batch operation, has been shown [1,10] to be nearly equal to that of nitrogen sufficient batch growth when carbohydrate is accumulated. For lipid accumulation, a higher heating value must be ascribed to the biomass so that, on a dry weight basis some decrease in biomass productivity must be expected. There is little or no evidence that algae have a high enough lipid content during active growth. Since nitrogen sufficient batch growth appears to be productive, and lipid accumulation must be done in batch, there is no real reason to operate any of the system continuously. Cells would accumulate storage products under nitrogen depleted continuous dilution, but much of the biomass harvested would not be induced, or only partially so.

The average productivity of the system is postulated to be 30 $gm/m^2/day$ averaged over a 360 day year. This is equal to a photosynthetic efficiency of 7.5% of visible light. Productivity is expected to be lower during the late fall, winter, and early spring and higher at other times. The density of cultures reflects this, being lower during cooler, less sunny times of the year. Since productivity is not a linear function of temperature and insolation, batch density would be lower when productivity is lower, even though detention times are longer. At times of high productivity, nitrogen sufficient batch growth would last from 1.5 to 2 days followed by a 2-3 day induction period. At times of low to moderate productivity, these periods would be, on average, 2-3 days and 3-4 days. The density achieved when nitrogen first becomes depleted from the medium are, for the two cases respectively, 400-500 ppm and 250-300 ppm. This would increase under induction conditions by a factor of about two to 800-1000 ppm and 500-600 ppm for the two average cases. Thus, at the time of harvest the density of the cultures would range between 500 and 1000 ppm with a lipid content of about 50% and a nitrogen content of about 4%, with 20-25 % protein 15-20% carbohydrate and a balance of 10% other chemicals.

The induced biomass is to be harvested either by a two-stage settling process or by flotation followed by foam collection. Either process would be followed by centrifuging material concentrated by a factor of 50. The primary harvesting processes are not expected to be 100% efficient. This is not a problem since an inoculum must be left for start of the succeeding batch. A harvesting efficiency of 90% is postulated for the Base Case. The centrifugation of the 2-5% slurry increases the density to 10-20% ash free solids.

Since substantial performance is being expected of the primary harvesting process, it may become necessary to use chemical flocculants to aid clarification rate and efficiency. The dose of chemical is postulated to be low (<$.02/kg biomass). In the Base Case, the cost of polymer addition is set at $.01/kg harvested biomass.

The organism required to function in this system must possess several important characteristics. It must be competitive, productive, induce lipid prolifically, not be subject to inhibition by light, oxygen or fluctuations in temperature, and it must be able to recover from depigmentation quickly and become highly productive soon after nitrogen is replenished in the medium. Each of these characteristics has been observed in at least some organism, but not all in any one organism. However, a concerted effort to find such an organism has just recently begun at SERI. Probably the most severe criterion of those listed is the ability to store large amounts of lipid quickly. A recently isolated chrysophyte, called Chryso/F1, is tentatively chosen as a test organism. It acculates lipids prolifically (personal communication, B. Barclay of SERI), has a broad salinity tolerance, but must be tested for its tolerance to high temperature. The Platymonas sp. grown in Hawaii, although not a lipid producer, will be used as a productivity standard, since it is, as of now, the most productive organism grown in intensive, outdoor culture.

Carbon is to be supplied as purified CO_2, introduced into the ponds via sumps 1.5 m deep. The source of carbon, at a 400 hectare scale must either be commercial or, in the best of circumstances a CO_2 well near the site. Power plant carbon containing gases are not a feasible source at this relatively small scale. The algal biomass, after extraction of the lipid component, will be anaerobically decomposed in a covered lagoon from which the liberated gases are recovered. The methane energy is to be converted to heat and electricty with the CO_2 remaining recycled for the growth of more algal biomass. 34% of the carbon needed for new growth can be provided in this manner, reducing the demand of new carbon to algae ratio from 1:1.7 to 1:2.4. The recycle of nutrients from the anaerobic digester is not included in the Base Case.

The water resource specified for use in this system is similar to the Type II water described by SERI. This water has a TDS of 4 ppt, an alkalinty of 13-15 meq/l and a Ca concentration of 10 mM. The water needs to be conditioned by equilibration, in holding ponds, with the atmosphere to induce precipitation of the calcium as calcium carbonate. The precipitate will be removed by alum or lime flocculation, if necessary, and sedimentation. The resulting water is low in calcium (1 mM is the practical lower limit) and still contains about 4 meq/l alkalinity. An eight-fold average concentration of solutes is achieved when an average amount of water is blowndown which equals 1/7 of the average evaporation. Since the system is to be operated in batch, the salinity will fluctuate somewhat, but this is a secondary effect. Thus the pond medium will avarage 32 ppt TDS.

The approximately 1.5 mt of salts produced per mt of algal biomass will be transported to a large body of salt water, or other disposal site, at a cost of .67 cents per kg. The sludge from the lagoon will be dried in sludge drying beds and plowed in on site.

10.2 OPERATIONAL COSTS OF A 1000 ACRE SYSTEM

The parameters of the Base Case design of the large scale system are summarized in Table 10-1. This design is used in Section 11.0 as the reference case for cost sensitivity analysis.

Table 10-1. Base Case Design Parameters

Parameter	Specification
Pond size, hectares	8
Ponds/module	24
Module size, hectares	192
Number of modules	2
Land required/1000 acres growth ponds, acres	2000
Evaporation rate, $m/m^2/yr$	3.6
Blow down ratio	.125
Water requirement, m^3/yr	16.8×10^6
Average well depth, m	50
Pond depth, m	.20
Pond volume, m^3	16000
System volume, m^3	2.7×10^9
TDS of water resource, ppt	4
Operating TDS, ppt	32
Salt disposal, mt/yr	67000
Growing season, days	360
Daily productivity, $gm/m^2/d$	30
Total production, mt/yr	45000
Lipid content, %	50
Pond channel velocity, cm/s	20
Carbon source	Purified CO_2
Nitrogen source	Ammonia
Phosphorus source	Superphosphate

The operating costs of the Base Case design are summarized in Table 10-2. All costs are given on a per hectare per year basis, a per module (192 ha) per year basis, and on a total cost per 1000 acre per year. For the chemical inputs, the unit requirement and the unit cost are given as well. Power unit cost is taken at 6.5 cents per kilwatt hour. Disposal of salts contained in the blowdown (evaporated and transported to a disposal site) is taken as .67 cents per kg salt produced. The breakdown of the maintanance and labor costs are given in Table 9-2 and 10-3 respectively.

The carbon source is purified CO_2, at a low cost of $35/mt. This value is varied in the sensitivity analysis presented later. 2.2 kg CO_2 is required to produce a kg of algal biomass at 50% lipid. Losses are assumed to be less than 5%. Nitrogen, at $250/mt is derived from ammonia, and used with a 75% efficiency to cover losses. The phosphorus, at $900/mt is derived from superphosphate. Iron is added as ferrous sulfate and costs $500/mt Fe. All other nutrients are assumed to be present in the water resource, or as a trace contaminant of the above nutrients. Even with the low price of CO_2 assumed, this one input alone dominates the production price of the algal biomass.

A high molecular weight polymer is included to flocculate the algal biomass prior to primary concentration in a settling pond. It is assumed that 1 ppm of polymer is required for each 500 ppm biomass. The cost of the polymer is $5.00/kg.

The only operating cost associated with the water resource is the cost of pumping from a 50 m deep well and transport to the holding ponds. Water conditioning is presumed to occur during the eight day holding time, and is due to the precipitation of calcium carbonate by the alkalinity in the water upon equilibration with the atmosphere. Power costs are included for mixing, harvesting (dominated by centrifuge power), nutrient supply pumping, and building utilities.

The maintenance operating costs are for materials only, as detailed in Table 9-2 of Section 9.0. The labor for maintenance is included in the personnel breakdown listed in Table 10-3 Estimates are based on five shifts per week, 360 days per operation, and rough judgement of people needed on hand.

For this base case, the total, non-annualized operating costs are $6.2 million for the 45,450 mt of algal biomass produced. At 50% lipid and a conversion of 7.14 bbl lipid per mt lipid, the "lipid oil" produced equals 160,000 bbl/yr. The capital cost summary was given in Section 9.0. Together with the operating cost summary Table 10-2, the total annualized production cost will be calculated in the next section using the SERI economic model.

Table 10-2 Total Operating Costs - Base Case

Basis: 112.3 mt algae/ha/yr (30 gm/sq m/day)

	QUAN REQ'D	UNIT COST	--- YEARLY COST, Thousands ---	192 ha	1000 ac
	kg/kg	$/mt	$/ha/yr	$/yr	$/yr
NUTRIENTS (1)					
CO2	2.2	35	$8.65	$1,660	$3,501
N, as NH3	0.053	250	1.47	283	597
P, as Superphosphate	0.005	900	0.51	97	205
Fe, as FeSO4	0.005	500	0.28	54	114
Total			$10.91	$2,094	$4,416
FLOCCULANT	0.002	5000	$1.12	$216	$455
	10^6kwh/yr*	c/kwh			
POWER					
Mixing	2.06	6.5	0.70	134	282
1 Harvesting	0.34	6.5	0.12	22	47
2 Harvesting	1.10	6.5	0.37	71	150
Water Supply	1.68	6.5	0.57	109	230
Nutrient Supply	0.10	6.5	0.04	7	14
Buildings	0.20	6.5	0.07	13	27
Total	5.48		$1.85	$356.14	$750.95
	kg/kg	c/kg			
SALT DISPOSAL					
	1.5	0.67	1.13	217	457
MAINTANENCE (matl's)	(2)	(2)	1.26	242	511
LABOR	(3)	(3)	1.39	267	562
TOTAL			$16.54	$3,176	$6,697

(1) kg/kg = kg required / kg algae produced
(2) See Table 9-2
(3) See Table 10-3
* 192 ha basis

Table 10-3. Labor Required for 1000 Acre System

Title	No.	Hrs/Yr	$/Hr	$/Yr
Plant Manager	1	2080	25	52000
Shift Supervisors	4	2080	17	141.440
Pond Operators	10	2080	10	166,400
Centrifuge Operators	5	2080	12	124.800
Laboratory Manager	1	2080	17	35,360
Laboratory Technicians	2	2080	10	41.600
			TOTAL	561,600
			$/ha	1390

SECTION 11.0

ECONOMIC ANALYSIS OF A 1000 ACRE ALGAL PRODUCTION SYSTEM

11.1 BASE CASE

The capital cost of the Base Case design was presented in Table 9-1, the operating costs in Table 10-2. Appendix VI contains these tables again as well as the capital and operating costs for all of the sensitivity analyses. In the Base Case the biomass productivity is 45.5 x 10^3 mt/yr at a lipid content of 50%. This is well in excess of any currently achievable lipid productivity. The biomass production value has been achieved, on a year round basis [3], but under conditions of nutrient sufficiency and at 15-20% lipid content. The price of CO_2 used in the Base Case is $35/mt which is about as much as a very large scale user would be charged for commercial CO_2 if transportation distance were less than 200 miles. The installed system cost is $76,600 per hectare using a low cost system design and favorable assumptions concerning the suitability of land and other resources needed for system construction. The total capital cost of construction is $25 million dollars. Thus the Base Case must be considered very optimistic in terms of lipid productivity, nutrient costs, and attainment of capital cost goals. In this section, the sensitivity of the production cost of algal biomass and lipid oil (unextracted), to these assumptions will be analyzed to determine where the future work must be concentrated.

Table 11-1 shows the input and output data of the Base Case using the SERI economic model. The parameters of this model are given in Table 11-2. A quick glance at the economic figures reveals that even at $35/mt, CO_2 costs dominate the Base Case economics. Half of the annualized production costs come from this. Unless the price, or quantity of CO_2 is reduced, the economics of producing low value products from algae cannot improve significantly. In addition, increases in productivity have relatively little effect on final costs when CO_2 is so dominant a factor. Thus only some initial sensitivity is presented with the Base Case as the reference level. The cost of the carbon input is significantly reduced by the addition of an anaerobic lagoon, in which extraction residues containing about half of the algal carbon, are degraded to volatile carbon gases which are recycled to the growth ponds. The design of an efficient, low cost recycling system, is an important feature of the large scale design and experimental system tests.

11.2 SENSITIVITY OF BASE CASE TO PRODUCTIVITY CHANGES

Table 11-3 shows the effect of varying the system biomass productivity on the annualized production costs. Capital costs of some items, like centrifuges, were increased using a scale factor of .7. However, no increases in capital costs were assumed necessary to achieve the two and three fold increases in productivity analyzed. The major increases come in operating costs, particularly in proportional increases in the CO_2 required. Doubling the

Table 11-1 Base Case Economic Analysis

INPUTS	BASE CASE
Depreciable capital investment	$13,940,000
Non-Deprec capital investment	11,060,000
Annual operating costs	6,180,000
Annual maintenance costs	510,000
System algal yield mt/yr	45450
System Lipid yield bbls/yr	150000
Carbohydrate yield mt/yr	9100
Protein yield mt/yr	9100

CALCULATED VALUES	
Cost of capital	0.0475
Cap Recov Factor (book life)	0.0692
Cap Recov Factor (depr life)	0.0947
Fixed Charge Rate	0.0951
PV of Capital Investment	$17,349,560
PV of NonDeprec Investment	13,765,150
PV of Operating costs	123,965,000
PV of Maintenance costs	10,930,980
TOTAL PV OF FACILITY	$166,010,700
Annualized Cost Algae Plant	$11,934,510
LIPID only price $/bbl	79.56
ALGAE price $/mt	262.59
Lipid price by weight $/mt	304.68
Lipid price by value $/mt	284.56
Protein price by weight $/mt	304.68
Protein price by value $/mt	327.87
Carbohyd price by weight $/mt	304.68
Carbohyd price by value $/mt	327.87

Annualized Capital Cost	DCI	$1,650,002
Annualized Capital Cost	NCDI	952,285
Annualized Operating cost	OPP	8,576,007
Annualized Mainten. Cost	MNT	756,215
Annualized Cost Total	AAP	11,934,508
Annualized Capital Cost	DCI	13.8%
Annualized Capital Cost	NCDI	8.0%
Annualized Operating cost	OPP	71.9%
Annualized Mainten. Cost	MNT	6.3%
Annualized Cost Total	AAP	100.0%

Table 11-2 Economic Model Input Parameters

FINANCIAL PARAMETERS	
Base year for constant dollars	1984
Year for investment outlay	1992
Year for cost information	1985
Year first commercial oper	1996
System oper life (book life)	25
Tax life for depreciation	15
Annual other taxes fn of CI	0.01
Annual ins premium fn of CI	0.01
Effective income tax rate	0.46
Investment tax credit	0.09
Debt:total capitalization	0.3
Common Stk:tot capitalization	0.5
Prefer Stk:tot capitalization	0.2
Ann rate return on debt	0.037
Ann rate return common stk	0.065
Ann rate return prefer stk	0.045
Rate of general inflation	0
Escalation capital costs	0.01
Escalation operating costs	0.015
Escalation maintenance costs	0.018
Cost capital (optional)	0
Cap Recovery Factor (optional)	0
Fixed Charge Rate (optional)	0
Lipid credit $/mt	200
Protein credit $/mt	100
Carbohydrate credit $/mt	100

productivity only decreases unit production costs by 20%. Tripling productivity only leads to a further decrease of less than 5%, or 23% reduction from the Base Case. Thus productivity enhancement will only be meaningful if the amount or cost of CO_2 is decreased. The cost was already assumed to be quite low. The recycling of carbon will be seen to dramatically affect the sensitivity of production costs to productivity increases.

Table 11-3. Base Case: Sensitivity to Productivity

Cost	BASE	2XPROD.	3XPROD.
LIPID only price.$.bbl	79.56	63.79	61.52
ALGAE price $/mt	262.59	210.30	202.83
Lipid price by weight $/mt	304.68	244.28	235.60
Lipid price by value $/mt	284.56	228.15	220.04
Protein price by weight $/mt	304.68	244.28	235.60
Protein price by value $/mt	327.87	262.88	253.53
CHO price by weight $/mt	304.68	244.28	235.60
CHO price by value $/mt	327.87	262.88	253.53
DCI, % total	13.8	10.3	7.8
NDCI, % total	8.0	5.4	3.9
OPP, % total	71.9	79.4	82.9
MNT, %total	6.3	4.9	5.4
AAP, %	100.0	100.0	100.0

11.3 SENSITIVITY OF THE BASE CASE TO CARBON RECYCLE

The low cost anaerobic digester is a major proposed innovation in that recovery of the input nutrients for return to the algal growth ponds has not been attempted on large scale. The efficiency of the recovery and reuse of nutrients from biomass degraded in a low cost, covered lagoon is a factor that must be empirically determined in the proposed experimental system. The lagoon not only allows recycle of carbon (33%) but also up to 75% of the nitrogen and 50% of the phosphorus. In addition, the methane produced is combusted to produce electrical power, at about 25-30% efficiency. This electricity can be used on site, for mixing power, etc. In some of the cases analyzed, the electrical output of the digestion system exceeds the needs of the system. In these cases, the surplus value of the power produced, calculated at 6.5 cents/Kw.hr, is added as a negative number to operating costs.

The recycle of carbon reduces the demand of new input carbon, per kg algal biomass, from 2.2kg to 1.6 kg. Table 11-4 shows the effect of the savings in CO_2 quantities on the annualized production costs of biomass and lipid. The capital costs of the Base Case were modified to include the costs of the covered lagoon and engine generator. Maintanance costs increased, but operating costs decreased due to the lower quantity of CO_2 used and the power generated. Since all of the unextracted biomass goes into the lagoon

Table 11-4 Economic Analysis: Base Case + Recycle

INPUTS	BASE CASE	RECYCLE
Depreciable capital investment	$13,940,000	$17,620,000
Non-Deprec capital investment	11,060,000	11,430,000
Annual operating costs	6,180,000	3,640,000
Annual maintenance costs	510,000	800,000
System algal yield mt/yr	45450	45450
System Lipid yield bbls/yr	150000	150000
Carbohydrate yield mt/yr	9100	
Protein yield mt/yr	9100	

CALCULATED VALUES

Cost of capital	0.0475	0.0475
Cap Recov Factor (book life)	0.0692	0.0692
Cap Recov Factor (depr life)	0.0947	0.0947
Fixed Charge Rate	0.0951	0.0951
PV of Capital Investment	$17,349,560	$21,929,650
PV of NonDeprec Investment	13,765,150	14,225,640
PV of Operating costs	123,965,000	73,014,990
PV of Maintenance costs	10,930,980	17,146,630
TOTAL PV OF FACILITY	$166,010,700	$126,316,900
Annualized Cost Algae Plant	$11,934,510	$9,307,186
LIPID only price $/bbl	79.56	62.05
ALGAE price $/mt	262.59	204.78
Lipid price by weight $/mt	304.68	443.82
Lipid price by value $/mt	284.56	221.91
Protein price by weight $/mt	304.68	443.82
Protein price by value $/mt	327.87	
Carbohyd price by weight $/mt	304.68	443.82
Carbohyd price by value $/mt	327.87	

Annualized Capital Cost	DCI	$1,650,002	$2,085,584
Annualized Capital Cost	NCDI	952,285	984,142
Annualized Operating cost	OPP	8,576,007	5,051,241
Annualized Mainten. Cost	MNT	756,215	1,186,219
Annualized Cost Total	AAP	11,934,508	9,307,186
Annualized Capital Cost	DCI	13.8%	22.4%
Annualized Capital Cost	NCDI	8.0%	10.6%
Annualized Operating cost	OPP	71.9%	54.3%
Annualized Mainten. Cost	MNT	6.3%	12.7%
Annualized Cost Total	AAP	100.0%	100.0%

for digestion, there are no carbohydrate or protein by products. Although not included in the model, the lagoon sludge is a potential fertilizer product. The production cost of biomass drops 22% compared to the base case, from \$263/mt to \$205/mt. The lipid only price drops from \$79.6/bbl to \$62/bbl. This is as much a decrease, relative to the Base Case, as tripling the productivity provided. More importantly, it will be shown below that now productivity increases have much more impact on final cost than they did without carbon recycle. This is due to the decrease, in terms of per cent of total annualized cost, of operating relative to capital outlays. In the Base Case operating costs contributed over 70% of the total, capital only 22%. With nutrient recycle and power generation, operating drops to 54% of the total and capital rises to one third.

11.4 RECYCLE CASE: PRODUCTIVITY SENSITIVITY

The results of variations in assumed biomass productivity, with 50% lipid, on the annualized production costs, are presented in Table 11-5. The 1.0 x productivity case refers to the nutrient recycle case discussed above. Doubling productivity now reduces production cost 30%, whereas without recycle the reduction was 20%. Biomass production cost is \$144/mt, lipid price is \$44/bbl. If productivity is only increased by 50%, a 20% savings is still achieved in production costs. When processing of the biomass is included, i.e., extraction and conversion to useable fuel, the impact of productivity increases is likely to be reduced. The production costs are sensitive to reductions in productivity, with a 30% reduction leading to almost a 30% increase in cost. Thus, a productivity of 30 $gm/m^2/day$ on average with a 50% lipid content, appears to be a high but economically reasonable goal. Decreases in productivity have substantial impact, but increases above 1.5 fold begin to have diminishing returns as well as being unrealistic. This is shown in Figure 11-1 as a decrease in the slope of the graph of lipid price versus productivity.

11.5 RECYCLE CASE: SENSITIVITY TO CO2 PRICE

Although the price of CO_2 used in all of the preceding analyses is not inconsistent with commercial CO_2 cost under favorable circumstances, the quantities needed for very large scale systems, and the remote siting of these systems, may preclude the use of commercial CO_2. In this case, the \$35/mt price is a low one, as purification of CO_2 from power plant stack gases is more expensive than from refinery off gases. To demonstrate the sensitivity of the design to variation of the assumed price of CO_2, two cases were analyzed: free CO_2 and \$70/mt CO_2. The results are shown in Table 11-6. Doubling the cost of carbon increases annualized production costs about 40%. When CO_2 is free (an exceptional case only applicable to small systems located near an existing well) the costs are reduced 40%. Thus again, the Base Case assumption on carbon price appears to be an economically reasonable one that must be achieved. The addition of nutrient recycle to this cost is one of the most necessary features to successfully develop. Carbon is too costly, in general, to allow half of it to escape from the system without being incorporated into product.

Table 11-5 Recycle Case: Productivity Sensitivity

INPUTS	R, 0.5xP	R, 0.67xP	R, 1.0xP
Depreciable capital investment	$14,570,000	$15,290,000	$17,620,000
Non-Deprec capital investment	10,820,000	11,030,000	11,430,000
Annual operating costs	2,650,000	2,990,000	3,640,000
Annual maintenance costs	660,000	690,000	800,000
System algal yield mt/yr	22725	30315	45450
System Lipid yield bbls/yr	75000	100000	150000
Carbohydrate yield mt/yr			
Protein yield mt/yr			
CALCULATED VALUES			
Cost of capital	0.0475	0.0475	0.0475
Cap Recov Factor (book life)	0.0692	0.0692	0.0692
Cap Recov Factor (depr life)	0.0947	0.0947	0.0947
Fixed Charge Rate	0.0951	0.0951	0.0951
PV of Capital Investment	$18,133,650	$19,029,750	$21,929,650
PV of NonDeprec Investment	13,466,450	13,727,810	14,225,640
PV of Operating costs	53,156,520	59,976,600	73,014,990
PV of Maintenance costs	14,145,970	14,788,970	17,146,630
TOTAL PV OF FACILITY	$98,902,580	$107,523,100	$126,316,900
Annualized Cost Algae Plant	$7,312,236	$7,931,843	$9,307,186
LIPID only price $/bbl	97.50	79.32	62.05
ALGAE price $/mt	321.77	261.65	204.78
Lipid price by weight $/mt	697.38	567.36	443.82
Lipid price by value $/mt	348.70	283.68	221.91
Protein price by weight $/mt	697.38	567.36	443.82
Protein price by value $/mt			
Carbohyd price by weight $/mt	697.38	567.36	443.82
Carbohyd price by value $/mt			
Annualized Capital Cost DCI	$1,724,572	$1,809,794	$2,085,584
Annualized Capital Cost NCDI	931,621	949,702	984,142
Annualized Operating cost OPP	3,677,414	4,149,233	5,051,241
Annualized Mainten. Cost MNT	978,631	1,023,114	1,186,219
Annualized Cost Total AAP	7,312,237	7,931,843	9,307,186
Annualized Capital Cost DCI	23.6%	22.8%	22.4%
Annualized Capital Cost NCDI	12.7%	12.0%	10.6%
Annualized Operating cost OPP	50.3%	52.3%	54.3%
Annualized Mainten. Cost MNT	13.4%	12.9%	12.7%
Annualized Cost Total AAP	100.0%	100.0%	100.0%

Table 11-5 Recycle Case: Productivity Sensitivity (continued)

INPUTS	R, 1.5xP	R, 2xP
Depreciable capital investment	$19,640,000	$21,900,000
Non-Deprec capital investment	12,000,000	12,550,000
Annual operating costs	4,630,000	5,620,000
Annual maintenance costs	1,000,000	1,120,000
System algal yield mt/yr	68175	90900
System Lipid yield bbls/yr	225000	300000
Carbohydrate yield mt/yr		
Protein yield mt/yr		
CALCULATED VALUES		
Cost of capital	0.0475	0.0475
Cap Recov Factor (book life)	0.0692	0.0692
Cap Recov Factor (depr life)	0.0947	0.0947
Fixed Charge Rate	0.0951	0.0951
PV of Capital Investment	$24,443,710	$27,256,480
PV of NonDeprec Investment	14,935,060	15,619,580
PV of Operating costs	92,873,460	112,731,900
PV of Maintenance costs	21,433,290	24,005,290
TOTAL PV OF FACILITY	$153,685,500	$179,613,300
Annualized Cost Algae Plant	$11,265,740	$13,132,360
LIPID only price $/bbl	50.07	43.77
ALGAE price $/mt	165.25	144.47
Lipid price by weight $/mt	358.15	313.12
Lipid price by value $/mt	179.08	156.56
Protein price by weight $/mt	358.15	313.12
Protein price by value $/mt		
Carbohyd price by weight $/mt	358.15	313.12
Carbohyd price by value $/mt		
Annualized Capital Cost DCI	$2,324,680	$2,592,184
Annualized Capital Cost NCDI	1,033,220	1,080,576
Annualized Operating cost OPP	6,425,067	7,798,891
Annualized Mainten. Cost MNT	1,482,774	1,660,707
Annualized Cost Total AAP	11,265,741	13,132,358
Annualized Capital Cost DCI	20.6%	19.7%
Annualized Capital Cost NCDI	9.2%	8.2%
Annualized Operating cost OPP	57.0%	59.4%
Annualized Mainten. Cost MNT	13.2%	12.6%
Annualized Cost Total AAP	100.0%	100.0%

