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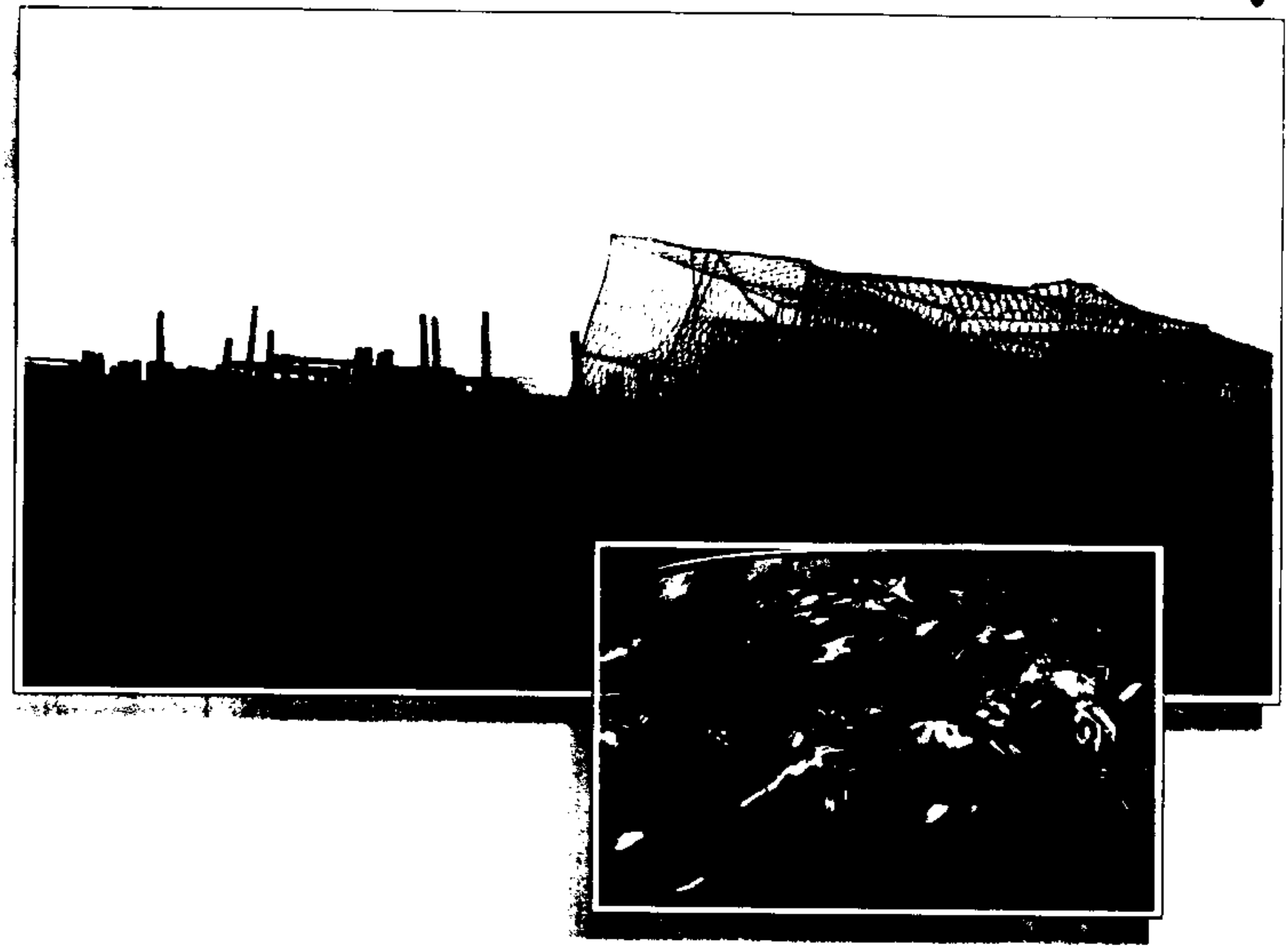
Waste Management and Byproducts Recovery for the Blue Crab Industry