Figure 11-1 Productivity Sensitivity

With Nutrient Recycle

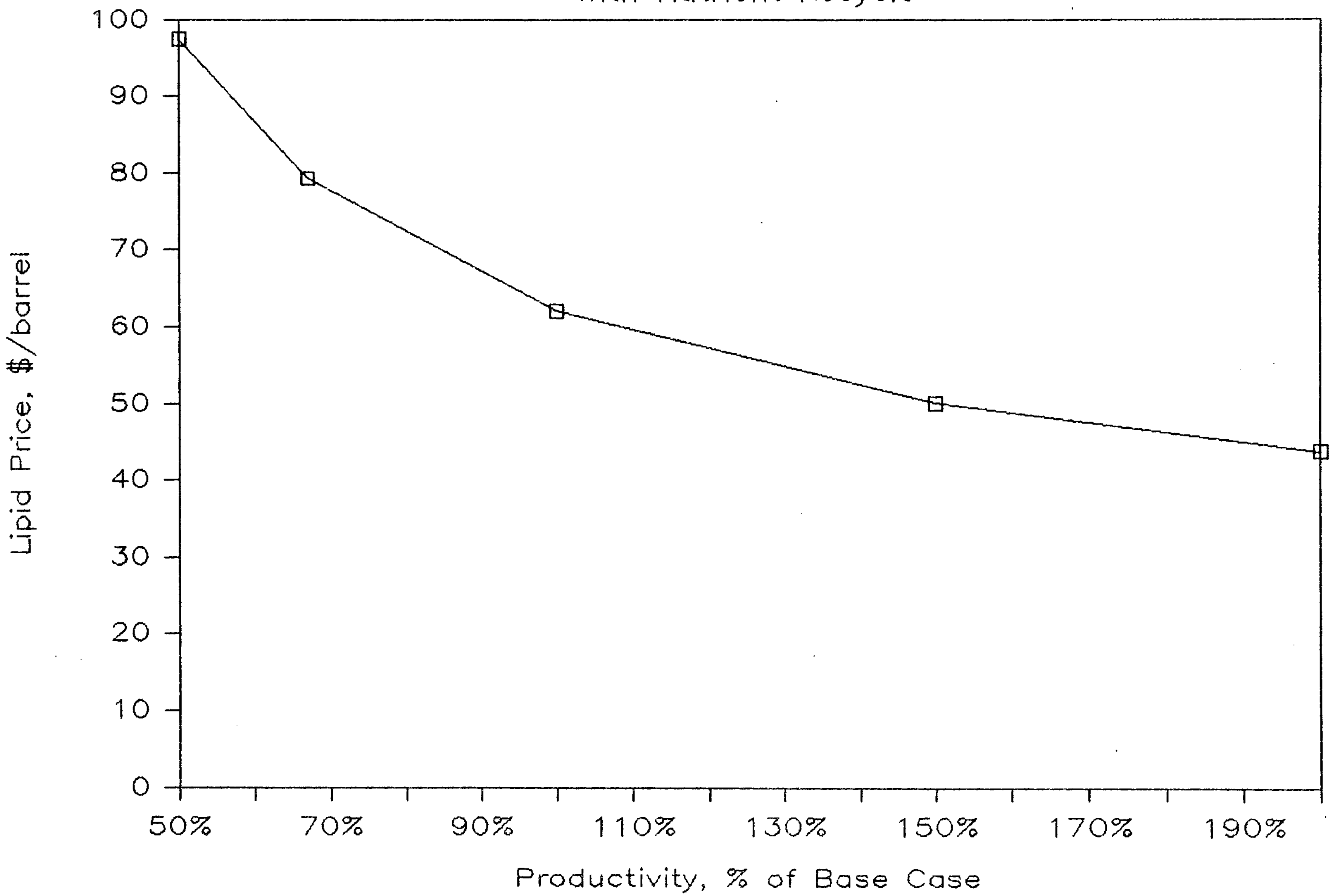


Table 11-6 Recycle Case: Sensitivity to CO2 Price

INPUTS	R, Free	R, $35/mt	R, $70/mt
Depreciable capital investment	$17,300,000	$17,620,000	$17,300,000
Non-Deprec capital investment	11,420,000	11,430,000	11,420,000
Annual operating costs	1,100,000	3,640,000	6,180,000
Annual maintenance costs	800,000	800,000	800,000
System algal yield mt/yr	45450	45450	45450
System Lipid yield bbls/yr	150000	150000	150000
Carbohydrate yield mt/yr			
Protein yield mt/yr			
CALCULATED VALUES			
Cost of capital	0.0475	0.0475	0.0475
Cap Recov Factor (book life)	0.0692	0.0692	0.0692
Cap Recov Factor (depr life)	0.0947	0.0947	0.0947
Fixed Charge Rate	0.0951	0.0951	0.0951
PV of Capital Investment	$21,531,380	$21,929,650	$21,531,380
PV of NonDeprec Investment	14,213,200	14,225,640	14,213,200
PV of Operating costs	22,064,970	73,014,990	123,965,000
PV of Maintenance costs	17,146,630	17,146,630	17,146,630
TOTAL PV OF FACILITY	$74,956,180	$126,316,900	$176,856,200
Annualized Cost Algae Plant	$5,743,681	$9,307,186	$12,793,210
LIPID only price $/bbl	38.29	62.05	85.29
ALGAE price $/mt	126.37	204.78	281.48
Lipid price by weight $/mt	273.89	443.82	610.06
Lipid price by value $/mt	136.95	221.91	305.03
Protein price by weight $/mt	273.89	443.82	610.06
Protein price by value $/mt			
Carbohyd price by weight $/mt	273.89	443.82	610.06
Carbohyd price by value $/mt			
Annualized Capital Cost DCI	$2,047,707	$2,085,584	$2,047,707
Annualized Capital Cost NCDI	983,282	984,142	983,282
Annualized Operating cost OPP	1,526,474	5,051,241	8,576,007
Annualized Mainten. Cost MNT	1,186,219	1,186,219	1,186,219
Annualized Cost Total AAP	5,743,682	9,307,186	12,793,214
Annualized Capital Cost DCI	35.7%	22.4%	16.0%
Annualized Capital Cost NCDI	17.1%	10.6%	7.7%
Annualized Operating cost OPP	26.6%	54.3%	67.0%
Annualized Mainten. Cost MNT	20.7%	12.7%	9.3%
Annualized Cost Total AAP	100.0%	100.0%	100.0%

11.6 RECYCLE CASE: SENSITIVITY TO LIPID CONTENT

Lipid content of the biomass produced (at 30 $gm/m^2/d$) was varied from 20% to 50% (the Base Case value) to determine the effect on lipid price. As shown in Table 11-7, reduced lipid dramatically increases lipid price. Thus the Base Case assumption of 50% lipid content needs to be achieved when lipid oil is the desired product. Recycle of carbon was assumed in this analysis and had the effect of causing a decrease in biomass production cost as lipid content was decreased. This analysis could be adapted to the case of methane fuel production if the non-lipid fraction is assumed accessible to degradation and if the amount of carbon recycle is corrected for removal of carbon as the methane product. No such analysis was done in detail as lipid production is the primary goal. The system would have to produce biomass at $75/mt to provide methane at $5/MBTU. The system designed is effectve in producing methane gas if the covered lagoon digester performs as well as assumed. This will be studied in the proposed experimental system.

11.7 RECYCLE CASE: SENSITIVITY TO DEPRECIABLE CAPITAL INVESTMENT

The capital investment does not appear to significantly affect the annualized production costs until it gets large. Three variations in DCI were analyzed for their impact. The first, 1.5 times the DCI of the recycle case would apply if the primary harvesting device were much more expensive than the proposed sedimentation system. An example of a device that would fall into this range of DCI increase is an engineered dissolved gas flotation reactor, costing about $12 million per 1000 acre system. The second change in DCI, to twice the baseline level, corresponds to lining the ponds with a high quality synthetic material costing @$5.00/$m^2$, or about $22 million dollars per 1000 acres of growth pond area. This type of liner would be used if ponds lined with less expensive earthen materials proved to lose too much water. The five fold increase in DCI would correspond to lining and covering a pond system. This is very costly and as yet there is no evidence that, on a large scale, this concept is at all workable. As shown in Table 11-8, the 50% increase in DCI leads to a 15% increase in product cost which implies that it is affordable if it is necessary for good performance of the harvesting system. The 100% increase in DCI, for a liner, increases cost of lipid product by 30%. This is affordable only if the increase is offset by some other savings or by productivity increases. Thus use of a liner is marginal. The 400% increase, for a covered system, results in 2.25 fold increase in production costs, which is by all means too much even if the covered system somehow made all other performance goals achievable. It must be kept in mind that synthetic materials used for lining and covering will likely inflate with crude oil prices, so that hoping for dramatic oil price increases to make such systems affordable is not warranted.

Table 11-7 Recycle Case: Sensitivity to Lipid Content

INPUTS	R,20% Lipid	R,30% Lipid	R,40% Lipid
Depreciable capital investment	$19,120,000	$18,370,000	$17,910,000
Non-Deprec capital investment	11,850,000	11,710,000	11,570,000
Annual operating costs	1,910,000	2,440,000	3,040,000
Annual maintenance costs	1,000,000	940,000	880,000
System algal yield mt/yr	45450	45450	45450
System Lipid yield bbls/yr	60000	90000	120000
Carbohydrate yield mt/yr			
Protein yield mt/yr			
CALCULATED VALUES			
Cost of capital	0.0475	0.0475	0.0475
Cap Recov Factor (book life)	0.0692	0.0692	0.0692
Cap Recov Factor (depr life)	0.0947	0.0947	0.0947
Fixed Charge Rate	0.0951	0.0951	0.0951
PV of Capital Investment	$23,796,530	$22,863,090	$22,290,580
PV of NonDeprec Investment	14,748,370	14,574,130	14,399,890
PV of Operating costs	38,312,810	48,944,110	60,979,550
PV of Maintenance costs	21,433,290	20,147,290	18,861,300
TOTAL PV OF FACILITY	$98,291,010	$106,528,600	$116,531,300
Annualized Cost Algae Plant	$7,416,723	$7,962,412	$8,639,566
LIPID only price $/bbl	123.61	88.47	72.00
ALGAE price $/mt	163.18	175.19	190.09
Lipid price by weight $/mt	884.18	632.82	514.98
Lipid price by value $/mt	442.10	316.42	257.50
Protein price by weight $/mt	884.18	632.82	514.98
Protein price by value $/mt			
Carbohyd price by weight $/mt	884.18	632.82	514.98
Carbohyd price by value $/mt			

Annualized Capital Cost	DCI	$2,263,131	$2,174,358	$2,119,910
Annualized Capital Cost	NCDI	1,020,305	1,008,251	996,197
Annualized Operating cost	OPP	2,650,514	3,385,996	4,218,618
Annualized Mainten. Cost	MNT	1,482,774	1,393,807	1,304,841
Annualized Cost Total	AAP	7,416,723	7,962,412	8,639,566
Annualized Capital Cost	DCI	30.5%	27.3%	24.5%
Annualized Capital Cost	NCDI	13.8%	12.7%	11.5%
Annualized Operating cost	OPP	35.7%	42.5%	48.8%
Annualized Mainten. Cost	MNT	20.0%	17.5%	15.1%
Annualized Cost Total	AAP	100.0%	100.0%	100.0%

Table 11-8 Recycle Case: Sensitivity to DCI

INPUTS		R,1.5xDCI	R,2xDCI	R,5xDCI
Depreciable capital investment		$26,430,000	$35,240,000	$88,100,000
Non-Deprec capital investment		11,430,000	11,430,000	11,430,000
Annual operating costs		3,640,000	3,640,000	3,640,000
Annual maintenance costs		1,060,000	1,320,000	2,910,000
System algal yield mt/yr		45450	45450	45450
System Lipid yield bbls/yr		150000	150000	150000
Carbohydrate yield mt/yr				
Protein yield mt/yr				
CALCULATED VALUES				
Cost of capital		0.0475	0.0475	0.0475
Cap Recov Factor (book life)		0.0692	0.0692	0.0692
Cap Recov Factor (depr life)		0.0947	0.0947	0.0947
Fixed Charge Rate		0.0951	0.0951	0.0951
PV of Capital Investment		$32,894,470	$43,859,290	$109,648,200
PV of NonDeprec Investment		14,225,640	14,225,640	14,225,640
PV of Operating costs		73,014,990	73,014,990	73,014,990
PV of Maintenance costs		22,719,290	28,291,940	62,370,880
TOTAL PV OF FACILITY		$142,854,400	$159,391,900	$259,259,700
Annualized Cost Algae Plant		$10,735,500	$12,163,810	$20,778,170
LIPID only price $/bbl		71.57	81.09	138.52
ALGAE price $/mt		236.20	267.63	457.17
Lipid price by weight $/mt		511.94	580.05	990.83
Lipid price by value $/mt		255.97	290.03	495.42
Protein price by weight $/mt		511.94	580.05	990.83
Protein price by value $/mt				
Carbohyd price by weight $/mt		511.94	580.05	990.83
Carbohyd price by value $/mt				
Annualized Capital Cost	DCI	$3,128,376	$4,171,168	$10,427,917
Annualized Capital Cost	NCDI	984,142	984,142	984,142
Annualized Operating cost	OPP	5,051,241	5,051,241	5,051,241
Annualized Mainten. Cost	MNT	1,571,740	1,957,261	4,314,872
Annualized Cost Total	AAP	10,735,499	12,163,811	20,778,171
Annualized Capital Cost	DCI	29.1%	34.3%	50.2%
Annualized Capital Cost	NCDI	9.2%	8.1%	4.7%
Annualized Operating cost	OPP	47.1%	41.5%	24.3%
Annualized Mainten. Cost	MNT	14.6%	16.1%	20.8%
Annualized Cost Total	AAP	100.0%	100.0%	100.0%

11.8 RECYCLE CASE: SENSITIVITY TO COST OF CAPITAL

All of the modeling done so far is based on assumptions which lead to a cost of capital of about 5%. While this may be reasonable in some circumstances, many private investors will want to see a greater return for risking capital or for investing in a capital intensive business. Using the recycle case, three cost of capital factors were used to test sensitivity to this parameter, each with the original DCI and with the 100% increased DCI. The results are given in Tables 11-9 and 11-10.

As expected, the cost of capital has great impact on final product costs and this impact increases as capital costs increase. The contribution of annualized capital cost to total annualized cost increases from 33% at the lowest rate and lowest level of capital requirement to 60% at the highest of each. This means that, except at the most favorable borrowing rate, the capital intensity of a system is an important factor. Very expensive systems must be justified in terms of very great increases in productivity. However productivity has an upper limit which is less two than times the Base Case value, i.e., it is doubtful that a lipid accumulating alga could ever be expected to be 15% efficient in converting solar energy. For these reasons, the design criteria used in this report, lowest cost with highest effect, appears to be justified.

11.9 RECYCLE CASE: SENSITIVITY TO POLYMER DOSE

Harvesting is one of the steps in the overall process that must be highly reliable. In the basic design of the system, it was assumed that a high molecular weight polymer would have to be added to the algal suspension to aid in flocculating the biomass prior to primary concentration. The dose of polymer used, 1 ppm per 500 ppm of biomass, is a low one. The results of the sensitivity analysis shown in Table 11-11 indicate that this dose could be doubled without increasing product cost more than 7%, and even a five fold higher polymer dose would only lead to a 20% increase. A low dose, 1-3 ppm/ppm, is becoming state of the art due to the capability of tailoring polymers to specific applications.

11.10 CONCLUSIONS

The sensitivity analyses performed confirm that the optimistic assumptions used in setting the large scale system performance standards are necessary to produce fuel products from algae economically. The analyses also indicate that a low cost, carbon efficient system has the best opportunity of achieving these standards. Productivity increases, which appear necessary despite the high lipid productivity assumed here, will have the most impact on final product cost when achieved within the framework of the system specified here. The questions of resource availability were not, of course, solved by these analyses. However, the system designed is very efficient in its utilization of resources.

Table 11-9 Recycle Case: Sensitivity to Cost of Capital (k)

INPUTS		R, k=.047	R, k=.07	R, k=.10
Depreciable capital investment		$17,620,000	$17,620,000	$17,620,000
Non-Deprec capital investment		11,430,000	11,430,000	11,430,000
Annual operating costs		3,640,000	3,640,000	3,640,000
Annual maintenance costs		800,000	800,000	800,000
System algal yield mt/yr		45450	45450	45450
System Lipid yield bbls/yr		150000	150000	150000
Carbohydrate yield mt/yr				
Protein yield mt/yr				
CALCULATED VALUES				
Cost of capital		0.0475	0.0700	0.1000
Cap Recov Factor (book life)		0.0692	0.0858	0.1102
Cap Recov Factor (depr life)		0.0947	0.1098	0.1315
Fixed Charge Rate		0.0951	0.1202	0.1581
PV of Capital Investment		$21,929,650	$23,373,740	$25,395,390
PV of NonDeprec Investment		14,225,640	15,162,420	16,473,860
PV of Operating costs		73,014,990	57,974,420	44,343,990
PV of Maintenance costs		17,146,630	13,572,460	10,342,760
TOTAL PV OF FACILITY		$126,316,900	$110,083,000	$96,556,000
Annualized Cost Algae Plant		$9,307,186	$10,250,610	$11,853,790
LIPID only price $/bbl		62.05	68.34	79.03
ALGAE price $/mt		204.78	225.54	260.81
Lipid price by weight $/mt		443.82	488.81	565.26
Lipid price by value $/mt		221.91	244.41	282.63
Protein price by weight $/mt		443.82	488.81	565.26
Protein price by value $/mt				
Carbohyd price by weight $/mt		443.82	488.81	565.26
Carbohyd price by value $/mt				
Annualized Capital Cost	DCI	$2,085,584	$2,810,038	$4,014,160
Annualized Capital Cost	NCDI	984,142	1,301,095	1,814,894
Annualized Operating cost	OPP	5,051,241	4,974,813	4,885,293
Annualized Mainten. Cost	MNT	1,186,219	1,164,659	1,139,442
Annualized Cost Total	AAP	9,307,186	10,250,605	11,853,790
Annualized Capital Cost	DCI	22.4%	27.4%	33.9%
Annualized Capital Cost	NCDI	10.6%	12.7%	15.3%
Annualized Operating cost	OPP	54.3%	48.5%	41.2%
Annualized Mainten. Cost	MNT	12.7%	11.4%	9.6%
Annualized Cost Total	AAP	100.0%	100.0%	100.0%

Table 11-10 Recycle Case: Sensitivity to Cost of Capital at 2x DCI

INPUTS	2x DCI R, k=.047	2x DCI R, k=.07	2x DCI R, k=.10
Depreciable capital investment	$35,240,000	$35,240,000	$35,240,000
Non-Deprec capital investment	11,430,000	11,430,000	11,430,000
Annual operating costs	3,640,000	3,640,000	3,640,000
Annual maintenance costs	1,320,000	1,320,000	1,320,000
System algal yield mt/yr	45450	45450	45450
System Lipid yield bbls/yr	150000	150000	150000
Carbohydrate yield mt/yr			
Protein yield mt/yr			
CALCULATED VALUES			
Cost of capital	0.0475	0.0700	0.1000
Cap Recov Factor (book life)	0.0692	0.0858	0.1102
Cap Recov Factor (depr life)	0.0947	0.1098	0.1315
Fixed Charge Rate	0.0951	0.1202	0.1581
PV of Capital Investment	$43,859,290	$46,747,480	$50,790,780
PV of NonDeprec Investment	14,225,640	15,162,420	16,473,860
PV of Operating costs	73,014,990	57,974,420	44,343,990
PV of Maintenance costs	28,291,940	22,394,550	17,065,560
TOTAL PV OF FACILITY	$159,391,900	$142,278,900	$128,674,200
Annualized Cost Algae Plant	$12,163,810	$13,817,670	$16,608,590
LIPID only price $/bbl	81.09	92.12	110.72
ALGAE price $/mt	267.63	304.02	365.43
Lipid price by weight $/mt	580.05	658.91	792.00
Lipid price by value $/mt	290.03	329.46	396.00
Protein price by weight $/mt	580.05	658.91	792.00
Protein price by value $/mt			
Carbohyd price by weight $/mt	580.05	658.91	792.00
Carbohyd price by value $/mt			

Annualized Capital Cost	DCI	$4,171,168	$5,620,076	$8,028,321
Annualized Capital Cost	NCDI	984,142	1,301,095	1,814,894
Annualized Operating cost	OPP	5,051,241	4,974,813	4,885,293
Annualized Mainten. Cost	MNT	1,957,261	1,921,687	1,880,080
Annualized Cost Total	AAP	12,163,811	13,817,671	16,608,588
Annualized Capital Cost	DCI	34.3%	40.7%	48.3%
Annualized Capital Cost	NCDI	8.1%	9.4%	10.9%
Annualized Operating cost	OPP	41.5%	36.0%	29.4%
Annualized Mainten. Cost	MNT	16.1%	13.9%	11.3%
Annualized Cost Total	AAP	100.0%	100.0%	100.0%

Table 11-11 Recycle Case: Sensitivity to Polymer Dose

INPUTS	R, 1x Poly.	R, 2xPoly.	R, 5xPoly.
Depreciable capital investment	$17,620,000	$17,620,000	$17,620,000
Non-Deprec capital investment	11,430,000	11,430,000	11,430,000
Annual operating costs	3,640,000	4,094,000	5,456,000
Annual maintenance costs	800,000	820,000	820,000
System algal yield mt/yr	45450	45450	45450
System Lipid yield bbls/yr	150000	150000	150000
Carbohydrate yield mt/yr			
Protein yield mt/yr			
CALCULATED VALUES			
Cost of capital	0.0475	0.0475	0.0475
Cap Recov Factor (book life)	0.0692	0.0692	0.0692
Cap Recov Factor (depr life)	0.0947	0.0947	0.0947
Fixed Charge Rate	0.0951	0.0951	0.0951
PV of Capital Investment	$21,929,650	$21,929,650	$21,929,650
PV of NonDeprec Investment	14,225,640	14,225,640	14,225,640
PV of Operating costs	73,014,990	82,121,800	109,442,200
PV of Maintenance costs	17,146,630	17,575,300	17,575,300
TOTAL PV OF FACILITY	$126,316,900	$135,852,400	$163,172,800
Annualized Cost Algae Plant	$9,307,186	$9,966,858	$11,856,910
LIPID only price $/bbl	62.05	66.45	79.05
ALGAE price $/mt	204.78	219.29	260.88
Lipid price by weight $/mt	443.82	475.28	565.41
Lipid price by value $/mt	221.91	237.64	282.71
Protein price by weight $/mt	443.82	475.28	565.41
Protein price by value $/mt			
Carbohyd price by weight $/mt	443.82	475.28	565.41
Carbohyd price by value $/mt			
Annualized Capital Cost DCI	$2,085,584	$2,085,584	$2,085,584
Annualized Capital Cost NCDI	984,142	984,142	984,142
Annualized Operating cost OPP	5,051,241	5,681,258	7,571,307
Annualized Mainten. Cost MNT	1,186,219	1,215,875	1,215,875
Annualized Cost Total AAP	9,307,186	9,966,859	11,856,908
Annualized Capital Cost DCI	22.4%	20.9%	17.6%
Annualized Capital Cost NCDI	10.6%	9.9%	8.3%
Annualized Operating cost OPP	54.3%	57.0%	63.9%
Annualized Mainten. Cost MNT	12.7%	12.2%	10.3%
Annualized Cost Total AAP	100.0%	100.0%	100.0%

SECTION 12.0

EXPERIMENTAL SYSTEM DESIGN AND CONSTRUCTION COSTS

12.1 SITE DESCRIPTION

The site chosen for the experimental system is in Brawley, California in the Imperial Valley on a two thousand acre asparagus farm. Over one thousand acres of land are at present not cultivated. The location of the pond system is on land that borders the southwest side of the Salton Sea, making that water resource available by pipeline or shallow well. The land is substantially flat, with sandy and clay soils and sparse vegetation. Figures 12-1 and 2 show the general location of the site and the specific location of the pond construction area.

The climate in the Imperial Valley is representative of the warmer, drier areas of the American Southwest. The annual rainfall averages 7cm, the net pan A evaporation averages 240cm, and the mean insolation is 480 Langleys/day total. Table 12-1 summarizes the mean monthly and yearly climatic data.

Table 12-1. Climatic Conditions at the Proposed Experimental Site

Month	Rain,cm[14]	Evap.,cm[14]	Lang/[15]	Max T	Min T	Av T ,°C[15]
January	1.0	11.2	275	27	-1	12
February	0.9	10.4	375	29	1	15
March	0.6	17.1	475	32	4	18
April	0.3	21.0	550	37	7	21
May	0.1	27.6	625	40	11	25
June	0.0	30.7	675	44	12	29
July	0.2	31.8	650	45	19	33
August	0.9	27.9	600	45	19	32
September	1.0	24.8	550	43	15	29
October	0.6	19.8	425	38	8	23
November	0.5	11.0	325	32	3	17
December	1.3	9.3	275	24	0	13
Total	7.3	244.4	Av 485			Av 22.5

Although the pond location is adjacent to a developed farm, it is essentially undeveloped. The benefits from the farm will be made available to the experimental system through a rental agreement. These shall include: access to water from the 1700 gpm well, access to power lines, lease of land required, option for expansion of land requirement, and use of office space. The agreed rent is $2000 per month.

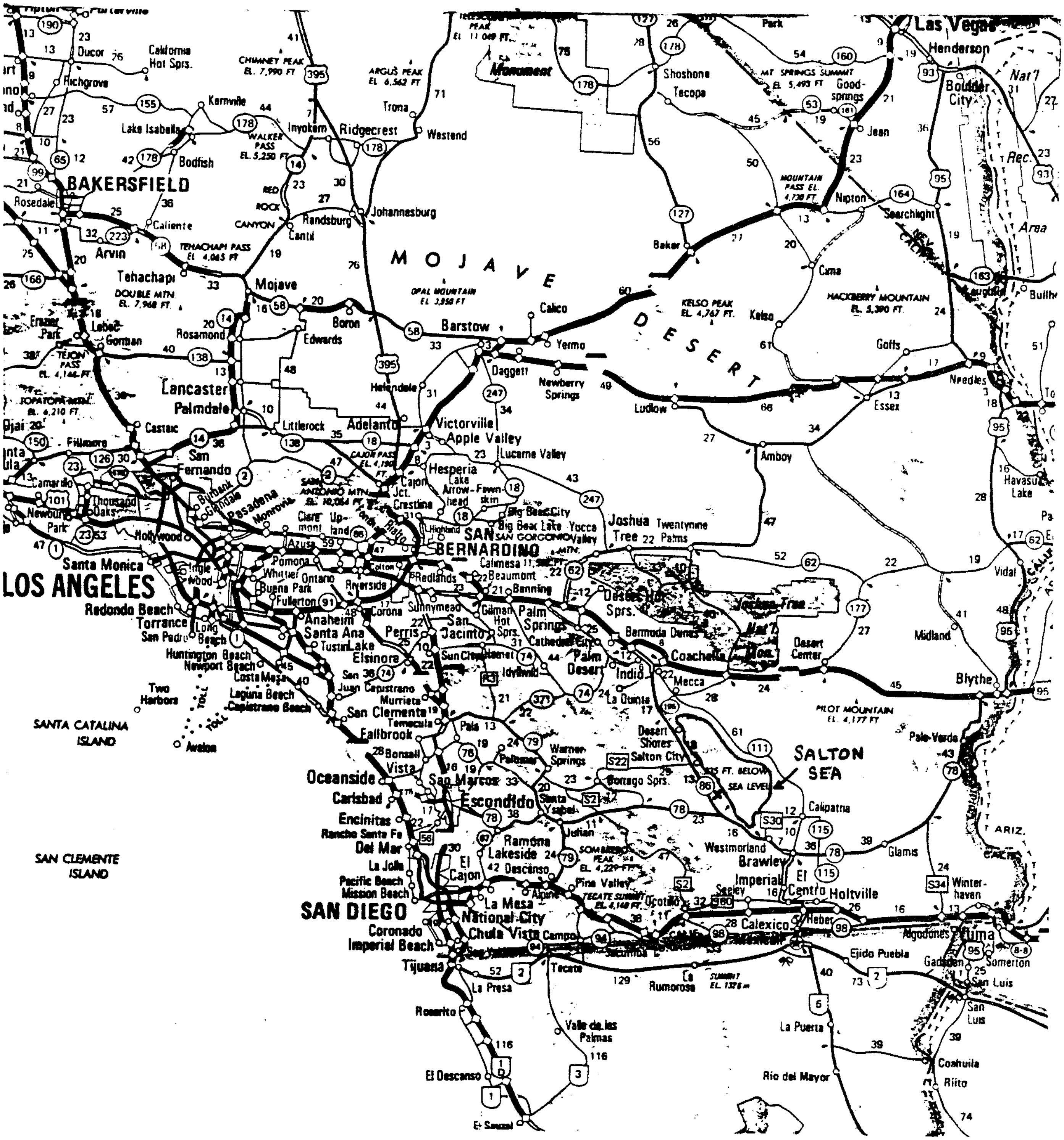


Figure 12-1 Location of Experimental System

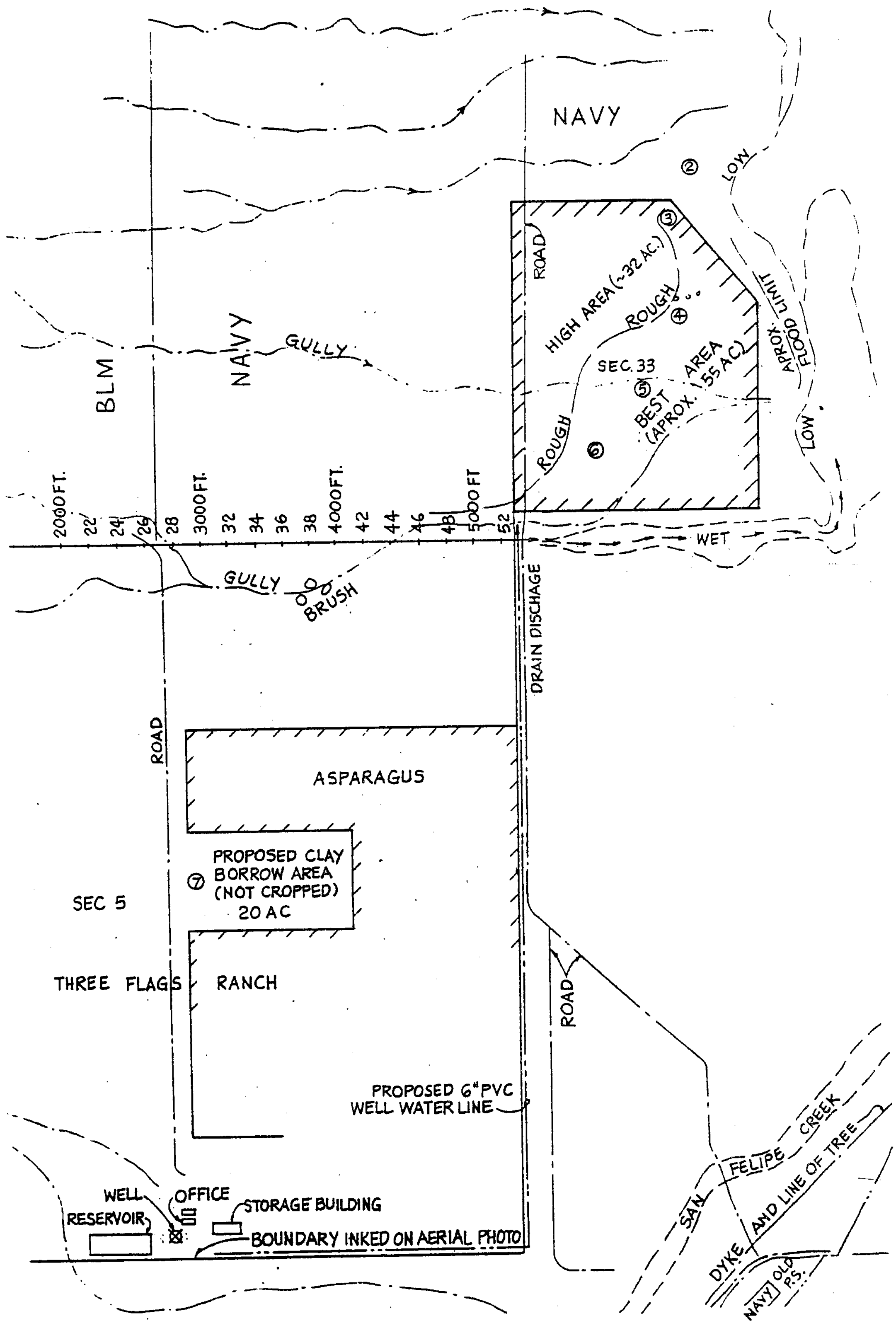


Figure 12-2 Experimental Site

The characteristics of the well water and of Salton Sea water are given in Table 12-2. Both waters are known to support the growth of photosynthetic organisms. The high calcium level in the Salton Sea water necessitates some form of water conditioning. After conditioning, this water can be mixed with the well water to obtain a resource with almost any salinity. The two starting salinities of interest are 4 and 8 ppt, allowing experimental blowdown ratios of .125 and .25 respectively, to attain a medium salinity of 32 ppt. The mixture ratios required to yield the two salinities are, well water to Salton Sea water, 19 to 1 and 4 to 1. In addition, the Saltion Sea water can be used to approximate ocean water.

Table 12-2. Chemical Composition of Experimental Site Water Resources

	Salton Sea[16]	Groundwater
pH	8.6	8.55
TDS, ppt	34.8	2.5
Alkalinity, meq/l	2.2	3.2
Ca, ppm	1089	40
Mg, ppm	1220	50
K, ppm	75	20
Na, ppm	970	800
Cl, ppm	14700	1200
SO_4, ppm	7800	100

12.2 EXPERIMENTAL SYSTEM DESCRIPTION AND JUSTIFICATION

Aside from laboratory and other support systems, the experimental system consists of six 1.5 m^2 ponds, three 50 m^2 ponds, two 4000 m^2 ponds (one acre each) and, if budget allows, one 4 hectare pond (10 acres). The goal of the experimental system is to demonstrate, on a scale that can be extrapolated with a large degree of confidence to a full scale system, the designs and operational performance goals of the full scale system. The 4 hectare pond is the minimum size that is still close enough to full module size to allow a direct transfer of results to a larger system. In that sense it is the optimal choice for this project. It is, however, costly. Thus the experimental system, without the 4 hectare pond, has been designed to test all of the full scale system parameters. A greater degree of extrapolation is needed, of course, to transfer the results. Especially in the cases of mixing and carbonation, the larger pond is desirable. However, the .4 hectare ponds can be operated to simulate, in head loss and carbon storage time, the larger ponds. All of the final engineering results must be obtained from the .4 or 4 hectare ponds.

It is not the goal of this experiment to encompass the activities of the other work carried on in the ASP. Species screening in general, biological adaptation studies, and development of resource analyses are all functions that are beyond the scope of this experiment. However, whenever an outdoor system is operated there is an abundance of laboratory culture work and small scale test work that must accompany the demonstration experiments. Each site has its own environmental input parameters which must be studied. Each organism obtained from a screening program responds differently to these specific site inputs. The 1.5 m^2 and 50 m^2 ponds specified are for use in these support studies. A large number of the former are included because they are easy to operate and manipulate, and also inexpensive. Being outdoors they are more realistic a test system that lab cultures. The 50 m^2 ponds bridge the gap between the small ponds and the real demonstration ponds. They do this literally by providing an intermediate volume for inoculum build-up. In addition, less expensive mock-ups of engineering designs can be tested in these ponds, yielding results which serve in the evaluation of the overall feasibility of a design and thus input into the decision whether to go ahead and test at the larger scale. Two of the 50m^2 ponds are for testing basic design parameters, two being the minimum number required for controlled, simultaneous experimentation. The third is to be used to vary subsystem designs, e.g., airlift vs. paddlewheel mixing. More detail is given in Section 13 on the use of each of the pond systems.

12.3 ALTERNATIVE SITES

Two back up sites have been identified for the proposed experiment. Neither is as optimal, overall, as the primary site in Brawley, Ca., but each has specific advantages. One of the sites is also located in the Imperial Valley, at Salton City. It borders the Salton Sea as well. The advantages of this site are that it has more presently constructed laboratory and shop space and, most importantly, an inexpensive source of CO_2 from a well. The disadvantages are that the supply of low TDS water is significantly more expensive and that the expansion potential is limited to under 8 hectares. Microbial Products, Inc. has an informal option to use this site. The second back up site is located in the area of Roswell, New Mexico. No specific site has been selected, however conversations with personnel at the New Mexico Solar Energy Institute indicate that many sites are available. The Institute and other state agencies, as well as private concerns are potentially interested in co-funding an algae based experimental and/or demonstration system. This is one of the primary advantages of this site. Water, of varying salinity is plentiful, laboratory and culture facilities are available for lease, and CO_2 pipelines are nearby. Thus this site would appear to be the least expensive at which to construct a facility. However, the climate is not as warm, especially in late fall and early spring, as the Imperial Valley of California. This could limit the length of the growing season. Although this may be a realistic condition for much of the Southwest, an experimental facility produces data more cost-effectively when the growing season is long. In any case, the design of the proposed system could be easily adapted to either of the alternative sites.

12.4 EXPERIMENTAL SYSTEM DESIGN

Since a primary goal of the experimental system is to validate the design concepts of the large scale system, its design will closely parallel that of the ponds and harvesting systems presented in Sections 8 & 9. Obviously, only a limited number of the more promising and important design options outlined in those chapters will be constructed, and the operation of the experimental system will focus on resolving the most important technical issues.

The proposed experimental site was described in the previous section. A more detailed view of the actual pond site is shown in Figure 12-3. The large ponds (5 & 10 acres) shown are for reference only, and are not included in the baseline experimental proposal. The proposed two 0.4 hectare ponds would be located in the lower middle portion of the site, as indicated on the figure. A section view of the site is shown in Figure 12-4, with approximate elevations above the Salton Sea.

In the sub-sections which follow, each element of the experimental system is described, along with its cost, starting with the two 0.4 hectare growth ponds. In most cases, the unit costs are significantly higher than in the 192 hectare system, reflecting the 240-fold difference in total system size. Design parameters for the single pond and the two pond system are listed in Tables 12-3 & 12-4.

12.4.1 0.4 Hectare Growth Ponds: Design and Cost

Earthworks

Rough grading costs were estimated conservatively pending a more detailed survey of the pond site. A local earthworks contractor quoted a price of \$.55/yd^3 for cutting and filling. An allowance was also made for possible diversion of the shallow gulley shown on the bottom of Figure 12-3, as well as for initial site clearing. A total of \$5000 was budgeted for this category. This should also cover the additional area required for the site buildings and the settling pond. Lazar levelling will cost about \$3000 total. The \$2000 finish grading cost is actually less than in the large scale design because the liner costs (one membrane lined, one crushed rock) are listed separately. \$900 is budgeted for sump excavation.

Walls and Structural

The cracking/seepage issue discussed in Section 7.3.3. will be examined by building the divider wall and a portion of the outside wall in the unlined pond out of poured concrete. The outside wall, which has the full 20 cm of head across, it is more likely to indicate a problem, if any exists. If so, remedies are available (e.g. covering the inside of the wall with a membrane)

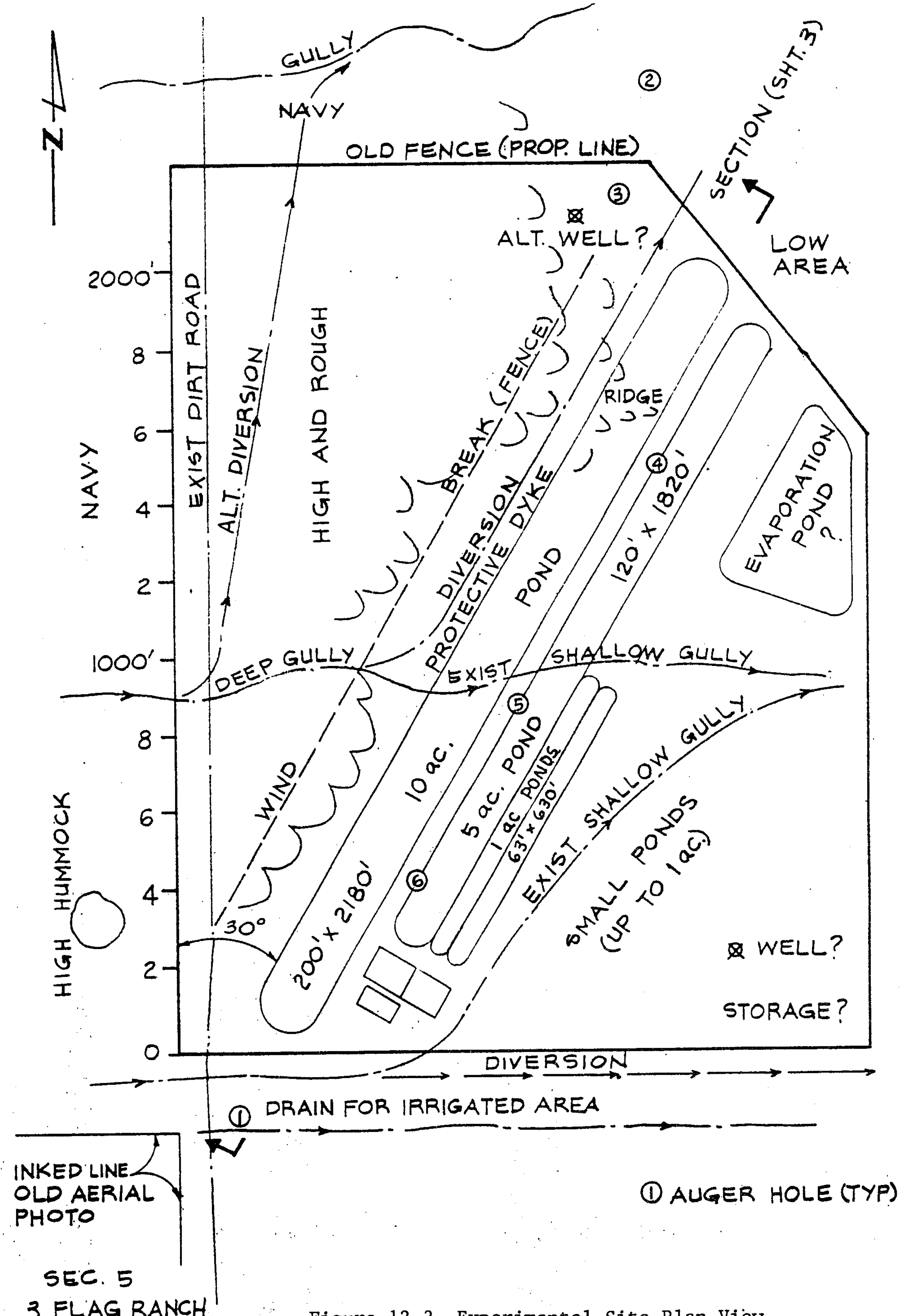


Figure 12-3 Experimental Site Plan View

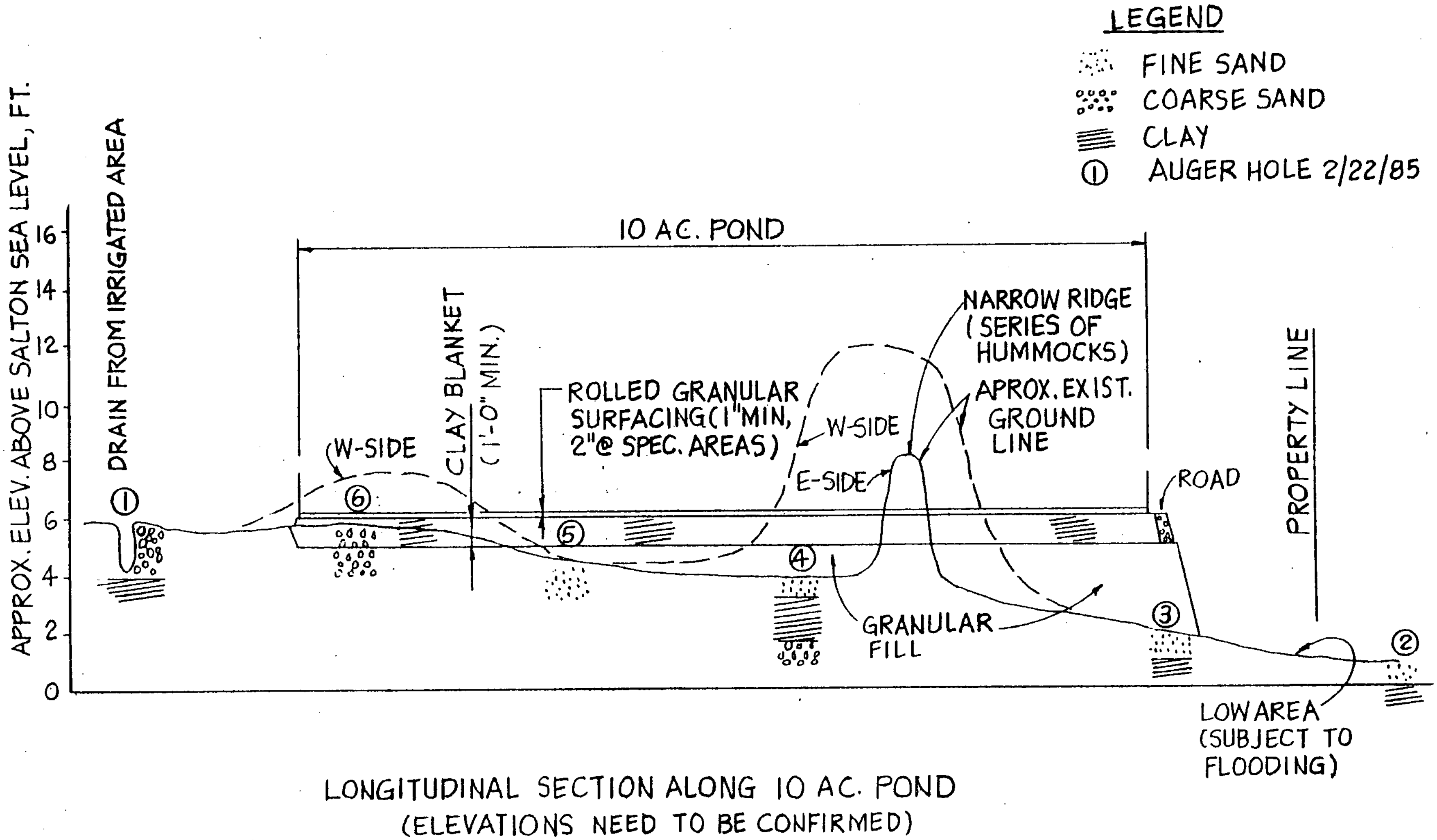


Figure 12-4 Experimental Site Section View

Table 12-3 0.4 Hectare Pond Design Calculations

INPUTS

DESCRIPTION	NAME	VALUE
Pond Area	AREA	0.4 hectares
# of Channels	#CHAN	2 -
L/W Ratio	L/W	20 -
Depth	DEPTH	20 cm
Channel Velocity	VEL	30 cm/sec
Mannings 'n'	MANNING	0.018 sec/m^0.33
Paddle Eff.	PW.EFF	0.6 -
Drive Eff.	DR.EFF	0.7 -
PW Width/Chan Width	PAD/CHAN	0.75 -
Evaporative Rate (max)	EVAP	1.50 cm/day
Blowdown Rate (max)	BLOWDOWN	0.21 cm/day
Detention Time	DET.TIME	4.00 days
Wall Ht. (above grade)	WALL.HT.A	40 cm
Wall Ht. (below grade)	WALL.HT.B	10 cm
Sump Depth	SUMP.DEPTH	1.5 meters

OUTPUTS

DESCRIPTION	NAME	S.I. UNITS	ENGLISH UNITS
Channel Width	CHAN.WID	9.69 meters	31.78 feet
Paddle Width	PAD.WID	7.26 meters	23.83 feet
Centerwall Length	CW.LEN	193.71 meters	635.52 feet
Single Channel Length	CHAN.LEN	208.92 meters	685.43 feet
Total Channel Length	TOT.CHAN.LEN	417.84 meters	1370.86 feet
Total Wall Height	TOT.WALL.HT	0.50 meters	
Slope	SLOPE	2.492 x 10^-4	2.492 x 10^-4
Total Head Loss	HEAD.LOSS	11.33 cm	4.46 inches
Total Efficiency	TOT.EFF	42%	42%
Hydraulic Power	HYD.PWR	594 watts	0.796 hp
Total Power	TOT.PWR	1413 watts	1.894 hp
Total Unit Power	TOT.UNIT.PWR	0.35 watts/sq m	0.35 watts/sq m
Velocity at Paddle	VEL.PAD	40.0 cm/sec	1.31 ft/sec
Pond Volume	VOLUME	809.4 cu meters	28584 cu ft
Evap. Flowrate	Q.EVAP	60.7 cu m/day	2144 cu ft/day
(alternate units)		42.2 liters/min	11.1 gpm
Blowdown Flowrate	Q.BLOWDOWN	8.5 cu m/day	300 cu ft/day
(alternate units)		5.9 liters/min	1.6 gpm
Harvest Flowrate	Q.HARVEST	202.4 cu m/day	7146 cu ft/day
(alternate units)		140.5 liters/min	37.1 gpm

Table 12-4 0.8 Hectare Pond System Design Calculations

(uses input from above, as well as the following)

INPUTS

DESCRIPTION	NAME	VALUE
# of Adjacent Ponds	#ADJ.PONDS	2
Sets of Adj. Ponds	#SETS	1

OUTPUTS All outputs refer to the entire pond system

DESCRIPTION	NAME	S.I. UNITS	ENGLISH UNITS
Total # of Ponds	#PONDS	2	
Total Pond System Area	S.AREA	0.8094 hectares	2.000027 acres
(alternate units)		8094 sq meters	87121.38 sq ft
Center Wall Length	S.CW.LEN	387 meters	1271 feet
Tot.Straight Wall Length	S.STR.W.LEN	969 meters	3178 feet
Curved Wall Length	S.CUR.W.LEN	122 meters	399 feet
Total Wall Length	S.TOT.W.LEN	1090 meters	3577 feet
Total Wall Area	S.TOT.W.AREA	545 sq meters	5868 sq ft
System Pond Volume	S.VOLUME	1619 cu meters	57167 cu ft
Evap. Flowrate	S.Q.EVAP	121.4 cu m/day	4288 cu ft/day
(alternate units)		84.3 liters/min	22.3 gpm
Blowdown Flowrate	S.Q.BLOWDOWN	17.0 cu m/day	600 cu ft/day
(alternate units)		11.8 liters/min	3.1 gpm
Harvest Flowrate	S.Q.HARVEST	404.7 cu m/day	14292 cu ft/day
(alternate units)		281.0 liters/min	74.2 gpm

to keep the pond operational. The issue is moot for the membrane lined pond since the membrane will extend up to the top of the wall. The curved walls will be built using Dodd's corrugated panel technique. The issue of existing patents on the flow deflector vanes will be examined, and if favorably resolved, vanes will be installed at one end of the ponds. The total cost of walls and flow deflectors is $25,100.

The sumps are identical to those described for the large scale system in section 7.3.4. Rails and piers will be installed installed ($3400), and $5000 is budgeted for solids removal equipment, which will not be built until the severity of the problem is evaluated.

Mixing System

Paddle wheels will be built, incorporating the features discussed in Section 7.3.6. The paddle wheel design specifications and costs are included in Appendix III. Since these are experimental ponds, the design is based on a maximum velocity of 30 cm/sec. The cost is $24,400 for the two mixing systems.

Carbonation

Carbonation efficiency is a key technical issue to be studied in the experimental system. The 0.4 hectare ponds, with their full-scale sized sumps, will provide data which can be extrapolated to larger ponds with confidence. The cost of the two carbonation systems is $7000, which includes the pH controller and associated instrumentation.

Lining

One of the ponds will be lined with a good quality membrane liner (Hypalon or CPE), while the other will be lined with a 0.3 meter (1 ft) clay base topped with a thin layer of crushed rock, rolled to a smooth finish. The clay is necessary because of the sandy soil at the pond site, and is available on-site for the cost of hauling. The crushed rock liner is an important cost saving feature of the large scale system which needs to be evaluated. The relatively long channels will allow accurate head loss measurements to be made, from which roughness coefficients can be determined.

Summary

A summary of the construction costs for the two 0.4 hectare experimental ponds is presented in Table 12-5.

Table 12-5 0.8 Hectare Experimental System Pond Costs

INSTALLED COST FOR 0.8 HECTARE SYSTEM

2 PONDS 0.4 HECTARES EACH

DESCRIPTION	QUAN	UNITS	UNIT $	TOT $
GROWTH PONDS				
Earthworks				
Rough Grading*	2.0	acre	$2,500	$5,000
Lazar Levelling	2.0	acre	1,500	3,000
Finish Grading	2.0	acre	1,000	2,000
Sump Excavation	362	cu yd	2.5	904
Walls & Structural				
Straight Walls	5213	sq ft	4.0	20,851
Curved walls	655	sq ft	5.0	3,275
Flow Deflectors	197	sq ft	5.0	983
Sump bottom	36.7	cu yd	150	5,500
Sump ends	8.6	cu yd	250	2,159
Rails & Piers	254	ft	13.5	3,431
Solids Remover				5,000
Mixing System				
Paddle Wheels	2	# PW	10,700	21,400
P.W. Structural	2	# ponds	900	1,800
P.W. Depression	6	cu yd	200	1,239
Carbonation System	2	# Ponds	3,500	7,000
Lining (membrane)				
Channels	43730	sq ft	0.55	24,051
Ends	6865	sq ft	0.60	4,119
Lining				
Clay	1874	cu yd	0.75	1,405
Crushed Rock	319	cu yd	9.00	2,867
GROWTH PONDS TOTAL				$115,985

* Includes drainage diversion & site clearing

12.4.2 50 sq meter Experimental Ponds

Three 50 m^2 experimental growth ponds will be constructed for biological and small scale engineering studies. Two of these will be mixed with paddle wheels, one with an air lift. All three ponds will be lined. The cost breakdown for these ponds is shown in Table 12-6.

Table 12-6 50 m^2 Pond Costs

Site Peparation (levelling, etc):	$1000
Walls: 430 ft @ $8/ft	3440
Sumps: 3 x $500	1500
Mixing Systems (2 P.W., 1 air lift):	5250
Carbonation: 3 x $2000	6000
Submersible Pumps: 2 x $300	600
Drains, piping: 3 x $800	2400
Misc.:	1500
TOTAL, 3 Ponds	$23,040

12.4.3 Experimental System Harvesting Design

Settling Pond

In order to validate the settling pond concept, a settling pond sized for the volume of a 0.4 hectare growth pond will be constructed. Rather than building a scaled down version of the large settling ponds of Section 8, the experimental settling pond will be an almost full scale version of one of the channels in the larger pond. The depth, width, and length will be approximately 80% of full size. An identical 2% bottom slope will be used. This pond will allow meaningful evaluation of settling characteristics and proposed concentrate collection system of the large pond. The side walls will be made as steep as possible (possibly lined with gunnite instead of a membrane) in order to reduce "edge effects." Two 20,000 liter tanks, with sloped bottoms, will be used for secondary settling, flush storage, and for primary settling tests with the 50 m^2 ponds. Figure 12-5 shows the proposed settling pond system, while Table 12-7 lists its cost.