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Waste Management and Byproducts Recovery for the Blue Crab Industry

The book characterizes and suggests treatments of crab processing wastewaters and possible uses for the byproducts. It includes waste treatability studies; flavor and pigment extraction from the blue crab processing byproducts; collection, analysis, and utilization of biogas generated by the anaerobic treatment of crab processing wastewater; removal of nutrients and ammonia from crab processing waters; and USAB (Upflow Anaerobic Sludge Blanket) and air stripping treatment technologies. Published by Virginia Tech in concert with the Virginia Sea Grant College Program and the Virginia Cooperative Extension. Authored by Gregory D. Boardman, George J. Flick, and associates, 1995. 142 pp.

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Waste Management and Byproducts Recovery for the Blue Crab (*Callinectes sapidus*) Industry

Part I: Waste Treatability Studies

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1.0 INTRODUCTION

1.1 Background

As environmental legislation becomes more stringent, the blue crab industry is being challenged to develop technically and economically feasible methods to manage their liquid and solid wastes. The blue crab industry is one of the largest seafood processing industries in Virginia. More than 50 blue crab processing firms are presently certified by the Bureau of Shellfish Sanitation, Virginia Health Department (Croonenberghs, 1991).

The industry is economically vital to many communities of the Chesapeake Bay and its tributaries. In 1989, there was a total landing of 206.7 million pounds of blue crab in the United States valued at 81 million dollars. The Chesapeake region produced 89 million pounds of this total (U.S. Department of Commerce, 1990).

In the 1970's, the United States Environmental Protection Agency (EPA) developed documents which defined effluent limits for segments of the seafood processing industry. The regulations were then adopted by the states. Compliance with these limits was obtained, in most cases, with minor plant modifications that did not require large capital investments and operating costs. During the last decade, citizen and environmental groups have encouraged more stringent legislation in an effort to achieve a cleaner environment. Individual states have established waste disposal standards that exceed current EPA requirements.

The restoration of water quality in the Chesapeake Bay has become a goal of states which border the Bay and its tributaries. In Virginia, the State Water Control Board (SWCB) has initiated a vigorous program to remove pollutants from the Bay. As a result, new standards for the disposal of liquid and solid wastes have been established. The new standards, many of which substantially exceed previous standards, have resulted in: the loss of public landfills; highly regulated waste transportation regulations; the inability of municipal systems to accept industrial wastes; waste disposal surcharges; and perhaps most importantly, new liquid effluent standards for discharge into receiving waters. Many blue crab processors have reported serious waste disposal problems as they are unable to consistently comply with their current allowable pollutant discharge limits.

Waste management is difficult for blue crab processing firms for several reasons. In general, crab firms are less capitalized and do not have the economics of scale when compared to some other seafood industries. Treatment systems with the ability to produce the desired effluent quality may not be economically feasible. Chemical, biological and physical treatment may be required to reduce the strength of processing plant effluents (Geiger et al., 1985; Wheaton et al., 1984). These treatment processes will be expensive and require adequate planning to avoid unfavorable economic impact.

Secondly, the location of processing plants also creates problems. Most blue crab processing facilities are located in rural areas on bodies of water. These facilities usually do not have access to municipal waste treatment plants. Land application and lagoon treatment are often not viable treatment options due to the high water table and wetland limitations. Some larger crab plants are located in cities with access to municipal treatment systems. However, these plants are often assessed substantial surcharges unless the concentration of their waste is reduced. There is also a possibility that municipal treatment systems may be forced to reject crab wastes due to the rapid population and industrial growth that is pushing municipal waste treatment systems to maximum operating capacity. Furthermore, plants located in metropolitan areas often lack the space needed for treatment systems. The plants are usually tightly bordered by water, parking lots and/or neighboring industries.