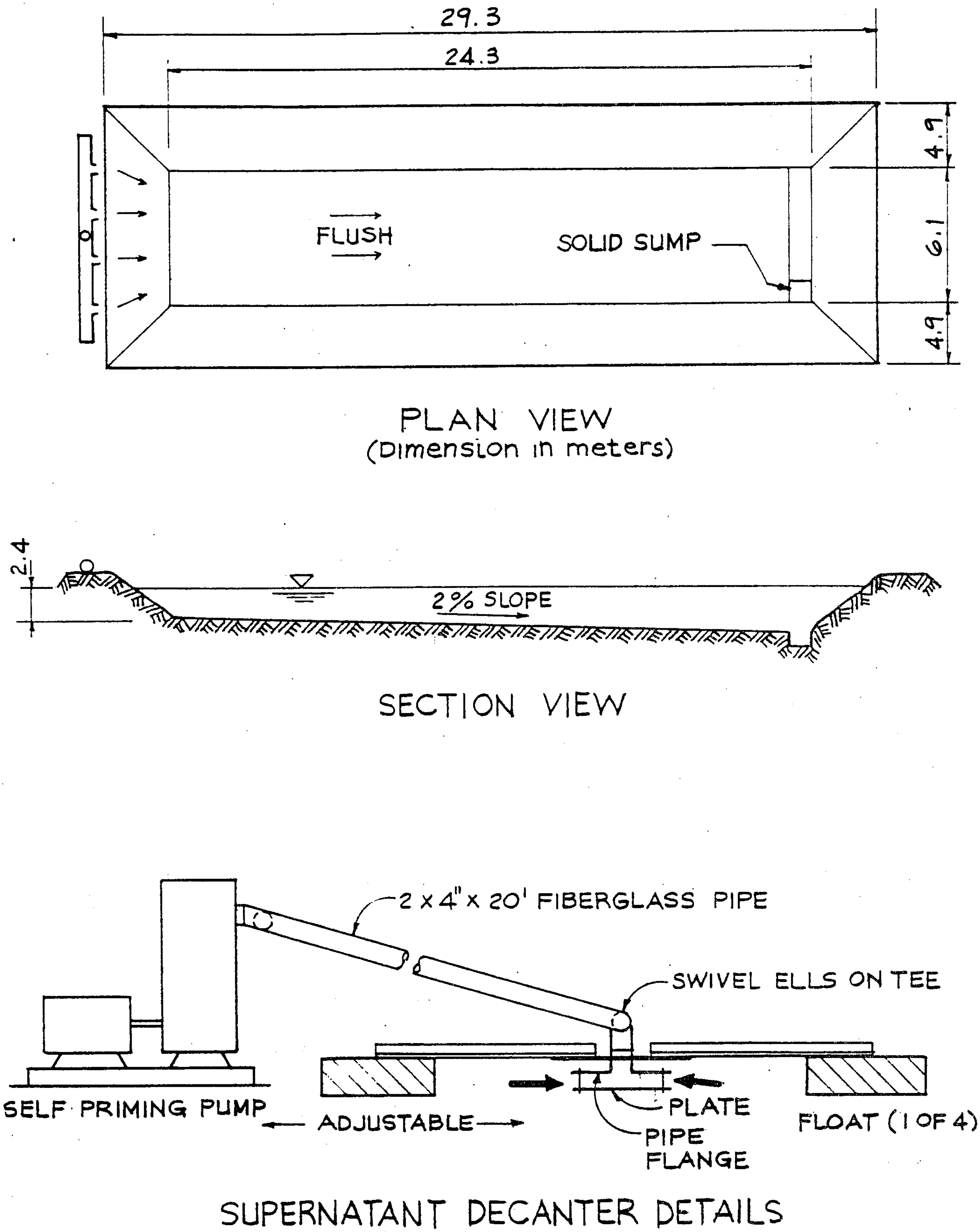


Figure 12-5 Experimental System Settling Pond

Table 12-7 Experimental System Settling Pond Cost

Excavation & shaping: 1000 yd^3 @ \$4.00/yd	\$4000
Lining: 5060 ft^2 @ \$0.65/ft	3300
Pumps & accesories:	
Supernatant, centrifugal 500 gpm	5000
Concentrate, solids handling 100 gpm	2200
Supernatant decanter (fabricated)	1000
Flush storage/secondary settling tanks 2 x 20,000 liters	7000
All piping & valves	3500
Instrumentation & misc.:	3000
TOTAL	\$29,000

Alternative Harvesting Design

A total of \$30,000 has been budgeted for the development of an alternative harvesting method, which would be either belt filtration or air/DO floatation. The belt filter was discussed in Section 8. Although a pilot scale unit would probably cost more than \$30,000, there a possibility of purchasing the unit built by Dodd in Singapore. The other alternative is to develop a air/DO floatation separator, which operates in much the same fashion as a dissolved air floatation, but uses fine bubble spargers in conjunction with the release of dissolve oxygen to float the algae to the surface. Low doses of polymer are used to enhance coagulation. A budjet of \$30,000 should be sufficient to develop a reasonable sized experimental unit.

Centrifuge

An additional \$30,000 is budgeted for the purchase of pilot scale centrifuge, since centrifugation is likely to follow any primary harvesting process. A small horizontal solid bowl unit can be purchased for this amount, from which the necessary scale-up data data can be compiled. Alternatively, a larger unit, many of which are available on the used equipment market, could be purchased.

12.4.4 System-wide Costs

Anaerobic Lagoon System

An covered anaerobic lagoon will be constructed to test the feasibility of nutrient recycling, and to stabilize algal solids before disposal. The lagoon will significantly reduce the quantity of solids to be dried and disposed of. Although there are no plans for utilization of the methane produced (since suitable engine-generators are quite expensive), the lagoon's performance will be monitored. A rough estimate of $20,000 was provided by the consultant who designed the anaerobic lagoon system for the large scale system, for a lagoon sized to handle the thickened effluent from the 0.4 hectare ponds.

Water Supply

An existing well at the proposed site has ample capacity to provide groundwater to the pond system. The cost of piping this water from the well to the pond site was estimated at $23,500. An additional $10,000 is budgeted to construct a shallow well near the Salton Sea as a source of highly saline water. With these two sources, a wide range of salinities can be formulated.

Water Storage

A lined water storage pond will be constructed as part of the water supply system. The pond will also serve as a settling pond for water treatment. The design and dimensions of the water storage pond is shown in Table 12-8.

Water and Carbon Dioxide Distribution, Nutrient Supply

Most of the piping for the experimental system is contained within a small area at the head of the growth ponds. $1500 is budgeted for water distribution to the experimental ponds and settling pond. The CO_2 will come from a rented storage tank. The carbonation system itself is included in the pond cost (Table 12-5). Most nutrients are added in salt form directly into the ponds, or in some cases, are dissolved in a small mix tank and pumped. Nitrogen will be added as aqua-ammonia from a storage tank (these are normally rented). A total of $3500 is budgeted for mix tanks, piping and valves associated with CO_2 distribution and nutrient additions.

Buildings

The offices, lab, and culture rooms will be located in rented trailers, with the exception of a small "lab annex", which contains balances and other

Table 12-8 Experimental System Water Storage Pond

STORAGE POND DESIGN - Volume of Truncated Prism

Slope= 2.0 (Run/Rise)
Depth= 8 ft (Do not include freeboard)

L1	L2	R	A1	A2	Am	Volume (cu ft)	Volume (gallons)
90	58	16	8100	3364	5476	44491	332790

STORAGE POND DIMENSIONS - Including freeboard

From table above:

Slope= 2 (run/rise)
Depth= 8 ft
L1 = 90 ft

Additional design parameters:

Desired freeboard = 1 ft
Addl. liner beyond top of berm = 2 ft

Computations:

		SINGLE LINER	
Length at top of berm	=	94	ft
Length of bottom	=	58	ft
Area of bottom	=	3364	sq ft
Area of bottom liner	=	3364	sq ft
Side panel height, to waterline	=	16	ft
Side panel height, to top of berm	=	18	ft
Side panel height, total	=	20	ft
Side panel area (based on total)	=	1520	sq ft
Area of all four side panels	=	6080	sq ft
Total liner area	=	9444	sq ft

COST:

Excavation & berms:	862 cu yd @ $3/cu yd =	2587
Liner:	9444 sq ft @$.60/sq ft =	5666
Misc.:		300
TOTAL		$8,554

instruments that must be on more solid ground. Additional buildings include a shop and centrifuge/chemical storage shed at the pond site. The cost of buildings is shown in Table 12-9.

Table 12-9 Experimental System Buildings Cost

Lab annex: concrete floor, air conditioned, 350 ft^2 @ \$30/ft^2	\$10,500
Shop: concrete floor, 1000ft^2 @ \$15/ft^2	15,000
Centrifuge shed, concrete floor, one side open, 500 ft^2 @ \$8/ft^2	4,000
TOTAL	\$29,500

Roads

A small amount (\$2500) is budgeted to extend an existing dirt road to the proposed pond site.

Instrumentation

This category covers the cost of instrumentation not already included in other categories, (e.g. solar monitor, pond temperature and level sensors, power measuring equipment, etc) and for a microcomputer-based data aquisition system. Ultimately, the system could be used to automate various pond operations, as most of the package data aquisition systems have digital output lines that can be used to open valves, turn on pumps, etc. A total of \$15,000 is budgeted for this category.

Electrical Distribution

The cost of electrical distribution and wiring is normally estimated as a percentage of total capital costs. For a relatively small project such as the proposed experimental system, 5% is typical.

Machinery

A total of \$15,000 is budgeted for machinery, to cover the cost of a pick-up truck, shop tools and equipment.

Engineering

The engineering budget, 5% of capital costs, or about $20,000 is to cover the cost of any outside engineering (e.g. surveying, drafting, soils analysis) required for the construction of the ponds. Most of the engineering will be done in-house, and is included in the operating cost budget in the following section.

Contingency

A 10% contingency is included to cover unforseen costs and unusual circumstances.

12.4.5 Summary of Construction Costs

A summary of all construction costs for the proposed experimental facility is given in Table 12-10 The two 0.4 hectare growth ponds account for about 25% of the total, the small ponds about 7%, the harvesting systems about 20%, system-wide costs about 33%, engineering 5%, and contingency 10%.

Table 12-10 Experimental Facility Construction Cost Summary

	TOTAL $
GROWTH PONDS: 2 x 1 acre	-------
Earthworks*	$10,904
Walls & Structural	41,199
Mixing System	24,439
Carbonation System	7,000
Lining (1 membrane, 1 crushed rock)	32,443
GROWTH PONDS - 3 x 50 sq meter	23,000
- 6 x 1.5 sq meter	9,000
HARVESTING	
Settling Pond System	29,000
Air/DO Floatation	30,000
Centrifuge	30,000
SYSTEM-WIDE COSTS	
Anaerobic Lagoon System	20,000
Water Supply (groundwater)	23,500
Water Supply (Salton Sea water)	10,000
Water Storage Pond	8,600
Water Distribution	1,500
CO2 & Nutrient Supply	3,500
Buildings	29,500
Roads	2,500
Instrumentation not included elsewhere	15,000
Electrical Distribution (5% of above)	17,554
Machinery	15,000
ENGINEERING (5% of above)	19,182
CONTINGENCY (10% of above)	40,282

TOTAL COST	$443,103
	========

SECTION 13.0

EXPERIMENTAL SYSTEM OPERATION AND COSTS

13.1 INTRODUCTION

The work plan and work schedule are presented below for the construction and operation of the experimental system. The plan is broken down into individual tasks, each describing an essential aspect of construction or operation. The schedule follows, with each task assigned a time span for completion. The cost of labor, equipment, supplies, and overhead is then delineated. The final budget, including construction and operating costs, follows this section

13.2 SYSTEM CONSTRUCTION

13.2.1 Design Review and Finalization - Task 1

The design, and any changes made during the review process, must undergo a final review for insertion of detail and finalization of the construction schedule. This work would actually be done primarily before a contract start date and continue during the first month.

13.2.2 Construction Contract Negotiations and Contract Execution - Task 2

Concurrent with Task 1, during the first month of the project, construction contractors will be sought and interviewed. After the solicitation of bids and negotiations, a final construction schedule will be drawn up and any variation from the initial budget reviewed.

13.2.3 System Construction - Task 3

Contingent on the weather, the construction of the entire proposed system would take four months. The acquisition of trailors for office space, the construction of storage and utility buildings, and earthworks will be done during the first of these four months. After laying water and power lines, the system itself will be constructed.

13.2.4 Laboratory Setup - Task 4

During the eighteen month contract period, the laboratory will be set up primarily in trailors. Acquisition of trailors and installation of laboratory equipment will begin once the site utilities are in place and the earthwork completed. Some of the laboratory equipment will have to be housed in a small pre-fabricated building on a concrete slab for stability. As soon as possible thereafter, a laboratory technician will be hired to initiate the algal culture collection and prepare for growth of inoculum. The 1.5 m^2 ponds will be operational at this time.

13.3 EXPERIMENTAL PLAN

13.3.1 Water Resource Experiments - Task 5

The water resource is an important parameter of the overall system. Since it is yet of rather undefined composition, the experimental system must be flexible enough to test a wide variety of water resources. The two water resources available at the proposed site provide this flexibility when combined with water conditioning. The Salton Sea provides the sodium chloride needed to achieve any conceivably useful TDS. It can be diluted with the well water to achieve any intermediate salinity allowing the system to be operated at almost any blow down ratio. The calcium concentration is very high in the Salton Sea, so that some water softening is required. The magnesium level is also high. Conditioning will be done with soda ash for calcium removal, and with lime and soda ash for removal of both. Input stream splitting will allow adjustment of both to different degrees. Still, the waters composed in this manner may only approximate specified water resource characteristics. It should be noted that these specifications are only averages so that precise duplication is not realistic. However, the small ponds, 1.5 m^2, and laboratory cultures can be used to formulate and test any given water composition. As of now, the water resources of interest are Type II concentrated eight fold and seawater. The following experiments will be performed regarding water chemistry and algal growth on these waters.

A. Water Conditioning
   1. Cost of conditioning vs amount removed and efficiency
   2. Precipitate removal by settlng
   3. Precipitate formation, after initial conditioning, upon additional concentration

B. Water Chemistry

1. Identification of species precipitating and remaining ion composition, by outside analytic laboratory

2. Approximate determination of inorganic carbon reaction equilibrium constants by inflection points in titrations. Determination of carbon storage capacity as a function of pH change.

C. Controlled Tests of Algal Growth on Waters Used

1. Bottle tests in laboratory, with Type II water as control for the growth of the lipid containing strains like Chryso/F1 and Cheatoceros

2. Growth tests in 1.5 m^2 ponds with Type II water as control for Chryso/F1 (or other promising lipid producer) and seawater as control for growth of Platymonas.

D. Media Recycle

1. Preliminary media recycle experiments will be conducted in the 1.5 amd 50 m^2 ponds to determine the minimum blow down ratio that is feasible. These experiments will be done for one month at a timne, which is short in terms of what must be actually practiced in a real system, but long enough to assess feasibility.

2. All experiments in ponds of .4 hectare size will be performed with media recycling. This is necessary for minimizing the operational cost of these experiments and crucial for assessment of the utiliization of saline groundwaters in any practical system which is to be operated in the arid areas of the U.S. Southwest. Productivity and species dominance as a function of the duration of cultivation on recycled medium are the main factors to be analyzed.

13.3.2 Carbonation System Tests - Task 6

Although in-pond sumps were specified for carbonation in the large scale design, both sumps and covered area carbonators will be tested in the experimental system. Sumps, however, will be provided in the larger experimental ponds, being the primary method. Covers will be added later if the results from 50 m^2 ponds are positive. The carbonation tests will be extensive and will be done prior to inoculation in both freshwater and media. The specific tests follow.

A. Sump Carbonation

1. Stripping rate per unit depth of sump
   - a. versus sump liquid depth
   - b. versus the ratio of gas flowrate to liquid flowrate
   - c. versus sparger type (initial bubble size) and gas flowrate

2. Overall transfer efficiency in 1.5 m deep sump
    a. versus sparger type
    b. for co-current, lateral, and counter-current flow
3. Performance of gas recycle, if necessary

B. Covered area carbonator
1. Overall transfer efficiency for given area of cover
2. Effect of water velocity under cover
3. Efficiency of ripple cover

C. Effect of algae and supersaturated oxygen on transfer efficiency
1. Measure CO_2 transfer and O_2 out
2. Determine if algae excretion products enhance transfer

Dissolved oxygen can be measured using an Orion DO probe and checked with the Winkler method. Carbon uptake can be measured by calculating the change in total inorganic carbon from pH change and the carbon equilibrium equations.

13.3.3 Mixing System Tests - Task 7

Although many paddlewheel systems are now used for mixing algal cultures, there is little data on the actual efficiency obtained in large ponds. Less data is available concerning airlift mixing systems for low head, low velocity systems. Thus it will be important to determine just how well these different options perform. Since the primary system to be installed in the experimental ponds is the paddlewheel, airlift mixing will be tested initially by sparging air into the carbonation sumps adapted for the purpose by installation of a baffle. Measurements will include head loss and various power numbers as a function of water velocity. In addition, CO_2 outgassing will be measured as a function of mixing speed. If the parameters determined for airlift mixing are favorable, a draft tube system for airlift will be designed and installed in a 50 m^2 pond for testing. Scale-up to .4 ha will depend on the evaluation of results and budget constraints.

The parameters determined from the carbonation tests and the airlift mixing tests will be used to evaluate and design a combined gaslift and carbonation system. The calculations will be performed for a flue gas system (high gas flow rates), a diluted CO_2 gas phase (35% CO_2, at three times the flow rate of a pure CO_2 system), and for pure CO_2 (low gas flow case).

13.3.4 Growth and Productivity Studies - Task 8

Studies on the growth and productivity of the algae will be performed in the 1.5 m^2 and 50 m^2 ponds. The best conditions will be scaled up to the .4 ha ponds for validation of the full scale design. Both indoor bottle cultures and the 1.5 m^2 cultures provide a convenient means to test quickly a moderate number of factors and values. Initial work indoors usually indicates whether a factor is important or not, except for those that depend on high light intensity. For these the outdoor small ponds work well. Thus media optimization studies can be performed indoors, while pH and DO work is best done outdoors. Since the mixing regime in a small pond does not simulate that in a larger one, the initial results obtained at the small scale need to be tested, in the short and long term, at the 50 m^2 and .4 size. Factors to be tested for affect on productivity include pH, DO, alkalinity, temperature, batch detention time, continuous vs batch operation, and nitrogen depleted productivity and storage product accumulation. Routine measurements include pH, temperature, DO, insolation, CO_2 input and biomass density. Chemical composition is measured as required to determine lipid productivity. Since correlation of productivity with various parameters is often a useful evaluative tool, much data needs to be taken and stored in an organized manner. A computer automated data acquisition system will be installed for this. In addition, with this in place, the status of productivity in a pond can be determined approximately on-line, without having to wait for cell density measurements, by monitoring the CO_2 input and the pH. When corrected for injection and outgassing losses, CO_2 input is proportional to productivity.

13.3.5 Harvesting Tests - Task 9

Two harvesting methods will be tested in the operation of the experimental system: flotation followed by foam collection and slow primary settling in large settling ponds followed by secondary settling in small tanks. Both methods will be followed by centrifugation to provide a final product at 10%-20% volatile solids. The biomass to be separated will be nitrogen starved and flocculated with small doses of high molecular weight polymers. Polymer flocculation is an important feature of the harvesting process. On site polymer flocculation tests will be conducted with the aid of consultants. The hypothesis being tested is that polymers can be tailored to the specific needs, i.e., chemical medium and algal type, while still maintaining enough versatility to function as a univerally effective flocculant. The collection method, sedimentation or flotation, will be evaluated in terms of flocculant type and dose required to achieve efficient and reliabe separation and concentration. The dose of flocculant must not exceed the equivalent of $.02 per kg dry biomass, and preferably half that. For gas flotation studies, some testing will be done right in the ponds. The density of biomass leaving the flotation station will be used to calcualte the harvesting efficiency. This will be measured as a function of pond DO, biomass density (since several passes through the harvester will be made, each with lower incoming biomass density), CO_2 input, duration of nitrogen depleted growth, and flocculant dose.

Unharvested biomass will re-enter the nitrogen sufficient growth phase. Thus it will be necessary to test strains for the ability to quickly absorb newly introduced nitrogen and enter an active growth stage.

The sedimentation tests will be performed in two phases: cell settling and sedimented slurry removal. The settling rates of the biomass will be measured as detailed previously [1] in indoor tests, with and without flocculant. When the velocity for most of the biomass exceeds 30 cm/hr, the pond will be fed flocculant and pumped into the settling basin. Sedimentation time will be determined on the basis of equivalency to settling 3 m in 10-15 hrs. After decanting the supernatent, the slurry recovery system will be tested for efficiency and rate of slurry removal. Then the rate of thickening in tanks will be determined.

Centrifugation tests are important at this point to determine the throughput and hence the size and number of centrifuges needed at the large scale. Centrifuge vendors are available to test the characteristics of the biomass slurry and consult on centrifuge selection and operation.

Once the harvesting parameters have been determined with biomass from 50 m^2 ponds, the process will need to be validated using the large growth and settling ponds. The latter have been designed to emulate directly the solids removal system specified for the full scale design. The experimental settling pond is a "slice" of the full scale settling pond, having the same characteristic dimension for solids flushing.

13.3.6 Digestion of Biomass in a Covered Lagoon for Nutrient Recycling Tests - Task 10

A small, covered lagoon will be constructed to digest, anaerobiaclly, the biomass harvested from the large ponds. These tests will be somewhat artificial in that the "pretreatment" due to lipid extraction will be absent. The results will still be extremely valuable in acertaining how the biomass components, that do degrade, partition in the lagoon. This information is necessary to validate the assumption upon which the extensive recycling of carbon, nitrogen, and phosphorus was based. The lagoon has been sized to accept the total biomass harvest of a one acre pond. The tests will provide the data necessary for evaluating the system's capability of producing a methane fuel product.

13.3.7 Liner Evaluation - Task 11

A significant capital cost savings incorporated into the large scale design is the use of clay sealer and crushed rock overburden for pond lining. One of the .4 ha ponds, and the 4 ha pond if constructed, will be lined in this manner. The liner must be evaluated, over the longterm, for water loss and buildup of deleterious organic solids. This must be done in a large pond in which the hydraulics are similar to a full scale pond. The other .4 ha pond will be lined with a high quality plastic liner and serve a control, as well as a test of the endurance of such a liner under field conditions. All of the 50 m^2 ponds will be similarly lined.

13.4 OPERATIONS SUMMARY AND EXPERIMENTAL PRIORITIES

The proposed system consists of six 1.5 m^2 ponds, three 50^2 ponds, and two .4 hectare ponds. A 4 hectare pond would yield results directly transferrable to large systems, but it is not cost effective research at this time. The 1.5 m^2 ponds are to used for initial screening of promising algal species under outdoor conditions of temperature and irradiance. Media and growth parameters will be determined in these small ponds. The 50 m^2 ponds will be used to test the various subsystem designs (carbonators, mixing systems, air flotation, sedimentation systems) at a scale which will indicate feasiblilty at lowest experimental cost. These ponds will also be used for growth studies and to provide inoculum for the .4 ha ponds. The latter will serve as the primary testing facilitiy for validation of the large scale design over longterm operation. As far as validation of the large scale dasign, the highest priority experiments are those which determine the utilization efficiency of CO_2, including mass transfer coefficients and losses throughout the system; those which test cost effectiveness of methods for primary concentration of the algal suspension, including polymer flocculation tests and evaluation of collection device performance and cost (settling ponds, flotation devices); use of an inexpensive anaerobic lagoon for recycling nutrients (especially carbon) that would not be removed in the product stream, including the partitioning of these nutrients in recoverable fractions of the lagoon; evaluation of the pond lining alternatives, most importantly the water losses when plastic liners are not used and the endurance of each of the liners; and the effect of longterm media recycling on system performance.

The biomass and lipid productivity goals will be evaulated, but attainment of these goals is substantially dependent upon provision of strains pre-screened for the ability to produce lipid. A preliminary choice of strains include the high lipid producing Chryso/F1, a productive and potentially high lipid producing strain of Chaetoceros, and the Hawaiian Platymonas for use as a biomass productivity standard. The possibility of removing photosynthetically produced oxygen from the pond medium will be examined--no determinationn is possible prior to experimentation--and the effects of such removal on productivity will be studied.

Two important growth characteristics of the organisms will be scrutinized during this experiment. First is the ability to recover quickly from nitrogen starvation after addition of new nitrogen into the medium. This involves resistance to photo-oxidation and photo-bleaching in already depigmented cells. Second is the ability to maintain high productivity in a medium that has undergone extensive recycling. This usually means that the organism must not excrete organics into the medium in significant quantity.

13.5 EXPERIMENTAL SCHEDULE

Figure 13-1 lists the eleven tasks described above and shows the anticiapted schedule of completion. Many of the tasks are of an ongoing nature, with periodic evaluation. Task 12, data evaluation and reporting, is indicated on the figure.

13.6 BUDGET DESCRIPTION

The complete budget is given at the conclusion of this section. The elements of the budget are described in Tables 13-1,2,3, and 4. Table 13-1 gives the cost of pond equipment. These items were included in the total pond construction costs given in Section 9.0. The costs listed in Table 13-2, are for laboratory equipment. Two cost columns are included in this table. The first gives the cost of acquiring new equipment as specified. Those items which are owned by Microbial Products, Inc. or are government owned but, at present available for use by Microbial Products, Inc. will be made available for use on this project. Thus the second column lists the cost of equipment which must be purchased for this project. The supplies and materials (other than those used in constructing the pond system) are detailed in Table 13-3. The costs shown are based on vendor quotes or on previous experience. At the experimental scale, nutrient costs are high because quantities are relatively small. The breakdown of direct labor costs is given in Table 13-4. The time alotments are in equivalent full months of work, not by scheduled work time. The budget form that follows at the end of this section gives the total project cost for the eighteen month effort, including overhead and fee.

Table 13-1. Experimental System - Pond Equipment

Item	Cost,$
Motors, speed reducers for 1 acre ponds	4800
pH controllers, 2	2000
Flowmeters, 5	1450
Motors, controllers, small ponds	1500
Pumps,5	8100
Centrifuge, harvest	30000
Air/DO, harvest (partial)	est. 5000
Computer, interfaces	15000
Total	67850

Task	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
1.Design Review	X																	
2.Budget Review	X																	
3.System Construct.	X	X	X	X														
4.Lab. Set up			X	X														
5.Water Res. Exp't.																		
Water Cond.				X	X	X												
Water Chem.			X	X	X	X												
Growth tests				X	X	X	X	X	X									
Media recycle					X	X	X	X	X	X	X	X	X	X	X	X	X	X
6.Carbonation Tests																		
Sump						X	X		X	X	X							
Covered area						X	X											
Media Effects						X	X		X	X	X							
7.Mixing Tests																		
$50m^2$ ponds								X	X									
1 acre ponds										X	X	X	X					
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
8.Growth Studies			X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
9.Harvesting Tests																		
Jar tests			X	X	X	X	X	X	X	X	X	X	X	X	X	X	X	X
$50m^2$									X	X	X	X	X					
1 acre										X	X	X	X	X	X	X	X	X
10.Digestion-Recycle									X	X	X	X	X	X	X	X	X	X
11.Liner Inspection									X				X					X
12.Reporting																		
Design review	X																	
Budget review		X																
Construct.progress			X															
Construct.final					X													
Semi-annual report							X						X					
Final Report																		X

Figure 13-1. Experimental Schedule

Table 13-2. Laboratory Equipment

Item	Availability	Acquisition Cost,$	Budgeted Cost,$
Air conditioning			
Culture room	P[1]	1500	1500
Buildings	P	2000	2000
Culturing equip.			
Light banks	M[2]	1500	--
Gas mixers	M	800	--
Shaker	M	700	--
Inoc. hood	M	3000	--
Air compressor	M	500	--
Tables	P	500	500
Laboratory equip.			
Fume hood	M	3000	--
Centrifuge	G[3]	12000	--
Autoclave, used	M	15000	--
Steam generator	M	2000	--
pH meters, 2	G	2500	--
Oven, drying	M	600	--
Furnace, muffle	M	800	--
Balance	M	5000	--
Microscope	G	3000	--
Oven, vacuum	M	500	--
Cell disrupter	G	4000	--
Spectrophotometer	M	10000	--
Furniture	P	2000	2000
Total		70,900	6000

1 Purchase required
2 Owned by Microbial Products
3 Government equipment

Table 13-3. Operating Materials and Supplies

Item	Basis, $	Total Cost, $
Culturing supplies	500 Monthly, 16 mo.	8000
Lab. analysis	500 Monthly, 16 mo.	8000
Water analysis, outside lab.	200-500/sample, 10-20 samples	5000
Polymer consultant	Quote	10000
Pond Chemicals		
CO_2	85/ton, 200 tons/yr,quote	17000
CO_2 vessel rent	30 ton capacity, 750/mo.,quote	13500
Ammonia	500/ton, 20 tons/yr, list pr.	10000
Phosphate	2000/ton, 20 tons/yr list pr.	4000
Iron	4500/ton, 20 tons/yr list pr.	9000
Sodium carbonate	300/ton, 20 ton/yr list pr.	6000
Pond utilities	500/mo., 18 mo.	9000
	Total	99500

Table 13-4. Direct Labor
Eighteen Month Project

Title	Equiv. Months	$/mo.	Total, $
Research Scientist	7	3750	26250
Research Engineer 1	8	3750	30000
Engineer	3	3750	11250
Lab. Tech.1	18	1667	30000
Lab. Tech.2	6	1667	10000
Pond Operator	12	2083	25000
		Total	132500

BUDGET

CATEGORY	COSTS,$
Materials and Supplies	
Pond construction (Table 12-9 minus $67,850 for equipment included below	375,253
Operating (Table 13-3)	99,500
Special Equipment	
Pond Equipment (Table 13-1)	67,850
Laboratory (Table 13-2)	6,000
Direct Labor (Table 13-4)	132,500
Total Direct Cost	681,103
Indirect Cost (1.318 x Direct Labor)	174,635
Total Direct plus Indirect	855,738
Fee (7% of total direct plus indirect)	59,901
TOTAL Project Cost	915,639

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APPENDIX I

CARBON DIOXIDE TRANSFER

A.1 INTRODUCTION

Transferring carbon to algae ponds is a major design problem. Unlike typical fermentations, which may have volumetric production rates of 15 g biomass/L-hr in deep, well-stirred tanks, algae ponds have 200-1000 times lower peak volumetric production rates in shallow, open ponds. Thus, large volumes of water must be transported to carbonation sites where CO_2 is injected. In the plug-flow regime of high rate ponds, enough CO_2 must be injected, and subsequently stored in the water to meet the carbon demand of the algae, as well as any CO_2 losses due to outgassing as the water travels between carbonation stations. Storage capacity is a function of depth, alkalinity and pH change. The alkalinity can be supplied by the water source, and concentrated by water evaporation from the pond, to a level determined by the blow-down ratio. The allowable pH changes are determined by biological tolerance of the desired microalgae population and water resource characteristics. Many inexpensive water sources contain hardness which tends to precipitate. The extent of precipitation depends on pH and alkalinity, i.e., carbonate concentration. Thus, the variables of water chemistry, blowdown rate, depth, pH changes, and C storage potential are all interrelated. CO_2 losses due to outgassing depend on pH, alkalinity and carbonator spacing, as well as mixing speed.

AI.2 PHYSICAL ABSORPTION OF CO_2

Gas-liquid mass transfer is a subject which has received much attention in the chemical and biochemical engineering literature. Theories and experimental correlations are well developed. However, they give only approximate and/or relative estimates of transfer rates. In the final analysis, rates must be empirically derived from prototypes and pilots. This is especially true of open channel carbonation, where irregular surface deformations occur due to turbulence, wind and bubbles. These disturbances may well have significant effect on gas transfer in ponds. The following analysis quantifies some of the more pertinent aspects of algal pond carbonation and evaluates, relatively, the sensitivities to depth and mixing intensity.

The theories of gas-liquid transfer into agitated liquids all involve some simplified manner of incorporating the hydrodynamics into the basic diffusional models. The rationale is that at some scale, close to the gas-liquid interface, turbulent eddy diffusion is damped to the extent that molecular diffusion is responsible for transfer. In the thin film model [1, 2] film thickness parameter, d, incorporates (in a mostly unspecified manner) all of the hydrodynamics, yielding a transfer rate equal to (D/d) (C^*-Co), where D is the molecular diffusion coefficient of the solute gas in cm^2/s,

$C*$ is the molar concentration of solute gas that would be in equilibrium with the partial pressure of solute in the gas phase, in mmoles/cm^3 and Co is the molar concentration of solute gas in the bulk liquid. The mass transfer coefficient, k_L is equal to D/d . k_L is defined from the empirically observed relationship $VdCo/dt = k_L a\ (C*-Co)$. Here dCo/dt is the transfer rate, mmoles/cm^3.sec, V is the volume in cm^3, a is the interfacial area in cm^2, K_L is in cm/sec. The intensity of turbulence determines the film thickness, but the latter is difficult to infer theoretically and hard to measure.

In other models, the liquid turbulence is viewed as causing replacement of small interfacial surface areas with liquid from the uniformly mixed bulk. The type of distribution of surface ages assumed determines the quantitative prediction of these models. Higbee's surface renewal model assumes all elements of area have a uniform lifetime [3]. A convenient model, and the one used below, is Danckwert's renewal model [4] in which the probability of any element of surface being replaced is proportional to the age of the surface element. Then the surface renewal is a Poisson process, having many calculational conveniences.

In this approximation, gas side resistance to diffusion is considered negligible, so that at the gas liquid interface equilibrium does exist between the bulk gas partial pressure of solute and the concentration of solute in the liquid at this interface. The relationship between gas side partial pressure and liquid side concentration is given by Henry's solubility law, with Henry's constant appropriately decreased for elevated temperatures and ionic strength and composition of the liquid. Especially for anything but small gas phase partial pressures of CO_2, where diffusion of solute molecules to the interface actually limit transfer, the equilibrium relationship is valid.

Once an element has been "renewed," transfer is assumed to go as in an infinitely deep, stagnant liquid. All transfer properties are determined as averages over the entire interfacial area, using the Poisson distribution as the weighting function. Purely physical transfer (without chemical reaction) goes as $(DS)^{1/2}(C*-Co)$, where S is the rate of surface renewal, or 1/S is the average age of any surface element after formation. Several other models are available, however, there is little empirical evidence for any one in particular. In most circumstances, all models give similar quantitative predictions and, in addition, relative prediction as to the effect of chemical reactions on gas absorption. As alluded to above, in many instances, the model chosen for use depends on calculational ease in the specific case.

The renewal rate, S, embodies all of the effects of turbulence on gas-liquid transfer. It is not possible to measure, directly, this idealized parameter, but its value can be inferred from experiments. Flow of water in flumes, streams, and sewers is closely related to that in open channels. The empirical formula developed for k_L as a function of slope, velocity and depth from sewer data (5) has been found to correlate (6) with numerous sets of data taken for flumes, streams and open channels. As indicated in Table AI-1, Danckwert's model, in which $k_L = (DS)^{1/2}$, was used to calculate S from k_L based on purely physical absorption. Although the original correlations were made for aeration, S is independent of the diffusion process and can thus be determined. The procedure is to (1) choose a channel roughness coefficient and use Manning's formula to determine slope from given depth and velocity values, (2) insert the depth, velocity, and slope values into the Parkhurst-Pomeroy (5) correlation to determine k_L, and 3) use Danckwert's model to determine S from k_L and DO_2. Table AI-2 shows values of k_L (for oxygen, for CO_2 multiply by (1.91/2.28)) and renewal rate calculated according to this procedure with n = .025.

TABLE AI-1. MASS TRANSFER COEFFICIENTS FOR FLOWING WATER

Basis: $S_1 = (1.49)^{-2}(d)^{-4/3}(nV)^2$

$$k_L = (1.69 \times 10^{-2})\ (1 + 0.17V^2/gd)\ (S_1V)^{0.375} \qquad (6)$$

$$k_L \propto (d^{-.5})\ (V^{1.125})$$

$$S = k_L{}^2/Do \quad (4)$$

Where: S_1 = slope

d = depth, ft.

V = linear flow velocity, ft/sec.

g = gravitational acceleration, 32.2 ft/sec^2

k_L = transfer coefficient, cm/sec measured at 20°C

Do = diffusivity of oxygen = 2.28 x 20^{-5} cm^2/sec at 20°C

S = average rate of surface renewal, sec^{-1}

The purpose of this procedure is to get approximate values of carbonation coefficients using data from aeration experiments and develop a means of, in theory, determining the effect of chemical reaction on carbonation relative to purely physical absorption. The first objective could be met without calculating renewal rates. The results would again depend on the particular model chosen since k D^m where m is 1 or 0.5, or a value in between. Since relative molecular diffusivities are not substantially different, model discrimination is not possible based on experimental determination of m (7). However, this also means that, for an approximate analysis, the result is substantially independent of the model basis. The second objective is conveniently attained using the renewal rate when the chemical reaction is first-order or considered pseudo-first order.

AI.3 CHEMICAL ENHANCEMENT OF CO_2 ABSORPTION

The concentration gradient of a gaseous solute diffusing into a liquid can be steeper if the solute reacts with components of the liquid phase. The magnitude of this effect depends on the physical absorption time constant, the reaction rate constants, and the concentrations of reactants. Absorption of CO_2 into alkaline waters may be accelerated by one of two major uncatalyzed reaction paths, the hydration of CO_2 and subsequent acid-base reaction to form bicarbonate ion and the direct reaction of CO_2 with the hydroxyl ion to form bicarbonate. The rate of the former reaction is faster at pH values below 8, while the second dominates above pH 10. Between 8 and 10 both can be important. Several investigators (7) have calculated the magnitude of the enhancement of CO_2 absorption rates caused by these chemical reactions. Danckwert's renewal model is particularly convenient when the reaction is first order or pseudo-first order. Although the reaction of CO_2 and OH ion is, in general, second order, under some circumstances the OH ion concentration may be considered constant during that time interval between renewal of a surface element, when stagnant diffusion conditions are assumed to prevail. Thus if the rate of surface renewal is high enough, relative to the influx of CO_2 and subsequent depletion of OH in the surface layer, so that OH is not significantly depleted, the pseudo-first order conditions prevail.

When the renewal time is long, the absorption rate of CO_2 is high and/or the concentration of OH is low the reaction proceeds quickly, depleting the OH concentration just below the interface. When these conditions prevail, the rate of absorption is nearly the same as if the reaction were instantaneous. The instantaneous reaction gives the maximal enhancement for the given conditions. The enhancement factors are pH dependent because the concentration of reactant, OH, depends on pH. The alkalinity affects OH concentration by buffering pH changes upon absorption of CO_2. Thus, although the reaction rate is governed by the diffusion of CO_2 down from the interface and diffusion of dissolved OH from the bulk liquid, in certain pH ranges the diffusion of many more molecules of CO_2 (from the saturated interface) is needed than of OH from the bulk, due to the buffer action. For example, at pH 9, over 100 molecules of CO_2 must diffuse downward for each molecule of OH actually depleted from the reaction zone.

TABLE AI-2. CO_2 MASS TRANSFER COEFFICIENTS AS A FUNCTION OF POND HYDRAULICS

Depth	Velocity ft/s	(cm/s)	Slope	k_L cm/sec	S, sec^{-1}	t,sec
.164 ft (5cm)	.16	(5)	8.2×10^{-5}	2.5×10^{-4}	2.7×10^{-3}	370
	.33	(10)	3.4×10^{-4}	5.4×10^{-4}	1.3×10^{-2}	78
	.50	(15)	7.8×10^{-4}	8.9×10^{-4}	3.5×10^{-2}	29
	.66	(20)	1.3×10^{-3}	1.2×10^{-3}	6.3×10^{-2}	16
	1.0	(30)	3.0×10^{-3}	1.9×10^{-3}	1.6×10^{-1}	6
.328 ft (10cm)	.16	(5)	3.0×10^{-5}	1.7×10^{-4}	1.3×10^{-3}	770
	.33	(10)	1.3×10^{-4}	3.9×10^{-4}	6.7×10^{-3}	150
	.50	(15)	3.1×10^{-4}	6.3×10^{-4}	1.7×10^{-2}	59
	.66	(20)	5.5×10^{-4}	8.7×10^{-4}	$3.3 \times 1-^{-2}$	30
	1.0	(30)	1.2×10^{-3}	1.4×10^{-3}	8.6×10^{-2}	12
1.0 (30cm)	.16	(5)	7.0×10^{-6}	1.0×10^{-4}	4.4×10^{-4}	2270
	.33	(10)	3.0×10^{-5}	2.2×10^{-4}	2.1×10^{-4}	476
	.50	(15)	7.0×10^{-5}	3.6×10^{-4}	4.8×10^{-3}	208
	.66	(20)	1.2×10^{-4}	4.9×10^{-4}	1.0×10^{-2}	100
	1.0	(30)	2.7×10^{-4}	7.8×10^{-4}	2.7×10^{-2}	37

* For CO_2 at 20°C this overestimates k_L by 10%. However, since diffusion coefficients are significantly greater at higher temperatures more accurate estimation is not called for.

The hydration reaction, on the other hand, differs in that it is not pH dependent at the surface (below the surface it is, which limits the reverse reaction), and it is first order in CO_2. Thus, enhancement due to this reaction is essentially the same at all pH values between 7 and 11. Table AI-3 gives the forward and reverse reaction rates, and Table AI-4 the concentrations of reactant, at the surface and in the bulk at pH 7.5 and 10.3. All rate constants are for 20°C and infinite dilution, giving lower limits to enhancement. Reverse reactions are negligible.

TABLE AI-3. CHEMICAL ENHANCEMENT FACTORS (9) AND REACTION RATE CONSTANTS

Enhancement of First Order Reactions: $(1 + k/S)^{1/2}$

Enhancement of Instantaneous Second Order Reaction: $1 + D_{OH}(OH)_B/zD_{CO2}(CO_2)^*$

Reactions: $H_2O + CO_2 \underset{k_2}{\overset{k_1}{\rightleftharpoons}} H_2CO_3 \overset{fast}{\rightleftharpoons} H^+ + HCO_3^-$

$CO_2 + OH \underset{k_4}{\overset{k_3}{\rightleftharpoons}} HCO_3^-$

k - 1st order reaction rate constant or pseudo first order constant

B - Bulk liquid concentration

* - Saturated surface concentration

Z - Effective stoichiometric coefficients

Constants: $k_1 = 0.02 \text{ sec}^{-1}$

$k_2 = 20 \text{ sec}^{-1}$

$k_3 = 10^4 \text{ L mole}^{-1} \text{ sec}^{-1}$

$k_4 = 2 \times 10^{-4} \text{ sec}^{-1}$

All 20°C, infinite dilution

Considering the second order reaction of CO_2 and OH at pH 10.3, the parameters which determine which regime is appropriatae (pseudo-first order or instantaneous), are (1) the rate of CO_2 absorption, (2) the ratio of OH to dissolved CO_2 at the surface, and (3) the ratio of the time any element of fluid remains at the surface (the turnover time = inverse of the renewal rate) to the maximal rate of the pseudo-first order reaction, k_3(OH). With 1 atm CO_2 in the gas phase, the physical absorption rate is .01 x 10^{-3} mmoles/cm^2/sec., or after 200 sec. of quiescence of an element of fluid .1 cm deep, 20 mmoles CO_2/L is absorbed. This amount of absorption overwhelms the buffer, causing a drop in OH concentrations. Thus, the pseudo-first order approximation does not apply. Similar calculations for other pH ranges Table AI-4) confirm this over the entire pH range of interest.

The effective stoichiometry of the reaction is pH dependent, as shown in Table AI-4. This stoichiometry, along with the bulk concentration of OH and the saturated level of CO_2 determines the maximal enhancement of an instantaneous reaction regime. Even though the stoichiometry can become quite high, the concurrently falling OH concentration keeps the maximal enhancement from going above 1.05. Even this is not achieved due to the finite reaction rate constant of 10^4 L/mole-sec. Thus the direct reaction of CO_2 and OH does not serve to enhance the reaction when the turnover time is on the order of 200 sec.

Even when the turnover time is only 6 sec., there is no appreciable chemical enhancement. Around pH 9.5 the reaction is close to first order because the amount of CO_2 absorbed during short contact time can be buffered by the solution, keeping the OH concentration relatively constant. However, this concentration is low, keeping the reaction rate low relative to the rate of physical absorption. Increases in absorption rates under these conditions is due almost entirely to the hydrodynamic enhancement discussed previously.

The other reaction path, the hydration of dissolved CO_2, is a first order reaction which can never be considered instantaneous. According to any of the models the enhancement factor is 2-3. Again if hydrualic conditions are chosen to incrase the renewal rate, and hence the physical absorption rate, then chemical enhancement is reduced. For example, with a 6 sec vs. a 200 sec turnover time, physical absorption rates increase 6-fold, but chemical enhancement decreases by almost the same factor, giving overall only a 12% increase in absorption rates. This is a consequence of the low value of the reaction rate constant, 0.02 sec^{-1}. The enhancement varies as $(1 + k/s)^{1/2}$ so at higher values of S, the reaction competes poorly with turnover in absorbing CO_2.

Since even with 30 meq/L of alkalinity the buffer capacity is not great enough to enhance the absorption rates, the net effect of chemical reactions is the same for the low pH, low alkalinity case. Here again, due to the first order reaction of CO_2 and water, rates are enhanced 2-3 fold at a slow turnover rate (corresponding to a 1 foot deep pond mixed at 0.5 fps) and about 12% at high turnover rates (a pond depth of 1/3 ft. and linear velocity of 1.5 fps under the cover).

TABLE A-I-4 CHANGES IN TOTAL CARBON CONCENTRATION IN THE SURFACE LAYER[1]

pH	H^+ M	OH M	α_0	α_1	α_2	Alk meq/L	C_T mM	pH Change	ΔC_T M x 10^3	OH	$\frac{1}{Z} - \frac{\Delta CO_2}{\Delta OH}$
10.3	5×10^{-11}	$.2 \times 10^{-3}$	4.7×10^{-5}	.472	.528	30	19.5	10.3-9.3	7.73	$.18 \times 10^{-3}$	43
9.3	5×10^{-10}	$.2 \times 10^{-4}$	8.98×10^{-4}	.899	.101	30	27.2				
9.0	10^{-9}	10^{-5}	1.89×10^{-3}	.945	.053	30	28.5	9.3-9.0	1.33	$.01 \times 10^{-3}$	133
8.3	5×10^{-9}	$.2 \times 10^{-5}$	9.8×10^{-3}	.979	.011	30	30.0	9.0-8.3	1.43	$.008 \times 10^{-3}$	179
7.5	3.16×10^{-8}	3.16×10^{-7}	.059	.94	----	6	6.38				
7.0	10^{-7}	10^{-7}	.167	.83	----	6	7.23	7.5-7.0	0.85	2.16×10^{-7}	4000

1 Surface film assumed to be 0.1 cm

α_0 CO_2/C_T

α_1 HCO_3/C_T

α_2 CO_3/C_T

Chemical enhancement can therefore be expected to lower areal carbonator coverage by 50-67% when that coverage was initially calculated to be high. If measures are taken to increase physical absorption rates under a cover, then the 80% reduction in coverage thereby gained will not be much more affected by chemical reactions. In sumps chemical enhancement is also not effective under pond conditions, unless the reactions are catalyzed.

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APPENDIX AII

by D. C. Augenstein

GASLIFT PUMPS FOR COMBINED PUMPING AND GAS CONTACTING IN ALGAL PRODUCTION SYSTEMS

AII.1 INTRODUCTION

Gaslift pumps have been proposed to fulfill the simultaneous needs for pumping and gas transfer in microalgal growth systems. This treatment examines, first, the factors governing pumping efficiency, then factors governing the gas transfer function. Gas transfer requirements dictate the most important design parameters. Likely performance of a gaslift meeting both pumping and gas transfer requirements is then examined.

Prior to presenting detail, some important conclusions can be summarized here. With a configuration guaranteeing adequate (>80%) fractional transfer of CO_2 from stack gas, energy available to pump liquid meets or exceeds pumping needs anticipated for high rate pond systems with channel velocities in the 15-30 cm/sec range. A good match between pumping capacity and system needs seems possible with several plausible combinations of other parameters. When CO_2 containing stack gas is sparged to supply CO_2 needs, it seems that energy requirements will be high but probably tolerable function - 5 to 25% - of system output. However, when transfer of pure CO_2 is considered, a gaslift configuration guaranteeing adequate (>90%) fractional CO_2 transfer seems unable to meet pumping needs. Energy demand for pure CO_2 sparging would be less than 3% of system output. Representative pumping efficiencies are in the range of 20-50% and optimum sump depths range between 5-20 meters. It appears that if power plant stack gas is to be considered for CO_2 source, more energy efficient transfer methods through gaslifts might be considered. Major remaining uncertainties affecting this treatment concern the channel roughness, n, which will acatually be encountered, and the absorption coefficient, Kg, for CO_2 in the pond liquid. Other details are presented in the report, which follows.

AII.2 GENERAL DESCRIPTION OF AIRLIFT CONTACTOR/PUMP

The schematic of an airlift system of the type envisioned for application in an algal growth system is shown in Figure AII-1. In the figure, a ducting system (A) brings liquid to the base of the unit. There, at the base of the vertical draft tube (B) gas is introduced through a sparger to create a dispersion of gas in liquid. Because of the lower effective density of the gas/liquid dispersion relative to the liquid, it rises through the vertical Section B as shown and is discharged at an elevated head (the pumping head) relative to the inlet. Necessary gas-liquid contacting (to be discussed later) is also accomplished in the vertical section B. In practice, in an algal growth system with wide channels, the gaslift configuration would be

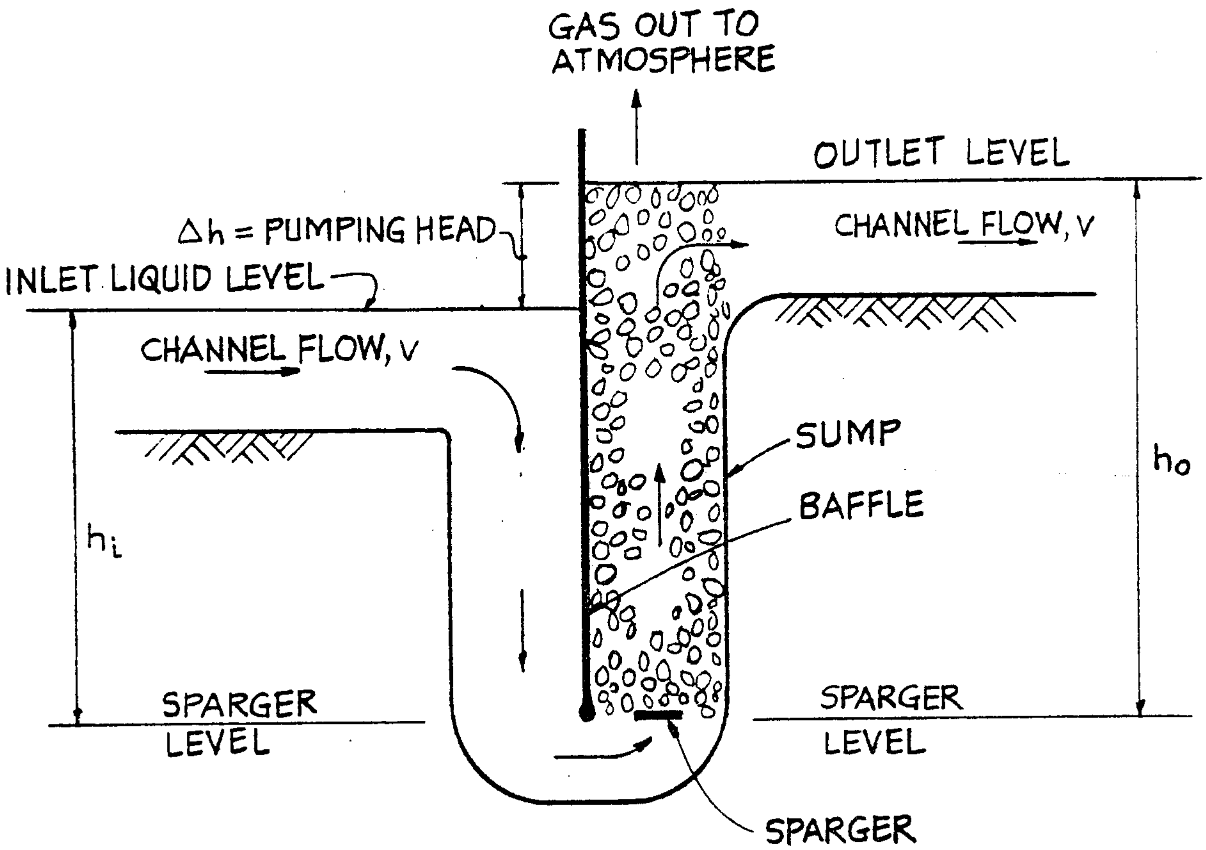

Figure AII-1 Gas Contacting Sump

one in which the necessary number of airlift units, side by side, span the channel, or a unit with cross section of Figure AII-1 (rectangular) spanning the channel, or even, just one draft tube. Analyses to follow are general to all these possible configurations. The advantage claimed for airlift use - relative to independently pumping at one location in the system and sparging gas in at another - is that it provides both pumping and gas contacting functions in one unit operation.

A11.3 GENERAL BACKGROUND

Recent chemical abstracts (1980-83) contained no relevent references on gaslift use solely for pumping liquids. Further communications with Dr. R. T. Hatch (who has extensively researched and published on airlift fermenters) resulted in the finding that the only recent use of gaslifts known to him has been in handling of liquid radioactive wastes in nuclear fuel processing plants, and in collecting ("vacuuming") manganese nodules from the sea floor. In essence, gaslift pumping has been used recently only when either its high reliability (due to no moving parts) and/or its ability to handle, unattended and remotely, very abrasive slurries such as manganese nodules at great depths is required. It seems - although no rigorous analysis is done here - that alternative pumping methods are in most cases reliable enough and more efficient and economical than the combined cost of the compressor/draft tube required for the gaslift approach. It is worth commenting, however, that before the development of improved materials and pumps, airlifts were used extensively in applications such as primary sewage, oilwell, and coal mine (acid water) pumping. References to these applications date to the early 1900's [2,3]. For reasons stated, and also because optimizing gas transfer (treated later) gives operating conditions outside the range for existing correlations [4,5] it will be necessary later to treat airlift pumping based on first principles.

A11.4 GENERAL ASPECTS OF PAST INVESTIGATIONS: PUMPING AND CONTACTING

Because of extensive pilot and commercial application of airlift fermenters, numerous articles are available discussing aspects of their operation - for example references 6-12. Pumping efficiency in airlift fermenters - which determine cycle time around the flow loop consisting of the annulus and draft tube in such fermenters - is almost always adequate and hence pumping has not been a concern. It is, however, addressed in reference 8. A major topic for analysis in the cited references 6-13 has been the ability of the unit to transfer oxygen efficiently. This transfer capacity relates directly to the ability of the unit to transfer CO_2, which is the concern with algal growth systems. Central to both gas transfer and pumping is the behavior of the gas-liquid dispersion in the updraft section, and, in particular, the behavior of the bubble swarm generated within the upflow section. As bubble behavior is critical in both these respects, this topic is addressed first.

A11.5 BEHAVIOR OF BUBBLE SWARMS IN GAS-LIQUID DISPERSIONS

A number of correlations have been developed to describe the behavior of swarms of bubbles which result from sparging of gas into liquid. The treatment to follow uses some simplifying assumptions which are valid for aqueous systems with low surfactant levels. Such low surfactant conditions are expected to exist in most algal growth systems.

Sparging of gas into liquid results in swarms of individual gas bubbles which rise away from the sparger. The size distribution of the bubbles leaving the sparger may initially be a strong function of the sparging method. However, as bubbles of differing size rise through the fluid at differing velocities, coalescence of the smaller bubbles occurs; on the other hand, bubbles above a certain critical diameter, about 0.5 cm in water, will be broken up by fluid forces. The net result of this coalescence of smaller bubbles and breakup of larger ones is that a steady-state bubble size distribution is attained once gas has risen a certain distance above the sparger. Coalescence (and hence the height above the sparger at which equilibrium size distribution is attained) depends strongly on fluid properties, but for fluids similar to water, at gas holdups e>.01, equilibrium would appear to be attained by the time gas had risen above 1 meter [13]. This can be intuitively understood by considering velocities of bubbles of various sizes. (for bubble diameter D_B < 0.2 cm, the Stoke's range, the bubble velocity V_B D_B^2) and the fact that a larger bubble sweeps out a cylindrical locus exceeding its own diameter (D_B). Within that cylinder it will catch up and coalesce with a smaller, slower bubble above it.

The outcome of an example calculation can illustrate the coalescence process. Consider an initial case of two bubbles with a diameter ratio of 2 ($D_{B,1}$ = 0.2 cm and $D_{B,2}$ = 0.1 cm, both in Stokes law range) and an initial vertical separation of 15 cm, with the smaller bubble above the larger. (This circumstance will exist on the average in the case where e = 0.01). The larger bubble will catch up and coalesce with the smaller bubble after only a 25 cm rise. (Note though that higher levels of surfactants and electrolytes can hinder the coalescence process, and the analysis is done only for fluids with properties near water, for example low TDS algal growth systems.) Power dissipation can affect the equilibrium bubble size, but for power dissipations expected in airlift operation this affect has been shown (albeit in other types of fermentations systems) to be negligible [16]. Gas holdup, e, may vary with pressure but any effect on bubble diameter from this is also expected to be negligible, both as pressure effects are negligible and as the coalescence dispersion process tends to maintain the characteristic size distribution.