Finally, in Virginia, because the SWCB is in a period of transition in implementing new standards, different crab processing firms may be regulated on the basis of different effluent constituents. This discrepancy can even appear in two adjacent crab plants. The inconsistency of the regulations makes it difficult for the industry as a group to address its waste disposal alternatives.

The blue crab processing industry faces serious solid and liquid waste disposal problems. Approximately 14% of the live crab is used as food for human consumption, with the remainder as by-products or waste (EPA, 1974). Consequently, processing firms will need to develop in-plant programs that will include: water conservation and recovery processes, improved by-product recovery systems and the development of industrial products from wastes (including foods, feeds, or biologics). Solutions to the waste management problems are essential for the future of many blue crab processing facilities. Some of the solutions will require the application of innovative technology, while others will result in substantially increased production costs.

The achievement of such a goal presents substantial economic and technical obstacles and may be one of the greatest challenges in the future.

1.2 Objectives

The objectives of this project were to characterize process wastewaters, identify ways to reduce waste quantities and strengths, and determine viable waste treatment methods for the blue crab processing industry.

2.0 LITERATURE REVIEW

2.1 Crab Processing and Waste Generation

The crab processing industry is the fourth largest among America's fishing industries; approximately one quarter of all crabs caught in the United States are taken from the Chesapeake Bay region (Warner, 1977). This concentration of the industry translates to a significant amount of regional economic activity. However, as with almost any industry of this size, there are associated problems which must be addressed.

One of the problems, which has received increasing amounts of attention in recent years, is the pollution associated with processing crabs. Currently, most processing plants discharge their wastewaters directly to a surface water, such as a river or estuary. Pollutive effects of the processing activities observed in receiving water bodies include odor problems, increased oxygen demand, solids accumulation, and excess nutrient levels (EPA, 1980). While the impact of odor problems is limited primarily to the aesthetic quality of the receiving body, the last three problems have the potential to impact the supporting capacity of the water.

Increases in the oxygen demand of a water body are the result of bacterial decomposition of organics present in the wastewater. This increased demand can reduce the dissolved oxygen content of the water which would otherwise be available to support various forms of sealife.

The accumulation of solids on the bottom of the receiving water can lead to the smothering of both plant and animal species which would normally contribute to the quality of the water above. The introduction of screens and other solids recovery steps on most processes has enabled the industry to minimize the discharge of solids and thereby decrease the effects associated with such discharges. The presence of excessive levels of nutrients can often have a negative impact on the receiving water's quality. These effects can include the development of ammonia toxicity problems, algal blooms, and increased NO_2 and NO_3 levels.

The susceptibility of aquatic animals to ammonia toxicity has been found in various studies to be a function of species, as well as water salinity, pH, and temperature. These effects can be seen in the data from three studies presented in Table 1 (Hazel et al., 1971; Diamond et al., 1993; Herbert and Shurben, 1965).

Table 1. Aquatic species ammonia toxicity test data.

Species	Water Type	pH	Temp °C	Unionized NH_3 Conc. mg/L		
				TLm	LC50	NOEC
Bluegill[1]	Fresh	8.0	20	-	1.02	0.29
Bluegill[1]	Fresh	8.0	12	-	0.53	0.36
Bay Silverside[1]	Fresh	8.0	20	-	1.13	0.78
Bay Silverside[1]	Fresh	8.0	12	-	0.73	0.23
Striped Bass[2]	Fresh	7.5	15	1.36	-	-
Striped Bass[2]	Brackish	7.5	15	1.36	-	-
Striped Bass[2]	Sea	8.0	15	0.97	-	-
Striped Bass[2]	Fresh	7.5	23	0.92	-	-
Striped Bass[2]	Brackish	7.5	23	1.02	-	-
Striped Bass[2]	Sea	8.0	23	0.73	-	-
Stickleback[2]	Fresh	7.5	15	1.02	-	-
Stickleback[2]	Brackish	7.5	15	2.53	-	-
Stickleback[2]	Sea	8.0	15	5.05	-	-
Stickleback[2]	Fresh	7.5	23	0.87	-	-
Stickleback[2]	Brackish	7.5	23	1.17	-	-
Stickleback[2]	Sea	8.0	23	1.12	-	-
Rainbow Trout[3]	Fresh	7.5	13.6	0.55 *	-	-
Rainbow Trout[3]	Brackish	7.5	13.6	1.27 *	-	-
Rainbow Trout[3]	Sea	7.5	13.6	0.71 *	-	-