Rise velocities of bubbles have been established as a function of bubble diameter [15]. Bubbles may be classified into two categories, "rigid sphere" up to D_B 0.2 cm which rise according to Stokes' law, at velocities up to 30 cm/s for D_B = 0.2 cm, and larger bubbles whose gas/liquid interface is fluid and which - over a wide range of sizes, D_B 0.25 to 2.0^+cm - rise at about 25-32 cm/sec [15]. At an equilibrium bubble size distribution discussed most gas, 95+%, is in bubbles with D_B >0.25 cm and a very important consequence of this is that the average rise velocity of gas through liquid, at least at gas holdups up to E = 0.25, is about 28-32 cm/sec. An average bubble rise velocity of 30 cm/sec relative to the liquid will be assumed for analysis to follow. Another assumption is that cross section of flowing liquid in the gaslift - hence the average liquid velocity - is constant. This assumption is to some extent artificial but introduces little error and simplifies the mathematics.

Given a net rise velocity of 30 cm/sec of gas relative to liquid, a constant liquid velocity V_L and a flow rate ratio of gas to liquid of Q_{gc}/Q_L (where Q_{gc} is the gas flow corrected for pressure, temperature and moisture effects - these corrections are treated later), the following relations hold:

$$Q_L = A_L V_L \tag{1}$$

$$Q_{gc} = A_{gc}(V_L + 30) \tag{2}$$

where A_L and A_{gc} are the flowing cross section of liquid and gas, respectively. Gas holdup, e, which can alternatively be defined as the void fraction of gas in the gas-liquid dispersion, is given by

$$e_c = A_{gc}/(A_L + A_{gc}) \tag{3}$$

and, making the necessary substitutions, average holdup can be derived in terms of Q_g/Q_L, and V_L as

$$e = \frac{Q_g Cpe}{Q_L(1 + 30/V_L) + Q_g Cpe} \tag{4}$$

where Q_g is the gas flow, at STP, into the sparger and the term Cpe is a pressure correction factor. Expression (4) will be referred to later.

A11.6 CORRECTION FOR PRESSURE

As noted, pressure must be compensated for in determining gas holdup. If gas flow through the liquid is expressed at 1 atmosphere, then the ratio of actual gas volume and volumetric flow to these values at one atmosphere can be easily computed as a function of pressure and liquid head. An average correction factor, termed Cpe here, can also be computed for the gas stream between the sparger and surface in the draft tube section. This factor, Cpe, can be expressed either in terms of pressure or alternatively in terms of h_i and h_o (refer to Figure AII-1). This correction has been compiled graphically with results shown in Table AII-1. Another assumption is that water vapor and temperature corrections (<5% total) can be neglected. No corrections are in fact necessary if both 1) the temperature of the gaslift fluid equals compressor inlet temperature, and if 2) inlet air to the compressor has the same weight ratio of water to gas as the saturated gas rising through the draft tube.

A11.7 PUMPING BY GASLIFT PUMPS

Refering to Figure AII-1 it is evident that a flow equation can be developed:

$$h_i \rho_L g = h_o \rho_L g(1-e) + (K_C+K_B+K_E)\rho_L V_L{}^2/2 + 4FL\rho_L V_L{}^2/2D \quad (5)$$

where

- h_i = inlet liquid surface elevation above sparger
- h_o = outlet (and h_i - h_i = pumping head)
- ρ_L = liquid density
- g = 980 cm/su^2
- e = holdup (= $e_o C_{pe}$)
- K_C = contraction coefficient of liquid on entering gaslift inlet duct
- K_B = Frictional loss coefficient of liquid in inlet duct-draft tube turn
- K_E = exit coefficient
- f = friction factor (dimensionless)
- L = length of fluid flow path in gaslift (inlet conduit and draft tube)
- D = hydraulic diameter (= 2X width for long rectangular cross section)
- V_L = Fluid velocity in airlift (assumed equal to channel velocity)

It is convenient to define the overall friction coefficient, as KF = (K_L + KB + K_C + 4fL/D). Then equation 5 becomes

$$h_i g = h_o\, g(1-e) + K_F V_L{}^2/2 \quad (6)$$

TABLE AII-1. Computed Values of C_{PE} vs. Depth*

Depth, cm	C_{PE}
100	.96
200	.92
300	.89
400	.86
500	.823
600	.803
700	.775
800	.752
1000	.712
1200	.676
1400	.645
1600	.619

*Gas volume correction factor for depth and pressure. See Text.

It is evident that with values for C_{PE} available equation 6 can be used for various predictive purposes. Discussion of the use of equation 6 will come later after it is demonstrated how gas transfer and pumping requirements fix some of the independent variables.

Before going on, it is interesting to consider likely magnitudes of frictional loss terms within the gaslift due to fluid flow, if gaslifts are applied to algal growth systems channel flow. Velocities proposed for high-rate pond systems range from 10-30 cm/sec [17]. The fluid head increment at an airlift pumping station is likelly to be in the order of 30 to 100 cm, or even more, to provide economical pumping station spacing [17]. It is interesting to note that a 30 cm pumping head is equivalent to 65 velocity heads (= $\rho v^2/2$). It can be readily calculated that if fluid flow in the gaslift is 30 cm/sec, and that if (for one "bad case" example) fluid is both drawn from and existing into stagnant liquid, in which case by Reference 20 K_C = 0.43 and K_L = 1.0 - that the frictional entrance and exit losses are only about 2.4% of the work expended to pumping liquid through the pumping head. The balance of expected frictional losses, from a (possible) 180^o bend between gaslift inlet duct and draft tube, and the total frictional loss from channel flow at L/D = 20, will total less than 3% of the pumping work. Thus for this example total frictional losses in the airlift pump are small, about 5%. Because V_L and V_{ch} can be equal, and a bend configuration can be chosen to minimize frictional loss, frictional losses due to fluid flow should in fact be in the range of 1% of the pumping work expended for cases of interest.

While frictional losses are minor, relative to useful work done in pumping, other inefficiencies exist. One of these, which is major, can be understood intuitively as due to the 30 cm/sec slip velocity of gas relative to liquid discussed earlier. This and other factors relating to efficiency are discussed next.

A11.8 EFFICIENCY CALCULATIONS

A11.8.1 Work Done by Gas on Liquid

The work done by an element of gas on the liquid can be stated as (force exerted by gas on liquid) X (distance moved by liquid during time force is exerted). Note that the distance is not the total moved by the gas element, as the forceXdistance product of the 30 cm/sec gas movement through the liquid is lost, i.e., dissipated as heat. (Also note here that the following analysis is in C.G.S. units, and based on 1 cm^3 gas or viewed by the compressor.)

The average force exerted by 1 cm^3 gas on the liquid is $gC_{pe}\Delta\rho$, with Cpe as calculated earliler, for the time the gas is in the liquid. If gas rises from the sparger to the surface, the time gas is in contact with the liquid is $h_o/(V_L + 30)$. Thus the work done by the 1 cm^3 gas element on the liquid, w_{gL} during the rise of the gas element, is

$$w_{gL} = g\ C_{pe}h_o(V_L/(30+V_L))(T_{gl}/T_c) \quad (7)$$

If gaslift temperature T_{gl} is different from compressor inlet temperature T_C, the term T_{gl}/T_C compensates for this. The term $\Delta\rho$ is assumed to be 1 gm/cm^3 and so is omitted in 7. Other terms were defined earlier.

A11.8.2 Work Expended to Compress Gas

For low head compression, which is the case for algal growth systems, single stage isentropic compression will be applied, and the expression for the work done to pump 1 cm^3 gas, W_{gc}, is given by

$$W_{gc} = nRT_c((p_2/p_1)^{(k-1)/k} - 1)X \quad (8)$$

where

n = mols gas in 1 cm^3 at compressor inlet temperature
T_C = gas temperature entering compressor, ^oK
k = C_p/C_v of gas, = 1.4 for air
P_1 = compressor inlet pressure, ~1 atm (or 10^6 $dyne/cm^2$)
P_2 = compressor outlet pressure
X = polytropic compression factor, 3.5 (dimensionless)

In CGS units, $R = 8.31 \times 10^7$ $erg/mol/^oK$. For 1 cm^3 gas entering the compressor at one atmosphere the product nT_C is constant (whatever the temperature) at $1.218 \times k10^{-2}$ mol ^oK, and with this value (8) becomes, in ergs expended to pump 1 cm^3 (compressor inlet conditions)

$$W_{gc} = 3.546 \times 10^6((p_2/p_1)^{(k-1)/k} \quad (9)$$

Isentropic compression is less efficient than isothermal compression since a portion of the isentropic compression work appears as unusable heat, but it is interesting that efficiency of isentropic compression is still high relative to isothermal at pressures likely to be needed. For example, at a compression ratio of 2/1 (or ca 10M depth for h_i), isentropic compression is 91% as efficient as isothermal.

A11.8.3 Overall Efficiency of Gaslift Pumping (Electricity to Net Pumping Work)

The work done by an element of gas on the liquid has been given previously as equation (7). Given equation (6) and if it is assumed for the moment that channel and gaslift flow velocities are equal then this energy expenditure divides between pumping head and fluid frictional losses in the ratio

$$\frac{(h_o - h_i)g}{(h_o - h_i)g + \frac{K_F V^2}{2}} \qquad (10)$$

The expression for efficiency, $E_{g/p}$, defined as the ratio of pumping work accomplished to gas compression work requirement then becomes

$$E_{g/p} = \frac{c_{pe}\, g\, h_o \left(\frac{V_L}{30+V_L}\right)}{3.546(10^6)\left(\frac{p_2}{p_1}^{\frac{K-1}{K}} - 1\right)} \cdot \frac{(h_o - h_i)g}{(h_o - h_i)g + \frac{K_F V^2}{2}} \qquad (11)$$

The remaining factor to be considered is the motor/compressor efficiency by which 11 is multiplied to give E_o, the overall expression for efficiency:

$$E_o = E_{mc} E_{g/p} \qquad (12)$$

It is of interest to note that overall pumping efficiency E_o should range by equation (12) between about 25% and 50%. One set of typical operating conditions (based on 80% CO_2 transfer need, computations presented later) might consist of a fluid gaslift and channel flow velocity of 15 cm/sec, draft tube height = 5.4 meters, and a pumping head of 14.5 cm. Efficiency under these circumstances at E_{mc} = 0.8 would be about 29.5%. If CO_2 transfer efficiency rises to 95%, and fluid flow velocity is increased to 30 cm/sec, the pump head needed will increase to 1.16 m . With other parameters kept the same, the draft tube height must increase to 13.5 meters and overall efficiency would increase somewhat, to 40.8%. Higher efficiencies are obtainable with higher channel and gaslift flow velocities; however, such higher velocities can result in energy consumption which is a large fraction, easily around 50%, of total expected system output (see later discussion) and in addition, there is no evidence that such flow velocities are required. Such velocities might in fact be deleterious in terms of suspending solids in the types of unlined channel systems which are likely to be affordable.

The pumping needs of a channel growth system could be met with any suitable combination of gas/liquid flow ratios $f(= Q_g/Q_L)$, and gaslift draft tube height and area. However, configurations must also be able to meet CO_2 transfer needs. Because of the importance of gas (CO_2) transfer and design constraints posed by gas transfer, this topic is addressed next.

A11.9 GAS TRANSFER FROM BUBBLE SWARMS TO LIQUID

Oxygen transfer is an important unit operation in fermentation and waste treatment. It has been established that the overall transfer coefficient, K_L, is in one range for small rigid sphere bubbles D_BL <2 mm, and in another range - about twice as high, for larger bubbles D_B >2 mm. A number of correlations (for example as discussed in 21) have evolved for these bubbles of various size ranges. This would seem on the surface to imply a difficult analytical situation. However, even with this disparity of transfer coefficient with bubble size it fortunately transpires that the coalescence and redispersion of bubbles in a rising swarm leads to a steady state size distribution a short distance above the sparger as discussed earlier (and as intuition would suggest). In this steady state the ratio of K_{La}, the combined mass transfer coefficient (whose units can be h_r^{-1} or mmol/L.hr.atm) to volume of the rising gas, is constant. Also, gas pressure (hence solubility) and area terms cancel. A very important implication of this is that the fractional approach to liquid equilibrium activity of a given gas component in the gas phase is constant per unit time. (It has been established that power dissipation in the liquid does not affect K_L.) If thermodynamic activity of gas dissolved in the liquid phase stays constant relative to gas phase acativity, then the fraction of gas phase component which is lost from the gas is constant per unit time. Thus,

$$dP_g/dt = P_g k_g / \{(P_g - P_L)/P_g\} \tag{13}$$

where

P_g = activity of species in gas phase, atm
P_L = activity of species in liquid phase, atm
K_g = transfer coefficient based on gas activity

more specifically the transfer coefficient can be expressed as

$$K_g = \frac{\text{Atmospheres gas lost/unit time}}{\text{Atmosphere gas driving force}}$$

The constant fractional stripping of of oxygen per unit height has been established in particular for reference 8 (Hatch) and in reference 13 in the deep tank work of Schmidt and Redmond.

For oxygen, the fractional absorption per unit time from the gas phase is .0065 - .0075 sec^{-1} at 20°C (8, 13). A value of 0.007 will be used in this analysis. However, the critical problem for algal growth systems is not oxygen transfer, but CO_2 transfer. Though no literature was found directlly applicable to CO_2 transfer in the situation envisioned for the algal growth system, a well established mass transfer correlation may be used to compare the Kg for CO_2 transfer from that of O_2 [7, 13]. Specifically, at a given gas phase partial pressure the liquid side mass transfer coefficient per unit area, which controls K_L, is proportional to $D_L^{0.67}$ (where D = gas diffusivity in liquid). Kg is in turn proportional to K times solubility. Thus

$$\frac{Kg_{CO_2}}{Kg_{O_2}} = \left(\frac{D_{L CO_2}}{D_{L O_2}}\right)^{2/3} \left(\frac{S_{CO_2}}{S_{O_2}}\right) = \left(\frac{1.77\times10^{-5}}{2.28\times10^{-5}}\right)^{.667} \left(\frac{29.1}{1.38}\right) = 17.8 \qquad (14)$$

and Kg for CO_2 can be computed to be about 0.125 sec^{-1} (though a cautionary note is that smaller bubble size with surfactant present could give a much higher Kg). With this Kg, a swarm of gas bubbles rising through gas-free liquid would lose about 13% of their remaining CO_2 each second. (In deriving CO_2 transfer data from O_2 data, note here that it has been established that there is negligible chemical acceleration of CO_2 uptake - see Ref. 17).

A11.10 COMPUTATION OF TIME AND HEIGHT NEEDED FOR TRANSFER

The limiting case will sometimes exist where back pressure of CO_2 is negligible. This will be true in particular when the system is operated at low CO_2 and high alkalinity to prevent CO_2 escape to the atmosphere [17]. In this case computation of the required contact time is straightforward from equation (13) and Kg = 0.125 sec^{-1}. Required contact times for 80% and 95% absorption of CO_2, for both 15 cm/sec and 30 cm/sec draft tube velocities in each case, are shown in Table AII-2. These values of draft tube height and flow velocity are those which were used earlier to compute representive efficiencies.

In the majority of cases CO_2 transfer will take place where backpressure is not zero. The partial pressure of CO_2 (P_L in equation 13) is needed and can be derived as a function of any of several interrelated parameters. One convenient derivation is as a function of pH (or H^+ concentration) and alkalinity, A. Here, alkalinity is defined as milliequivalents per liter excess of cations over all anions other than carbonate species. It is also assumed as a simplifying measure in this analysis that no buffers other than carbonate are present. The simplification can also be made for pH values of interest (5-11) that H^+ and OH^- are negligible relative to carbonate species. Data in the analysis to follow is from reference 18 and 19. The term C CO_2 is used to define the sum of CO_2 and H_2CO_3 in solution.

A = meq/L alkalinity = $HCO_3^- + 2CO_3$ =

given that $$\frac{(H^+)(CO_3^=}{HCO_3^-} = 4.4 \times 10^{-11} \text{ Refs. 18, 19} \tag{15}$$

then $$CO_3^= = \frac{4.4\times10^{-11}\ HCO_3^-}{H^+} \tag{16}$$

and

$$A = HCO_3^- \frac{H^+ + 8.8\times10^{-11}}{H^+} \tag{17}$$

so $$HCO_3^- = \frac{AH^+}{H^+ + 8.8\times10^{-11}} \tag{18}$$

Also, given that

$$(H^+)(HCO_3^-)/C_{CO2} = 4.31 \times 10\text{-}7$$

and $$C_{CO2} = \frac{H^+}{4.3\times10^{-7}}\ \frac{AH^+}{H^+ + 8.8\times10^{-11}}$$

$$C_{CO2} = \frac{A(H^+)^2}{(4.3\times10^{-7})(H^+ + 8.8\times10^{-4})} \tag{19}$$

letting $K_1 = 4.31 \times 10^{-7}$ and $K_2 = 4.4 \times 10^{-11}$,

total carbon in solution, TC, is given by

$$TC = \frac{A(H^+)^2}{K_1(H^+ + 2K_2)} + \frac{AH^+}{H^+ + 2K_2} + \frac{K_2 A}{H^+ + 2K_2} \qquad (20)$$

As solubility of CO_2 is 29 mmol/L .atm (20°C) the term P_L for use in equation (13) would be 1/29.1 atm X C CO_2 or 0.034 C_{CO2} at 20o.
Other temperatures can be treated in the same way using the CO_2 solubility for those temperatures.

A11.11 COMPUTATION OF REQUIRED RESIDENCE TIME IN DRAFT TUBE

The entire proacedure for computing transfer time (or height) is complex and only the briefest possible description will be given here. Given equation 19 and 20 a graph can be constructed showing total carbonate species (i.e. mmol/L C in liquid) and PCO_2 as a function of pH. It is assumed here that gas and liquid flows are known, and that the starting compositions of both the gas and liquid phases (at the sparger in the draft tube) are known. Thus a point can be drawn above the inlet liquid composition/pH point showing pCO_2 of the inlet gas and liquid. Gas and liquid flows will be concurrent where the gaslift is used for pumping as well as CO_2 transfer; thus the sum of gas and liquid fluxes upward is constant. Specifically,

$$Q_g(V_L + 30)P_{CO2}/RT = Q_L V_L \ TC \qquad (21)$$

where Q_g and Q_L are flow rates of gas (S.T.P.) and liquid. The material balance of equation 21 can be used to construct a PCO_2 for the gas phase corresponding to each PCO_2 of the liquid phase. The time needed for transfer can then be computed through the following iteration:

1) Starting with initial PCO_2 of gas and liquid elements, choose a PCO_2 increment for each (negative for gas, positive for liquid) by equation 21 small enough that use of averages for P_L and P_g gives acceptable error. Then the average driving force for transfer is $(PCO_2gI + PCO_2gF) - (PCO_2LI + PCO_2LF) = PCO_2$ average. The subscripts g, L, I, and F are gas, liquid, initial, final respectively.

2) Compute time necessary for transfer as

$$\Delta T = \frac{(P_{CO_2,g,I} - P_{CO_2,g,F})}{RT\,k_g \cdot \Delta P_{CO_2,\,average}}$$

3) Repeat procedure for the next chosen increment of gas and liquid PCO_2.

The procedure is continued until the final desired P_{CO2} values for gas and liquid are attained. The height needed for transfer is easilv derived from this time and the gas velocity (V_L + 30).

There certainlv exists a rigorous analytical solution for the height calculation, but it is almost certainly too messy to be useful. The preceding computation should, however, be easy to accomplish with an appropriate computer program.

A11.12 COMBINED PUMPING AND GAS TRANSFER

To this point in the analysis expressions have been developed for gas flow required for the pumping function, and for the draft tube height needed for CO_2 transfer. Regarding which independent variables are to be fixed, two sets of circumstances are of interest: one situation is where the draft tube height is sufficient only to accomplish needed CO_2 transfer, with the balance of pumping - if needed - accomplished by other means. The other situation is where transfer requirements are met and gas does all of the pumping work as well.

In a typical channel flow system, depth and flow velocity will be fixed on the basis of various criteria such as 1) sufficiency to mix algae, 2) avoidance of erosion and sediment transport in unlined pond sysstems, 3) spacing of gas transfer and mixing stations, and other factors (more discussion is presented in reference 17). Choice of depth and channel flow velocity dictate pumping energy requirements and grade, or liquid surface slope in the direction of flow. Grade in turn determines pumping head (= grade X length of flow path). Carbon dioxide addition must be sufficient to meet the needs of photosynthesis by the algae. The assumption here is that algal productivity will average 75 metric tons/hectare yr., with a maximum productivity of 20 gm/m^2.dav and that 2 gms of CO_2 must be transferred to the system per gram of algae which is grown [17]. It is assumed thaat CO_2 loss from the liquid surface - or in alternate terms, back diffusion to the atmosphere - is minimal so that the CO_2-containing gas stream can be introduced into the system, and mixing can be carried out, 24 hr/day. This is an operating characteristic important for system economics. For purposes of illustration and analysis, model systems will be assured with characaterístics shown in Table AII-2.

TABLE AII-2

SYSTEM CHARAACTERISTICS USED FOR ILLUSTRATIONS (See Text)

A. Stack gas use

I. Channel flow velocity = 15 cm/sec
Channel Depth = 30 cm
Pumping head = 7.2 cm (From Manning equation, n = .02, pumping station = 3240 m, 6 hr circulation time)
Algal Productivity = 20 gm/m^2.day
CO_2 requirement = 40 gm/m^2.day
80% absorption of CO_2 from gas stream
Required Q_g/Q_L = 0.155 (Q_g expressed at 25°C, 1 atm, 15% CO_2)
Required h_L for 80% absorption = 5.4 M

II. As above, except

Channel Flow Velocity = 30 cm/sec
Pumping station spacing = 6480 meters
Pumping head = 58 cm (6 hr circulation time)
Q_g/Q_L = 0.139 (Q_g at 25°C 1 atm, gas has 15% CO_2 by volume)
Required h_L for 90% absorption = 13.4 meters

B. Pure CO_2 use

With pure CO_2 (other assumptions above hold):

For I, Q_g/Q_L = 0.023

II, Q_g/Q_L = 0.0205

See text for further discussion of pure CO_2 case.

A11.13 COMPARISON OF GASLIFT PUMPING CAPACITY WITH CHANNEL GROWTH SYSTEM NEEDS

The choices for CO_2 content of the sparged gas, in combination with desired fractional CO_2 stripping and desired fluid flow velocity in the gaslift draft tube fix the gaslift draft tube height. (Calculated heights for the examples are shown in Table AII-2.) Values chosen for these parameters also fix the gas holdup in the system, and power input which is directly proportional to gas holdup. Holdup needed to provide a given pumping head, from equation 6, can be compared to the actual holdup from chosen parameters by equation 4. If holdup by equation 4) is less than needed from equation 6) then auxilliary pumping power must be provided.

Holdups calculated by equation 4 for case AI and AII are 0.0406 and 0.045 respectively. If the pumping head needs for the two systems are 7.2 and 58 cm, respectively, then pumping power availability is a factor of above three times need for case AI, but is in almost perfect balance (holdup computed at 1.04 times need) for the assumptions of case AII. As there are uncertainties (discussed under "uncertainties" later) this must be considered a perfect match within computational precision.

For the pure CO_2 transfer case, it is necessary to take into account the volumetric change as CO_2 is absorbed during bubble rise. The situation is complicated by back diffusion of oxygen into the bubbles. The assumption made here is that the CO_2 volume of the rising gas decreases exponentially with height (as would occur with the constant bubble size distribution discussed earlier) but that back diffusion of oxygen is equal to 25% of the CO_2 which diffuses out. For the 80% CO_2 absorption case, this leads to a final gas volume of 40% of the initial, and for the 95% absorption case, a volume which is 29% of the initial. The average gas flow (STP) used in computing holdup is considered to be the log mean of inlet and outlet, times a pressure correction factor, P_1/P_2, for a depth of 2/3 of h_L (3.6 M for case BI, 9M for BII) to account for the fact that the majority of gas is at lower than mean depth. (This correction factor is to be distinguished from the term C_{PE} used earlier). With these assumption, the holdups for cases BI and BII are 0.0035 and 0.00304. These values are only 15% and 5% of need, respectivelly, for 15 and 30 cm/sec channel flow velocities, so that an auxilliary pumping process would be needed where pure CO_2 is used to supply carbon needs.

The above calculations are only illustrations, but it is evident that when CO_2 is supplied by stock gas with plausible requirements for mixing velocity and fractional stripping, that gaslift pumping capacity can be capable of meeting total system needs. It is worth noting that a value of 0.02 was used for the Manning equation roughness, n (See 22 for discussion). A value of n = 0.01 could equally well be encountered with a smooth channel bottom, which would mean for even channel flow velocities of 30 cm/sec that pumping power of the stack gas transfer system could be well in excess of needs. At n = 0.01, power availability for the pure CO_2 case could supply a large function, or possibly all pumping needs, if channel velocities of 5-10 cm/sec could be tolerated. Uncertainties which attend the estimates are discussed later, but a good match between gaslift pumping power and needs seems possible with reasonable assumptions.

A further important consideration is that the system must also be energy efficient, which is to say that the CO_2 transfer/pumping functions should not consume a large function of system energy output. Net energetics are examined next.

A11.14 NET ENERGETICS; GASLIFT ENERGY CONSUMPTION AS A FRACTION OF TOTAL GROWTH SYSTEM OUTPUT

In addition to choices made earlier, two additional parameters must be assumed in order to calculate energy demand of the gaslift system as a function of system output. These are 1) the gross system energy output per unit area per day, and 2) the thermal to mechanical conversion efficiency which is obtained when converting the gross energy output, in the form of dilute algae, to mechanical energy. The working assumption here for 1) is that the 20 gm/m^2 day of algae have a gross heating value of 5000 cal/gm (= 9000 BTU/lb) so that system gross energy output is 100 Kcal/m^2.day. A second assumption, without specifying mechanism, is that this gross energy can be converted to mechanical energy at a thermal - to mechanical conversion ratio, or efficiency, of 30%. Available mechanical energy, from which energy for gaslift operation must come, is with this assumption 1.26 X 10^2 erg/m^2.day. Using equation (8) and a value of E_{ML} = 0.8, the fraction of total system energy requirement for sample case AI of Table 2 is 7.9% and for AII, pumping through 13.4 M, it is 15.2%. These are fairly major, but not intolerable, levels of energy consumption. For cases BI and BII, where CO_2 is pumped tghrough the same heads, with other assumption as presented earlier, consumptions are 1.2% and 2.2% of total system output. Carbon dioxide from some sources (e.g. geological reservoirs) may be pressurized. No energy at all would be required to sparge this CO_2 if it is available on site at more than about 2 atmospheres.

A11.15 UNCERTAINTIES

A number of uncertainties exist in the preceding calculations regarding efficiency of gas transfer and net energy consumption by the gaslift. Many of these have already been mentioned, but it is worthwhile to review the possible sources of error and uncertainty here. Some sources of error and their magnitude are discussed below.

A11.15.1 Value for Transfer Rate for Carbon Dioxide in Rising Bubble Swarm

A value of 0.125 sec^{-1} for Kg was derived based on the premise that gas bubbles in algal systems will behave as air bubbles do in steady-state dilute aqueous systems. There is major uncertainty in the assumption, but a range cannot be given to the uncertainty at this time. Experimental values for Kg are being observed at 2-5 times the 0.125 sec^{-1} assumed here [22]. This may indicate high fractional CO_2 stripping from small bubbles near the sparger before coalescence/dispersion establishes a steady state, or else, prevention of coalescence (hence maintenance of smaller CO_2 bubbles and higher Kg) by surfactants in algal growth systems. Experimentation will be required to give some answers in this area. This topic is quite important as both energy consumption (but also available pumping capacity) are inversely proportional to Kg.

AII.15.2 Variations in Channel Roughness

Values for n, the channel wall and bottom roughness, could easily vary by a factor of 3, which could change power requirements by an equal factor. Establishment of the value for n (if not 0.02 as assumed) will require considerably more analysis and is beyond the scope of this treatment.

AII.15.3 Temperature/Water Vapor Corrections

In computing energy consumption by the compressor, and buoyancy effect of sparged gas, account must be taken of temperature differences between the compressor inlet and in the rising column of bubbles in the draft tube. Corrections should also be made for variations in water vapor content between the two points. However, these corrections are small with minimum total error at less than 5%.

AII.15.4 Non-uniform Velocity of Liquid in Draft Tube

The calculation of pumping work done on the liquid by rising gas bubbles was predicated on liquid "plug flow," i.e. that all liquid elements in the draft tube had the same velocity. Obviously, they do not, and there will be an efficiency loss relative to that calculated earlier if fluid has a velocity distribution characteristic of turbulent flow [21] even when gas bubbles are uniformly distributed over this area. This error in pumping efficiency calculating has been worked out [23; detail not shown] and is, surprisingly, less than 2%. This error is small because the turbulent flow velocity profile is rather flat, though it does vary. If gas "channels," or is unequally distributed in the cross section, or travels in fluid elements having much greater than average velocity, then pumping efficiency losses could be large. It is not possible to state what the efficiency decrement might be, but in any event this situation can be avoided by sparging so as to introduce gas uniformly over the draft tube cross section.

AII.15.5 Changes in Gas Stream Volume as Carbon Dioxide is Absorbed

Power input and efficiency were computed earlier for stack gas (15% CO_2) sparging on the basis of unchanging gas volume. However, the actual stream flow could actually change by as much as 15%. This 15% seems to be the likely maximum error from this source. Back diffusion of O_2, and the fact that some CO_2 remains with the gas stream during its rise, will probably keep this error to 10% or less, with stack gas. For pure CO_2 sparging, the gas absorption correction is obviously major, but CO_2 pumping is not a major energetic drain (if any energetic drain at all) and pure CO_2 sparging does not appear to supply enough pumping power in most likely cases, to meet system needs. Thus modeling gaslift pumping by pure CO_2 streams does not seem to be a pressing need.

There are likely to be other sources of error which have not been considered. However, it seems that the surface roughness and CO_2 absorption Kg uncertainties are dominant. Thus in summary, the conclusion must remain that a gaslift system capable of transferring stock gas should be capable of providing enough or more than enough pumping capacity to meet the needs of a high rate pond system with channel flow velocities of 15-30 cm/sec. Energy consumption for this approach should be a high (5-25%) but tolerable function of system output.

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AIII-1 TRAVELLING SOLIDS REMOVER FOR POND SUMPS

(By Joseph C. Dodd, Consulting Engineer)

Description (See Figure AII-1)

Continuous rails are provided so that a single solids remover may service a number of pond sumps where the ponds are arranged to allow this. The rails are supported on piers and pond walls, with spacing to allow use of light rolled berms such as 6" or 8" WF or B sections (span 15-20 ft.). Steel sections at about 15#/ft with a high quality protecive coating is preferred, although fiberglass sections are available but more expensive. The pier spacing must be coordinated with the walkway support requirements, which may be more critical if timber is used (2" x 10" stringers with 15' span previously used). Minimum pier thickness is 6" for rebar cover.

The travelling solids remover is fabricated from 6" Cs, rolling on 6" steel wheels with rubber treads (to protect coatings). Two wheels are driven by a hand crank to move the collector at the speed which just picks up the blanket of solids. A sight glass on the suction hose may be used to estimate the rate of travel. The collector is a stainless steel pipe arm with perforated tee pipe having rollers to maintain the desired bottom clearance. The arm is pivoted up to clear the walls when transferring to another pond. It may also be moved along the solids remover rail to sweep several strips along the bottom to remove as much settled solids as possible. The solids are pumped by a self-priming pump (e.g. progressing cavity).

Power supply to the pump and solids discharge hose are draped between a fixed mast at the center of the sump and a mast on the solids remover. Quick connect couplings allow transfer from one pond to the next. The discharge of solids is conveyed in a pipe supported along the walkway or rail, either to disposal, or to a settling tank(s) for thickening and return of supernatant to the pond. The latter method reduces salt loss and disposal requirements but adds to facilities cost and operational complexity.

Costs for 192 Hectare (475 acre) System

For saline systems, frequency of cleaning is less than with wastewater systems, so two removers can probably handle 12 ponds, or a total of 4 removers for the 192 hectare system. A 2" or 3" pump should be used, operated at a reduced speed (say 500 - 800 rpm) due to abrasive nature of solids. Settling tanks (if used) should be designed for 50 gpm overflow rate of 3000 gpd/ft^2. Need 24 ft^2 or approx. 6 ft. diameter, with hopper bottom.

Rough estimating costs:

Rails @ \$12/ft. including coatings, specials, anchor bolts, installation,

Pier concrete (assume 10 per pond)

0.5' X 1.5' x 4.5' = 33.8/27 = 1.25 yd^3/pond

1.25 x 24 ponds @ \$200/yd^3 =	\$6.000

Solids remover (ea).

Complete except for pump, elec. & piping	10,000
Pump, elec. & piping (to fixed mast)	5,000

Total	\$15,000

Assume piping & electrical on pond structure is included in other estimates (% of project costs).

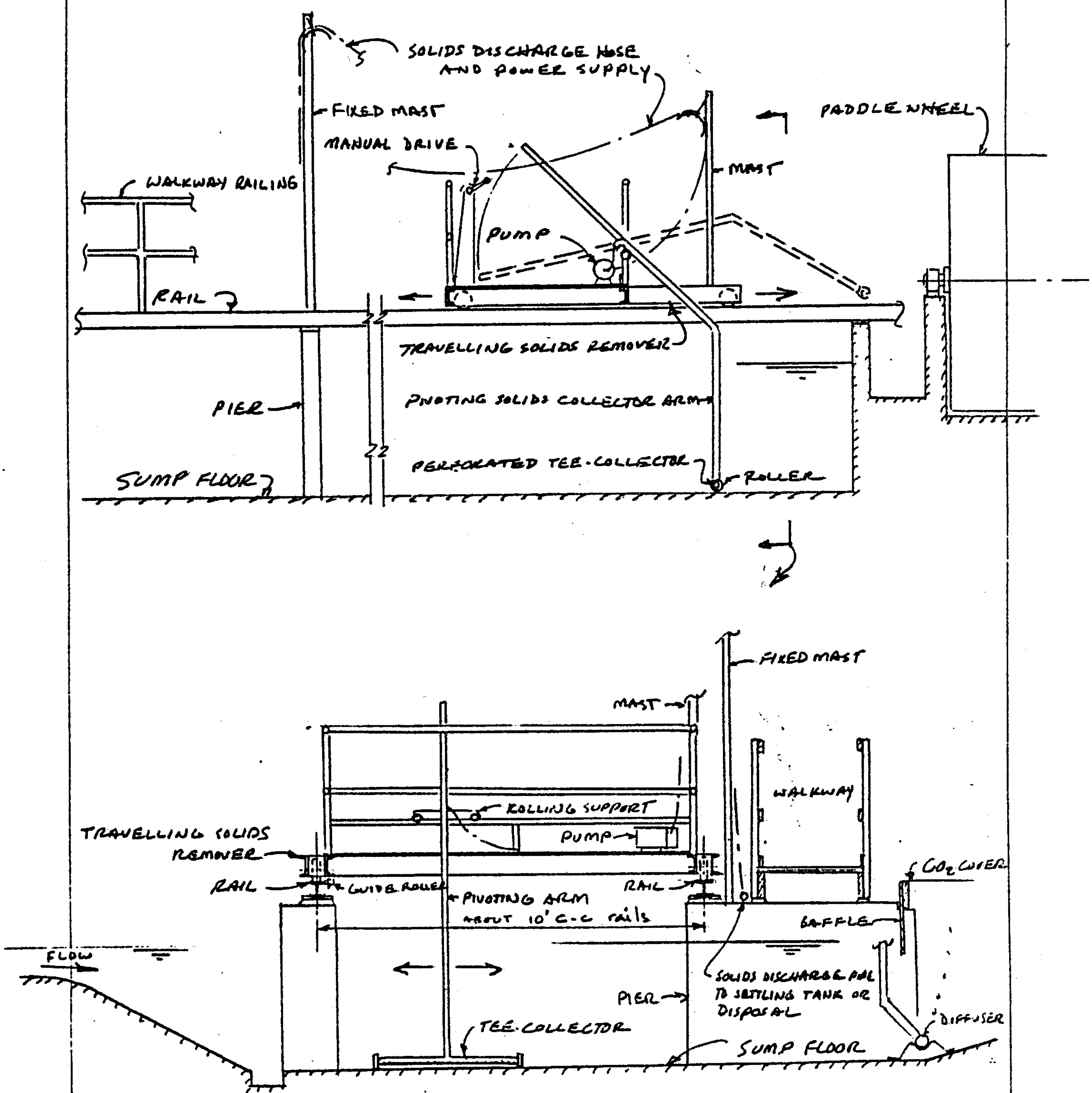
Total for 475 acre module:

Rails (4 rails x 3400' = 13,600 @ \$12/ft.)		163,000
Piers		6,000
Solids removers 4 x 15.000		60,000

Total without tanks		229,000
With tanks, add 4 @ \$5,000		20,000
	TOTAL	\$249,000
	TOTAL/HECTARE	\$1,297

Figure A-III-1

Travelling Solids Remover for Pond Sumps
(Joseph Dodd, consulting engineer)



¼" ~ 1'-0 approx

Appendix AIII-2

SLUDGE REMOVAL FROM SHALLOW GROWTH PONDS

(by Lloyd Bracewell, Consulting Engineer)

Shallow high rate algal growth ponds with mixing low velocities will eventually experience some settling out of organic and inorganic materials to form a sludge layer over the pond bottom. Since the growth ponds have only about 20 cm of depth, any significant layer of sludge would interfere in the pond's operation. It is therefore necessary to provide for the removal of sludges from the bottom of the growth ponds once the depth reaches about 15 percent of the water depth or 3 cm. The proposed method of construction of the growth ponds and their layout create severe constrints on the methods that can be used cost-effectively for sludge removal. In particular, access is limited to the ends and the pond bottoms will be of earthen construction.

A form of hydraulic removal of sludge would therefore be most appropriate if required on a frequent basis. If the pond mixing velocity is increased several-fold then the sludge would be resuspended in the water column and the water could be withdrawn for settling in another basin with adequate detention time for solids removal. This approach would require sizing the mixer drive motors much larger than necessary for normal mixing purposes. One possible solution to keep the installed size of the motors to a minimum would be to design the mixing systm to include a power take-off so as to accommodate the installation of a supplemental driver only when sludge removal was needed.

The sludge removal approach proposed whenever thorough pond cleaning is required, such as annually, is to decant the pond water, install a boom and sludge pump arrangement, operated by a winch and by a combination of dragging and pumping, move the sludge to small sumps at each end of each pond. The boom would be the width of the pond and dragged by several cables connected to winches installed on mounts for that purpose when needed to remove sludge. Small sludge pumps would be mounted on the boom with 100 ft discharge pipes suspended along the pulling cables to keep excessive amounts of sludge from accumulating in front of the boom as it is dragged along the pond bottom toward a sludge sump. This reduces the load on the cables and boom as the sludge is gradually moved to the end of the pond. The sludge collected in the end slumps would be pumped to a storage pond for later disposal. The boom would be constructed of 12 by 12 inch wooden beams 20 feet long and hinged at each connecting joint for the width of the pond. A pulling cable would be attached at each of the connecting joints and attached to individual winches. A 3 horsepower sludge pump would be mounted on each section of the boom with its suction manifold set along the front of that section of the boom to pump sludge as it acculates in front of the boom. The pump discharge hose would be extended along the pulling cable approximately 100 feet to keep the pumped sludge ahead of the boom as it travels the length of the pond. The sludge as it was pumped would tend to flow, albeit slowly, toward the end of the pond since the dragging would be done in the direction of the pond gradient.

The estimated cost for a sludge removal system of this type is about $60,000. A single system would be shared among all the ponds.

Appendix AIII-3

DIVIDER WALL COST ESTIMATES*

METHOD	----- MATERIALS ------			-------- LABOR --------			
	QTY PER FT	UNIT COST	TOTAL	HOURS	HOURLY RATE	TOTAL	SUBTOTAL
1. HYPALON MEMBRANE							
POSTS	0.250	$1.65	$0.41	0.025	$35.00	$0.88	$1.29
CONCRETE	0.004	$51.50	$0.19	0.004	$16.35	$0.06	$0.25
MEMBRANE	2.000	$0.50	$1.00	0.100	$16.35	$1.64	$2.64
RAILINGS	1.000	$1.00	$1.00	0.020	$16.35	$0.33	$1.33
TRENCHING	0.019	$1.00	$0.02				$0.02
						TOTAL	$5.51
2. CONCRETE BLOCK							
BLOCKS	1.780	$0.66	$1.17	0.130	$16.35	$2.12	$3.30
MORTAR	0.080	$2.60	$0.21				$0.21
GROUT	0.244	$2.25	$0.55	0.067	$16.35	$1.09	$1.64
STEEL	1.000	$0.23	$0.23	0.008	$16.35	$0.13	$0.36
FOOTING EXC	0.019	$1.00	$0.02				$0.02
CONCRETE	0.019	$50.00	$0.93	0.041	$16.35	$0.67	$1.60
						TOTAL	$7.12
3. POURED CONCRETE							
CONCRETE	0.035	$51.50	$1.80	0.061	$14.40	$0.88	$2.68
STEEL	1.000	$0.23	$0.23	0.008	$16.35	$0.13	$0.36
FORMWORK	2.670	$0.10	$0.27	0.116	$17.35	$2.01	$2.28
FOOTING EXC	0.019	$1.00	$0.02				$0.02
						TOTAL	$5.34
4. PIPING**							
PIPE	1.000	$6.77	$4.41	0.012	$94.00	$1.13	$5.54
FOOTING	0.025	$50.00	$1.23	0.055	$16.35	$0.90	$2.13
						TOTAL	$7.67

* Supplied by Lloyd Bracewell, consulting engineer
** Not described in text

Appendix AIII-5

Large Scale System Paddle Wheel Costs

* PADDLE WHEEL COST BREAKDOWN *

8.0 Hectare Basis

DESCRIPTION	MATERIALS QUAN	UNITS	UNIT $	COST	FAB. COST	TOTAL COST	COMMENT
Drive Tube	4656	lb	0.55	2,561	1,280	3,841	16" 150 psi steel
Hubs	498	lb	0.55	274	411	685	
Flanges	869	lb	1.00	869	0	869	
Spokes (f-glass)	431	ft	2.75	1,185	592	1,777	Fiberglas 2x2x0.25
Paddles (")	1473	sq ft	6.00	8,836	incl.	8,836	Fiberglas 0.30" thk
Fasteners				353		353	
Drive motor	1		900	900		900	15 hp, Severe Duty
Motor Starter	1		450	450		450	NEMA size 2, type 4
Motor Base	1					277	Base only, not struc.
Var. Speed Unit	1		2,700	2,700		2,700	Beier 15ACY, 29:1
Speed Reducer	1		1,400	1,400		1,400	Cyclo 1870
SR Base						275	Base only, not structral support
SR Sprocket	1		25	25		25	
Jackshaft	54	lb	6	324	324	648	3.38" dia S.S., 24"long, w/ keyways
JS Bearings	2		100	200		200	
JS Spr. #1	1		75	75		75	
JS Spr. #2	1		50	50		50	
PW Solid Shafts	112	lb	6	672	672	1,344	3.5" dia S.S., 12" long, one keyway
PW Drive Spr.	1		100	100		100	
Dr. Chain #1	1		25	25		25	
Dr. Chain #2	1		50	50		50	
Chain Guards	2		100	200		200	
PW Bearings	6		250	1,500		1,500	Roller bearing, split housing
PW Coupling	2		850	1,700		1,700	Falk 120T(1st), 100T(2nd)
Torque Limiter	1		350	350		350	
Misc.			800				
TOTAL				24,799		28,632	
INSTALLATION (not incl structural)						7,158	
TOTAL INSTALLED COST						$35,790	

Appendix AIII-6

Experimental System Paddle Wheel Costs

* PADDLE WHEEL COST BREAKDOWN *

0.4 Hectare Basis

DESCRIPTION	MATERIALS QUAN	UNITS	UNIT $	COST	FAB. COST	TOTAL COST	COMMENT
Drive Tube	819	lb	0.60	492	246	737	8" sch 40 steel
Hubs	139	lb	0.60	84	125	209	
Flanges	195	lb	1.00	195	0	195	
Spokes (f-glass)	140	ft	3.00	420	210	630	Fiberglas 2x2x0.25
Paddles (")	331	sq ft	6.00	1,987	incl.	1,987	Fiberglas 0.25" thk
Fasteners				79		79	
Drive motor	1		350	350		350	2 hp, Severe Duty
Motor Starter	1		200	200		200	NEMA size 0, type 4
Motor Base	1					47	Base only, not struc.
Var. Speed Unit	1		750	750		750	Beier 2ACY, 29:1
Speed Reducer	1		700	700		700	Cyclo 1870
SR Base						275	Base only, not structral support
SR Sprocket	1		25	25		25	
Jackshaft	14	lb	7	98	98	196	1.63" dia S.S., 24"long, w/ keyways
JS Bearings	2		50	100		100	
JS Spr. #1	1		50	50		50	
JS Spr. #2	1		25	25		25	
PW Solid Shafts	10	lb	7	70	70	140	1.75" dia S.S., 12" long, one keyway
PW Drive Spr.	1		75	75		75	
Dr. Chain #1	1		25	25		25	
Dr. Chain #2	1		50	50		50	
Chain Guards	2		100	200		200	
PW Bearings	2		150	300		300	Roller bearing, split housing
PW Coupling	0		850	0		0	
Torque Limiter	1		300	300		300	
Misc.	1		250	250		250	
TOTAL				6,826		7,897	
INSTALLATION (not incl structural)						2,764	
TOTAL INSTALLED COST						$10,661	

(unit prices have been adjusted for quantity)

TABLE A-IV-1 SETTLING POND HARVEST OPTION
1st Stage Pumps and Piping Costs

For one harvesting station serving 8 growth ponds (64 hectares).
Includes 2 settling ponds, one pump station + piping.
Does not include secondary thickening (see A-IV-2).

PIPE: All pipe is "100 ft. head" PVC, includes installation	
Pond drain lines	
22" dia X 2300 ft @ $20/ft =	$46,000
Effluent return lines	
22" dia X 1100 ft @ $20/ft =	22,000
Settling pond supernatant drain system	
16" dia X 200 ft @ $12/ft =	2,400
VALVES: All valves are low head type with epoxy coated cast iron bodies, stainless steel structurals, installed	
Pond Drain (2 per pond)	
16 X 22" canal gate @ $1200 =	19,200
Effluent Return	
8 X 22" canal gate @ $1200 =	9,600
4 X 18" adapt-to-line gates @ $1500 =	6,000
Check valves at pumps 3 @ $600 =	1,800
Air release valves 10 @ $250 =	2,500
PUMPS: Primary supernatant, includes motor, starter and installation	
3 X 12" vertical mixed flow pumps @ $9,000 =	27,000
SUMP, with pump support	7,000
FITTINGS and MISC.	12,500

TOTAL COST/HARVESTING STATION	$156,000
TOTAL COST/HECTARE	$8,125

TABLE A-IV-2. SETTLING POND HARVESTING OPTION
Secondary Thickening Cost Breakdown

EXCAVATION: 1200 yd^3 @ \$3.00/yd^3	\$3,000
CONCRETE (or gunnite) TANK	
Structural 36 yd^3 \$300/yd^3	10,800
Slab bottom 38 yd^3 @ \$100/yd^3	3,800
CONCRETE ALLEY in SETTLING POND	2,000
PIPING & VALVES (installed)	
Settling pond concentrate lines	
22" dia. X 80 ft @ \$18/ft	1,500
Concentrate valve at sump (2)	2,400
Flush system piping	
18" dia X 1200' @ \$14/ft	16,800
Flush system valves	
32 X 15" mud valves with hydraulic actuators	12,000
Thickener influent/effluent piping	2,000
Supernatant decanter and return piping	2,000
Concentrate piping to central processing	
6" dia X 1000' (ave) @ \$4.00/ft	4,000
Flush diverter valves	3,000
PUMPS (installed on slab)	
2^o supernatant, 2 centrifugals, 500 gpm each	5,000
2^o concentrate, 2 solids handling, 300 gpm each	8,000
1^o concentrate, 2 screw pumps, 1500 gpm each	20,000
FLUSH SYSTEM CONTROLS	
Sequencer and solenoids, hydraulic lines	5,000
MISC.	8,700

TOTAL COST/HARVEST STATION	\$110,000
TOTAL/HECTARE	\$ 1,720

APPENDIX V

A COVERED LAGOON BIOGAS SYSTEM FOR ALGAE PROCESSING WASTES

by

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INTRODUCTION

The covered lagoon biogas system approach has been found well-suited for a variety of liquid wastes. For over 15 years lagoons have been covered for odor control or potable water protection. Biogas produced by these first facilities was often viewed as a process byproduct requiring flaring. However in April of 1982, the first covered lagoon system designed specifically for biogas collection and utilization was placed into operation at a hog farm in California (Chandler et al, 1983). For the past three years this system has demonstrated itself as being efficient, low cost and easy to maintain system. Recently a similar system has been applied to distillery wastes. Thus the covered lagoon approach appears amendable to a large variety of agricultural and industrial liquid wastes.

Covered lagoons are typically large unheated basins using floating rubber-like membrane covers. They have been found to be very efficient at converting wastes to biogas at temperatures as low as 50 degrees fahrenheit. Covered lagoons are suitable for ambient temperature waste streams in mild climate regions of the U.S.. Since algae production facilities must also be located in warm regions, it appears that converting algae processing wastes to biogas using the covered lagoon approach holds promise.

In this report a covered lagoon biogas system for processing wastes from a 1000 acre algae production facility is examined. This design uses the most current information and experience from other waste streams. This is a preliminary analysis and further work is necessary to verify the feasibility of the proposed design.

DESIGN PARAMETERS

The proposed 1000 acre facility is expected to produce 24,990 tons of processing waste volatile solids (VS) annually. The waste stream as-produced is 15 to 20 percent total solids (TS) of which 83 percent are VS. It has been assumed that 3,124 tons of VS are produced each month during a 6 month summer period (1.5 times annual monthly average) and 1,041 tons of VS are produced load during the summer months while only one unit will be operational during the winter. Each unit will be equipped with a automatic tracking device which will match gen-set output to biogas availability. These units are intended to be operated under lean burn conditions (air/fuel = 30/1) inorder that air emission standards can be met without catalytic converters. Each unit will have its own heat rejection system. Both gen-set and heat rejection system will be trailer mounted by the system designer and delivered to the site as a turnkey system. All three gen-sets will feed power to a single substation.

DESIGN COST

Table 5 presents the estimated installed cost of the covered lagoon system. The preparation and storage basin intended to have near-vertical gunite walls inorder to minimize the use of poured-in-place concrete. The covered lagoons were assumed not to need a bottom liner. Standard 45 mil 5 ply Hypalon is specified for the floating covers. The total installed cost for the covered lagoon system including 15 percent contingency and 20 percent engineering fee is $1,614,000.

Table 6 shows the turnkey system costs for the 3 MW gen-set system. These costs were quoted by a commercial system fabricator with previous experience packaging megawatt-size gen-set systems. The total turnkey cost is $2,200,000.

DESIGN ECONOMICS

Table 7 presents a economic summary for the 3 MW covered lagoon biogas system. The total capital cost has been estimated at $3,814,000. Gross electricity revenue has been estimated to exceed $1,000,000 annually. Net electricity revenue is expected to exceed $745,000 and result in less than a 5 year payback.

It should be again noted that the system described was sized to accomodate the maximum summer processing waste production level. This to some degree has adversely affected system economics because of the drastic seasonal variation in processing waste production. For example, the 3 MW system has a very low installed cost of $1,270/kW. If waste production could of been maintained at summer condition levels year round, the system would possibly pay for itself in less than 3 years.

This simple analysis accounts only for the electricity derived revenue produced by the system. Obviously the system accomplishes some waste treatment and disposal benefit. It was assumed that the market value of solids removed from the lagoons and dried only equal their handling costs. Furthermore no credit has been taken for the gen-set recoverable waste heat. Thus the described covered lagoon biogas system could be beneficial in several ways to the 1000 acre algae production facility.

each month during the 6 month winter period (0.5 times annual monthly average). This results in a three-fold seasonal variation in waste stream production. It was assumed for this design that conversion and treatment of all of the processing wastes year round is desirable thus lagoons and engine generators (gen-sets) were sized to accomodate summer production levels.

Table 1 presents important design parameters for sizing the covered lagoons. It has been assumed that the processing waste stream will require dilution to 7 percent TS primarily due to it's high salt content. A high loading rate of 70 lbs/1000 cu ft - day was assumed. This loading rate has been employed with distillery wastes with low pH (4 or below). This is considered satisfactory given the long resulting hydraulic retention time (HRT) of 50 days and the near nuetral pH of the waste stream. In total, over 23 million gallons of covered lagoon volume is required.

Table 2 presents the estimated methane and biogas yield from algae processing wastes. It has been assumed that algae processing will produce a readily degradable waste stream. A volatile solids destruction efficiency of 65 percent has been assumed. The high methane and biogas yields shown are assumed to be achievable year round. Any reduction in biogas production due to decreased winter lagoon temperatures should more than be compensated by very long winter HRT's of over 150 days.

Table 3 shows the estimated summer and winter average daily biogas and electricity production. Over 1.6 million standard cubic feet (SCF) of biogas will be produced during summer months while only 0.5 million SCF is expected during winter months. Summer and winter electricity production capacities are 2970 and 990 kW respectively.

Table 4 presents a annual energy production summary for the covered lagoon system. Almost 373 million SCF of biogas equivalent to 2.2 million therms of natural gas will be produced annually. A total of 15.9 million kWh is expected to be produce. Recoverable waste heat from the gen-sets can range from 63,500 to 103,100 mmBTU annually.

DESIGN DESCRIPTION

Figure 1 presents a block diagram of the 3 MW capacity covered lagoon biogas system. The system consists of a 110,000 gallon waste preparation and storage basin where the as-recieved processing waste is diluted to 7 percent TS. The diluted waste is then pumped to three 8 million gallon capacity lagoons operated in parallel. Each lagoon has approximately 50,000 sq ft of liquid surface area. Liquid lagoon effluent is to be returned to the algae production ponds as a nutrient supplement while solids will periodically be pumped from the bottom of each lagoon and undergo further liquid solids separation in sand drying beds. Assuming less than 12 percent of inflow TS will reach the bottom and using a standard loading rate of 22 lb/sq ft-yr (U.S.E.P.A., 1974), 8 acres of sand drying beds have been specified.

The biogas produced will be removed from the lagoons by suction blowers and used to directly fuel 3 - 1 MW turbocharged gen-sets. It is anticipated that all three units will be operated near full

TABLE 1

COVERED LAGOON SYSTEM DESIGN PARAMETERS BASED ON SUMMER PRODUCTION LEVELS

Daily TS Production:	251,000 lb
Daily VS Production:	208,250 lb
Influent TS Content:	7%
Daily Flow:	432,000 gal
Loading Rate:	70 lb VS/1000 cu ft-day
Total Lagoon Volume Required:	23.3 million gal
Hydraulic Retention Time:	52 days

TABLE 2

ALGAE PORCESSING WASTE METHANE AND BIOGAS YIELD

VS Destruction:	65%
COD/VS (1):	1.25
Methane Content (1):	60%
Methane Yield:	4.83 SCF/lb VS added
Biogas Yield:	8.04 SCF/lb VS added

1. Adapted from: Jewell and Schraa, 1981

TABLE 3

SUMMER AND WINTER AVERAGE DAILY BIOGAS AND ELECTRICITY PRODUCTION

	Summer	Winter
VS Production (lb):	208,250	69,400
Biogas Production (SCF):	1,674,350	558,100
Elec. Prod. (kWh/hr) (1):	71,250	23,750
Ave. Capacity (kW/hr):	2,970	990

1: 23.5 SCF/kWh average conversion efficiency assumed

TABLE 4

COVERED LAGOON BIOGAS SYSTEM ANNUAL ENERGY PRODUCTION SUMMARY

Biogas (1):	372,828,000 SCF
Natural Gas Equiv. (2):	2,238,000 Therms
Electricity (1):	15,865,000 kWh
Gen-Set Waste Heat	
w/o Exhaust Recovery (3):	63,500 mmBTU
w/ Exhaust Recovery (4):	103,100 mmBTU

1: 334 days/yr production
2: Total BTU Production
3: Assumes 4000 BTU/kWh
4: Assumes 6500 BTU/kWh

TABLE 5

ESTIMATED COVERED LAGOON SYSTEM INSTALLED COST

	$
I. Preparation & Storage Pit	
1. 110,000 gal Gunite Basin @ $0.35/gal:	38,500
2. Pump System:	17,500
3. Plumbing & Controls:	25,000
4. Misc.	12,000
Subtotal:	93,000
5. Contingency (15%):	14,000
Total A:	107,000
II. Covered Lagoons	
1. 3 - 50,000 sq ft covers @ $2.35/sq ft:	352,500
2. Excavation 85,875 yds @ $2.25/yd:	193,200
3. Compaction 21,300 yds @ $3.00/yd:	63,900
4. Lagoon Solids Plumbing:	32,400
5. Solids Drying 8 ac @ $21,750/ac:	174,000
6. Gas Take-offs and Seals:	21,000
7. Gas Handling & Scrubbing:	150,000
8. Misc.:	90,000
Subtotal:	1,077,000
9. Contingency (15%):	162,000
Total B:	1,239,000
Total A+B:	1,346,000
III. Engineering (20%):	268,000
Total Installed Cost:	$1,614,000

TABLE 6

3 MW GEN-SET TURNKEY SYSTEM COSTS

	$
3 - 1 MW Turbocharged Gen-sets trailer-mounted:	1,770,000
3 - Heat Rejection Systems:	225,000
1 - Substation:	175,000
1 - Utility Metering Package:	30,000
Total:	$2,200,000

TABLE 7

3 MW COVERED LAGOON BIOGAS SYSTEM ECONOMIC SUMMARY

Total Installed Cost:	$3,814,000
Gross Electricity Revenue (1):	$1,031,200
System O & M (2):	$285,000
Net Electricity Revenue:	$745,600
Simple Payback (3):	4.6 yrs

1: Assumes Combined Production & Capacity @ $0.065/kWh
2: O & M Estimated @ $0.018/kWh
3: Assumes 10% ITC taken and subtracted from installed cost

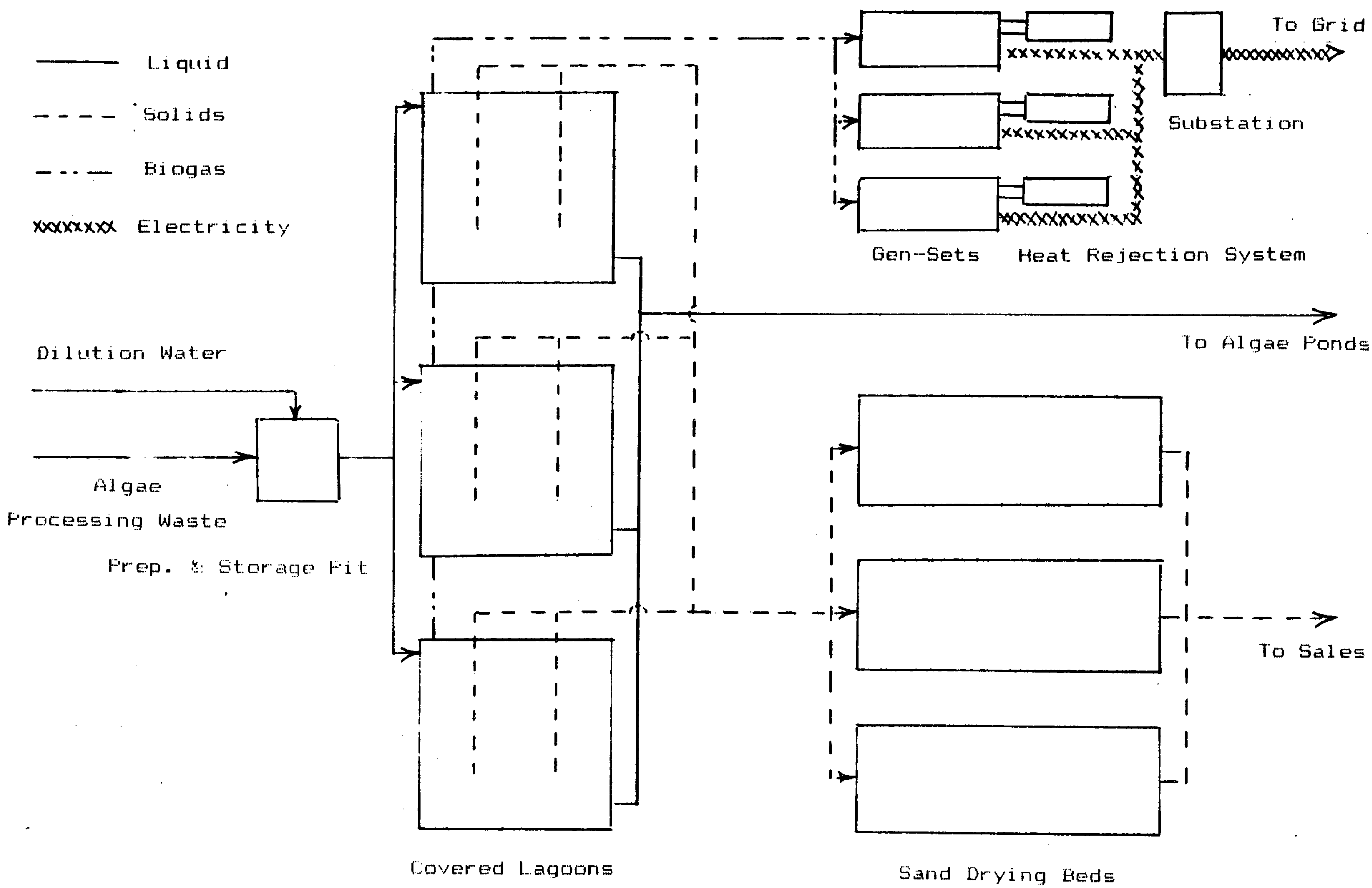

FIGURE 1: 3 MW COVERED LAGOON BIOGAS SYSTEM FOR PROCESSING WASTES FROM A 1000 ACRE ALGAE PRODUCTION FACILITY

REFERENCES

1. Anonymous, USEPA "Process Design Manual, Sludge Treatment & Disposal" EPA 625/1-74-006 October 1974

2. Chandler J. A., S. J. Hermes, K. D. Smith "A Low Cost 75kW Covered Lagoon Biogas System" Paper presented at ENERGY FROM BIOMASS AND WASTES VII, Lake Buena Vista, Florida January 25 1983

3. Jewell W. J., G. Schraa "Microalgae Separation, Concentration, And Conversion To Fuel With An Anaerobic Expanded Bed Reactor" USDOE Final Report #X8-9-82-63-1-3 July 1981

Table AVI-1a Capital Cost Summary - Base Case

Basis: 112 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS		
Earthworks*	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	740,667	3,858
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary**	3,027,088	15,766
Secondary (centrifugation)	1,602,080	8,344
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	105,400	549
Nutrient Supply System	316,200	1,647
Blowdown Disposal System*	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage*	210,800	1,098
Electrical Distribution (3% of above)	540,522	2,815
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above)*	1,896,475	9,877
CONTINGENCY (15% of above)*	3,129,184	16,298
LAND COSTS ($1250/hectare)*	1,011,840	5,270
TOTAL CAPITAL COST	$25,002,250	$130,220
DEPRECIABLE PORTION	$13,939,587	$72,602
NON DEPRECIABLE PORTION	$11,062,662	$57,618

Notes:

* non-depreciable item
** partially non-depr. ($375,000)
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-1b Operating Costs - Base Case

Basis: 112.3 mt algae/ha/yr (30 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
				1000 ac
	kg/kg	$/mt	$/ha/yr	$/yr
NUTRIENTS (1)				
CO_2	2.2	35	$8.65	$3,493
N, as NH_3	0.053	250	1.49	601
P, as Superphosphate	0.005	900	0.51	204
Fe, as $FeSO_4$	0.005	500	0.28	113
Total			$10.92	$4,412
FLOCCULANT	0.002	5000	$1.12	454
	10 kwh/yr	c/kwh		
POWER				
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	2.31	6.5	0.37	150
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
Total	11.55		$1.86	751
	kg/kg	c/kg		
SALT DISPOSAL	1.5	0.67	1.13	456
MAINTANENCE (matl's)			1.26	511
LABOR			1.39	562
TOTAL			$16.56	$6,691

(1) kg/kg = kg required / kg algae produced

Table AVI-2a Capital Cost Summary - Base Case + Nutrient Recycle

Basis: 112 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS	-------	---------
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	740,667	3,858
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	1,602,080	8,344
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	787,400	4,101
Non-depreciable Capital	382,600	1,993
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	105,400	549
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	575,622	2,998
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	2,016,985	10,505
CONTINGENCY (15% of above) (1)	3,328,025	17,333
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	2,200,000	11,458
	---------	-------
TOTAL CAPITAL COST	$28,726,701	$149,618
DEPRECIABLE PORTION	$17,618,719	$91,764
NON DEPRECIABLE PORTION	$11,427,334	$59,517

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-2b Operating Costs - Base Case + Recycle

Basis: 112.3 mt algae/ha/yr
(30 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
				1000 ac
	kg/kg	$/mt	$/ha/yr	$/yr
NUTRIENTS (1)	------	-----	--------	--------
CO_2	1.6	35	$6.29	$2,541
N, as NH_3	0.013	250	0.37	151
P, as Superphosphate	0.0025	900	0.25	102
Fe, as $FeSO_4$	0.005	500	0.28	113
			------	----------
Total			$7.20	$2,907
FLOCCULANT	0.002	5000	$1.12	454
	10 kwh/yr	c/kwh		
POWER	----------	-----		
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	2.31	6.5	0.37	150
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
	-----		-----	-----
Total	11.55		$1.86	751
Elect. Produced	-15.87	6.5	(2.55)	(1,031)
	-----		-----	-----
	-4.32		(0.69)	(281)
	kg/kg	c/kg		
SALT DISPOSAL	-----	----		
	1.5	0.67	1.13	456
MAINTANENCE (matl's)			1.97	797
LABOR			1.39	562
			-----	-----
			$10.99	$4,441
TOTAL			======	======

(1) kg/kg = kg required / kg algae produced

Table AVI-3a Capital Cost Summary - 0.5 x Prod. + Recycle

Basis: 56.2 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS	-------	---------
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	457,073	2,381
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	988,660	5,149
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	393,700	2,051
Non-depreciable Capital	191,300	996
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	65,043	339
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	529,950	2,760
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	1,860,181	9,688
CONTINGENCY (15% of above) (1)	3,069,299	15,986
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	1,100,000	5,729
	---------	-------
TOTAL CAPITAL COST	$25,643,129	$133,558
DEPRECIABLE PORTION	$14,566,771	$75,869
NON DEPRECIABLE PORTION	$10,820,503	$56,357

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-3b Operating Costs - .5xProd. + Recycle

Basis: 56.15 mt algae/ha/yr
(15 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
	kg/kg	$/mt	$/ha/yr	1000 ac $/yr
NUTRIENTS (1)				
CO2	1.6	35	$3.14	$1,270
N, as NH3	0.013	250	0.19	75
P, as Superphosphate	0.0025	900	0.13	51
Fe, as FeSO4	0.005	500	0.14	57
Total			$3.60	$1,454
FLOCCULANT	0.002	5000	$0.56	227
	10 kwh/yr	c/kwh		
POWER				
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	1.43	6.5	0.23	93
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
Total	10.67		$1.72	693
Elect. Produced	-7.93	6.5	(1.28)	(516)
	2.74		0.44	178
	kg/kg	c/kg		
SALT DISPOSAL	1.5	0.67	1.13	457
MAINTANENCE (matl's)			1.64	663
LABOR			1.39	562
TOTAL			$8.20	$3,312

(1) kg/kg = kg required / kg algae produced

Table AVI-4a Capital Cost Summary - 0.67 x Prod. + Recycle

Basis: 75 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS	-------	---------
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	560,646	2,920
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	1,212,690	6,316
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	527,558	2,748
Non-depreciable Capital	256,342	1,335
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	79,782	416
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	546,188	2,845
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	1,915,929	9,979
CONTINGENCY (15% of above) (1)	3,161,283	16,465
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	1,474,000	7,677
	---------	-------
TOTAL CAPITAL COST	$26,722,339	$139,179
DEPRECIABLE PORTION	$15,293,753	$79,655
NON DEPRECIABLE PORTION	$11,033,277	$57,465

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-4b Operating Costs - .67xProd. + Recycle

Basis: 75 mt algae/ha/yr
(20 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
				1000 ac
	kg/kg	$/mt	$/ha/yr	$/yr
NUTRIENTS (1)				
CO2	1.6	35	$4.20	$1,697
N, as NH3	0.013	250	0.25	101
P, as Superphosphate	0.0025	900	0.17	68
Fe, as FeSO4	0.005	500	0.19	76
Total			$4.81	$1,941
FLOCCULANT	0.002	5000	$0.75	303
	10 kwh/yr	c/kwh		
POWER				
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	1.75	6.5	0.28	114
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
Total	10.99		$1.77	714
Elect. Produced	-10.58	6.5	(1.70)	(687)
	0.41		0.07	27
	kg/kg	c/kg		
SALT DISPOSAL	1.5	0.67	1.13	457
MAINTANENCE (matl's)			1.71	690
LABOR			1.39	562
TOTAL			$9.10	$3,676

(1) kg/kg = kg required / kg algae produced

Table AVI-5a Capital Cost Summary - 1.5 x Prod. + Recycle

Basis: 168 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS		
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	985,598	5,133
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	2,131,873	11,104
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	1,181,100	6,152
Non-depreciable Capital	573,900	2,989
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	140,255	730
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	617,459	3,216
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	2,160,627	11,253
CONTINGENCY (15% of above) (1)	3,565,034	18,568
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	3,300,000	17,188
TOTAL CAPITAL COST	$31,643,768	$164,811
DEPRECIABLE PORTION	$19,644,484	$102,315
NON DEPRECIABLE PORTION	$11,999,284	$62,496

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-5b Operating Costs - 1.5xProd. + Recycle

Basis: 168.4 mt algae/ha/yr (45 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
	kg/kg	$/mt	$/ha/yr	1000 ac $/yr
NUTRIENTS (1)				
CO2	1.6	35	$9.43	$3,810
N, as NH3	0.013	250	0.56	226
P, as Superphosphate	0.0025	900	0.38	153
Fe, as FeSO4	0.005	500	0.42	170
Total			$10.79	$4,359
FLOCCULANT	0.002	5000	$1.68	680
	10 kwh/yr	c/kwh		
POWER				
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	3.07	6.5	0.49	200
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
Total	12.31		$1.98	800
Elect. Produced	-23.80	6.5	(3.83)	(1,547)
	-11.48		(1.85)	(746)
	kg/kg	c/kg		
SALT DISPOSAL	1.5	0.67	1.13	457
MAINTANENCE (matl's)			2.49	1,004
LABOR			1.39	562
TOTAL			$13.95	$5,635

(1) kg/kg = kg required / kg algae produced

Table AVI-6a Capital Cost Summary - 2.0 x Prod. + Recycle

Basis: 225 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS		
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	1,205,472	6,279
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	2,607,465	13,581
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	1,574,800	8,202
Non-depreciable Capital	765,200	3,985
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	171,544	893
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	656,812	3,421
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	2,295,737	11,957
CONTINGENCY (15% of above) (1)	3,787,967	19,729
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	4,400,000	22,917
TOTAL CAPITAL COST	$34,452,918	$179,442
DEPRECIABLE PORTION	$21,904,291	$114,085
NON DEPRECIABLE PORTION	$12,548,628	$65,357

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-6b Operating Costs - 2xProd. + Recycle

Basis: 224.6 mt algae/ha/yr
(60 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
	kg/kg	$/mt	$/ha/yr	1000 ac $/yr
NUTRIENTS (1)	-----	----	--------	-------
CO2	1.6	35	$12.58	$5,081
N, as NH3	0.013	250	0.75	302
P, as Superphosphate	0.0025	900	0.51	204
Fe, as FeSO4	0.005	500	0.56	227
			------	----------
Total			$14.39	$5,814
FLOCCULANT	0.002	5000	$2.25	907
	10 kwh/yr	c/kwh		
POWER	----------	-----		
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	3.76	6.5	0.60	244
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
	-----		-----	-----
Total	13.00		$2.09	845
Elect. Produced	-31.73	6.5	(5.11)	(2,062)
	-----		-----	-----
	-18.73		(3.01)	(1,218)
	kg/kg	c/kg		
SALT DISPOSAL	-----	----		
	1.5	0.67	1.13	457
MAINTANENCE (matl's)			2.94	1,188
LABOR			1.39	562
			-----	-----
			$16.84	$6,803
TOTAL			======	======

(1) kg/kg = kg required / kg algae produced

Table AVI-7a Capital Cost Summary - Free CO2 + Recycle

Basis: 112 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS		
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	742,055	3,865
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	1,605,083	8,360
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	787,400	4,101
Non-depreciable Capital	382,600	1,993
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	105,598	550
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	575,759	2,999
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	2,017,458	10,508
CONTINGENCY (15% of above) (1)	3,328,805	17,338
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	2,200,000	11,458
TOTAL CAPITAL COST	$28,732,680	$149,649
DEPRECIABLE PORTION	$17,304,093	$90,125
NON DEPRECIABLE PORTION	$11,428,586	$59,524

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-7b Operating Costs - Free CO2 + Recycle

Basis: 112.3 mt algae/ha/yr
(30 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
	kg/kg	\$/mt	\$/ha/yr	1000 ac \$/yr
NUTRIENTS (1)				
CO2	1.6	0	\$0.00	\$0
N, as NH3	0.013	250	0.37	151
P, as Superphosphate	0.0025	900	0.25	102
Fe, as FeSO4	0.005	500	0.28	113
Total			\$0.91	\$366
FLOCCULANT	0.002	5000	\$1.12	454
	10 kwh/yr	c/kwh		
POWER				
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	2.31	6.5	0.37	150
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
Total	11.55		\$1.86	751
Elect. Produced	-15.87	6.5	(2.55)	(1,031)
	-4.32		(0.69)	(281)
	kg/kg	c/kg		
SALT DISPOSAL	1.5	0.67	1.13	456
MAINTANENCE (matl's)			1.97	797
LABOR			1.39	562
TOTAL			\$4.70	\$1,900

(1) kg/kg = kg required / kg algae produced

Table AVI-8b Operating Costs - $70/mt CO2 + Recycle

Basis: 112.3 mt algae/ha/yr (30 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
	kg/kg	$/mt	$/ha/yr	1000 ac $/yr
NUTRIENTS (1)				
CO2	1.6	70	$12.58	$5,081
N, as NH3	0.013	250	0.37	151
P, as Superphosphate	0.0025	900	0.25	102
Fe, as FeSO4	0.005	500	0.28	113
Total			$13.48	$5,448
FLOCCULANT	0.002	5000	$1.12	454
	10 kwh/yr	c/kwh		
POWER				
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	2.31	6.5	0.37	150
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
Total	11.55		$1.86	751
Elect. Produced	-15.87	6.5	(2.55)	(1,031)
	-4.32		(0.69)	(281)
	kg/kg	c/kg		
SALT DISPOSAL	1.5	0.67	1.13	456
MAINTANENCE (matl's)			1.97	797
LABOR			1.39	562
TOTAL			$17.28	$6,982

(1) kg/kg = kg required / kg algae produced

Table AVI- 9a Capital Cost Summary - 20% Lipid Content + Recycle

Basis: 112 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS	-------	---------
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	742,055	3,865
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	1,605,083	8,360
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	1,259,840	6,562
Non-depreciable Capital	612,160	3,188
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	105,598	550
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	596,819	3,108
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	2,089,764	10,884
CONTINGENCY (15% of above) (1)	3,448,110	17,959
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	3,520,000	18,333
	---------	-------
TOTAL CAPITAL COST	$30,967,351	$161,288
DEPRECIABLE PORTION	$19,117,593	$99,571
NON DEPRECIABLE PORTION	$11,849,757	$61,717

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-9b Operating Costs - 20% Lipid Content + Recycle

Basis: 112.3 mt algae/ha/yr (30 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
				1000 ac
	kg/kg	$/mt	$/ha/yr	$/yr
NUTRIENTS (1)	-----	----	-------	-------
CO_2	0.9	35	$3.54	$1,429
N, as NH_3	0.013	250	0.37	151
P, as Superphosphate	0.0025	900	0.25	102
Fe, as $FeSO_4$	0.005	500	0.28	113
			------	---------
Total			$4.44	$1,795
FLOCCULANT	0.002	5000	$1.12	454
	10 kwh/yr	c/kwh		
POWER	---------	-----		
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	2.31	6.5	0.37	150
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
	-----		------	------
Total	11.55		$1.86	751
Elect. Produced	-25.38	6.5	(4.08)	(1,650)
	-----		------	------
	-13.84		(2.23)	(899)
	kg/kg	c/kg		
SALT DISPOSAL	-----	----		
	1.5	0.67	1.13	456
MAINTANENCE (matl's)			2.47	1,000
LABOR			1.39	562
			------	------
			$7.21	$2,913
TOTAL			======	======

(1) kg/kg = kg required / kg algae produced

Table AVI-10a Capital Cost Summary - 30% Lipid Content + Recycle

Basis: 112 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS		
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	742,055	3,865
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	1,605,083	8,360
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	1,102,360	5,741
Non-depreciable Capital	535,640	2,790
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	105,598	550
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	589,799	3,072
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	2,065,662	10,759
CONTINGENCY (15% of above) (1)	3,408,342	17,752
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	3,080,000	16,042
TOTAL CAPITAL COST	$30,222,460	$157,409
DEPRECIABLE PORTION	$18,372,703	$95,691
NON DEPRECIABLE PORTION	$11,709,367	$60,986

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-10b Operating Costs - 30% Lipid Content + Recycle

Basis: 112.3 mt algae/ha/yr
(30 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
	kg/kg	$/mt	$/ha/yr	1000 ac $/yr
NUTRIENTS (1)				
CO2	1.1	35	$4.32	$1,747
N, as NH3	0.013	250	0.37	151
P, as Superphosphate	0.0025	900	0.25	102
Fe, as FeSO4	0.005	500	0.28	113
Total			$5.23	$2,113
FLOCCULANT	0.002	5000	$1.12	454
	10 kwh/yr	c/kwh		
POWER				
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	2.31	6.5	0.37	150
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
Total	11.55		$1.86	751
Elect. Produced	-22.21	6.5	(3.57)	(1,444)
	-10.66		(1.72)	(693)
	kg/kg	c/kg		
SALT DISPOSAL	1.5	0.67	1.13	456
MAINTANENCE (matl's)			2.32	939
LABOR			1.39	562
TOTAL			$8.36	$3,376

(1) kg/kg = kg required / kg algae produced

Table AVI-11a Capital Cost Summary - 40% Lipid Content + Recycle

Basis: 112 mt algae/ha/yr

	1000 acres TOTAL $	$/HECTARE
GROWTH PONDS	-------	---------
Earthworks (1)	$4,102,083	$21,365
Walls & Structural	3,360,763	17,504
Mixing System	1,991,083	10,370
Carbonation System	742,055	3,865
Instrumentation not included elsewhere	202,368	1,054
HARVESTING - Settling Pond Option		
Primary (2)	3,027,088	15,766
Secondary (centrifugation)	1,605,083	8,360
ANAEROBIC LAGOON SYSTEM		
Depreciable Capital	944,880	4,921
Non-depreciable Capital	459,120	2,391
SYSTEM-WIDE COSTS		
Water Supply	1,391,280	7,246
Water Distribution	398,412	2,075
CO2 Distribution	105,598	550
Nutrient Supply System	316,200	1,647
Blowdown Disposal System (1)	337,280	1,757
Buildings (harv. blds. not included)	231,880	1,208
Roads & drainage (1)	210,800	1,098
Electrical Distribution (3% of above)	582,779	3,035
Electrical Supply	238,204	1,241
Machinery	168,640	878
ENGINEERING (10% of above) (1)	2,041,560	10,633
CONTINGENCY (15% of above) (1)	3,368,573	17,545
LAND COSTS ($1250/hectare) (1)	1,011,840	5,270
ELECTRICAL GENERATOR (3)	2,640,000	13,750
	---------	--------
TOTAL CAPITAL COST	$29,477,570	$153,529
DEPRECIABLE PORTION	$17,908,593	$93,274
NON DEPRECIABLE PORTION	$11,568,977	$60,255

(1) non-depreciable item
(2) patrially non-depr. ($375,000)
(3) firm bid for turnkey system - engr. & contingency not required
112 mt/ha/yr = 30 g/m2/day
Land area = 2 x growth pond area

Table AVI-11b Operating Costs - 40% Lipid Content + Recycle

Basis: 112.3 mt algae/ha/yr
(30 gm/sq m/day)

	QUAN REQ'D	UNIT COST	YEARLY COST, Thousands	
	kg/kg	$/mt	$/ha/yr	1000 ac $/yr
NUTRIENTS (1)				
CO2	1.35	35	$5.31	$2,144
N, as NH3	0.013	250	0.37	151
P, as Superphosphate	0.0025	900	0.25	102
Fe, as FeSO4	0.005	500	0.28	113
Total			$6.21	$2,510
FLOCCULANT	0.002	5000	$1.12	454
	10 kwh/yr	c/kwh		
POWER				
Mixing	4.34	6.5	0.70	282
1 Harvesting	0.72	6.5	0.12	47
2 Harvesting	2.31	6.5	0.37	150
Water Supply	3.53	6.5	0.57	230
Nutrient Supply	0.22	6.5	0.04	14
Buildings	0.42	6.5	0.07	27
Total	11.55		$1.86	751
Elect. Produced	-19.04	6.5	(3.06)	(1,238)
	-7.49		(1.21)	(487)
	kg/kg	c/kg		
SALT DISPOSAL	1.5	0.67	1.13	456
MAINTANENCE (matl's)			2.17	878
LABOR			1.39	562
TOTAL			$9.70	$3,919

(1) kg/kg = kg required / kg algae produced

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15. Supplementary Notes

Technical Monitor: R. McIntosh

16. Abstract (Limit: 200 words) The designs and systems developed include many innovative concepts and experiments, including the design and operation of a low-cost system. Cost-effectiveness is realized by minimizing capital costs of the system and achieving efficient use of inputs. Extensive engineering analysis of carbonation, mixing, and harvesting subsystems has elucidated both the lowest cost, most efficient options and the essential parameters needed to construct, test, and evaluate these subsystems. The use of growth ponds sealed with clay and lined with crushed rock results in construction cost savings of 50% over ponds lined with synthetic membranes. In addition a low-cost but efficient design allows improvments in technology to have maximum impact on final product cost reductions. In addition to the innovations in low-cost construction, the operational efficiency of the design is both higher and more feasible than that attained by any previous system concept of comparable scale. The water chemistry analysis has led to operational specifications that minimize water use and virtually eliminate losses of carbon dioxide to the atmosphere. The carbon dioxide injection system is designed for 95% efficiency, but is still low in cost. The construction of a large-scale, covered anaerobic lagoon to recycle carbon, nitrogen, and phosphorus has not been attempted at the scale analyzed here. Yet efficient recycling is essential for achieving economic affordability.

17. Document Analysis

a. Descriptors

Biomass ; Carbon Dioxide ; Cultivation ; Cultivation Techniques ; Ponds ; Productivity ; Unicellular Algae

b. Identifiers/Open-Ended Terms

c. UC Categories

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