TLm = 96 hr Mean Tolerance Limit LC50 = 50% Mortality Conc.
NOEC = No Observed Effect Concentration
* = 24 hr TLm NH_3 Conc. (converted from total NH_4Cl-N Conc.)
1 = Hazel et al., 1971; 2 = Diamond et al., 1993;
3 = Herbert and Shurben, 1965.

While toxicity can arise from both the ionized (NH_4^+) and the molecular unionized (NH_3) forms, research has identified the second of these as having the greater toxic effect. The EPA has implemented a water quality criterion of 0.02 mg/l NH_3 (Thurston et al., 1979).

2.2 Effluent Discharge Regulations

Point source discharges resulting from blue crab processing are governed by federal, state and, occasionally, local regulations. Permits, that set limits on the quantity and quality of the wastewater being discharged, are required. These limits are typically based on maximum daily and/or average monthly concentrations and/or loadings. If wastewater is discharged indirectly to a publicly owned treatment works (3POTW), the POTW may specify wastewater quantity and quality conditions and apply surcharges on wastewater volumes, rates and pollutant concentrations. Processing plants indirectly discharging to POTWs are required to comply with the pretreatment standards set forth in the National Pretreatment Program (40 CFR 403).

2.2.1 Federal Effluent Discharge Regulations

The Federal Water Pollution Control Act Amendments (FWPCA) of 1972, required National Pollutant Discharge Elimination System (NPDES) permits for wastewater discharges from most industrial categories including the "canned and preserved seafood processing" industry. A national goal of "zero discharge" to navigable waters by 1985 was established in the Act. Technology-based interim guidelines were set forth as a means of working toward achievement of the zero discharge goal.

The first set of interim guidelines was based on the best practicable control technology currently available (BPT), which went in effect in 1975 for blue crab processors. A second, more stringent set of guidelines based on the best available technology economically achievable (BAT) was to be implemented by July 1, 1983.

BAT standards were later considered too stringent for industries discharging conventional pollutants. Amendments to the FWPCA provided for more lenient standards for these industries and were based on the effluent quality attainable by the application of the best conventional pollutant control technology currently available (BCT). Current BCT levels are identical to the original BPT levels for the blue crab processing industry.

Federal discharge permit limits for the blue crab processing industry are differentiated by the following criteria:

- conventional (manual) blue crab processing versus mechanical processing
- existing point source versus new point source
- direct offsite discharge versus indirect discharge to a POTW

Federal effluent guidelines and standards for conventional and mechanical blue crab processing plants are shown in Table 2 and Table 3, respectively, as of June 1993. These limits are technology-based and are listed in the *Code of Federal Regulations, Title 40, Part 408, Subpart B and C.*

2.2.2 Virginia Effluent Discharge Regulations

The FWPCA requires that the States adopt and periodically revise water quality standards for state waters and empowers them to develop and enforce more stringent effluent limits to meet these standards. Virginia has a federally approved effluent permit program called the Virginia Pollutant Discharge Elimination System (VPDES) which is enforced by the Virginia State Water Control Board (SWCB). These effluent permit limits are often more stringent than federal limits and are based on federal effluent requirements, Best Professional Judgement (BPJ), water quality standards, water quality models, water quality management plans, etc.

The SWCB currently regulates temperature, pH, BOD_5, TSS, and O&G for blue crab processing industry effluent. Currently, criteria that determine the type and degree of pollutant regulation depend on the facility process (conventional versus mechanical), age, location and production rate. Each discharge permit is developed on a case-by-case basis and as a result, discharge permit limits differ between plants throughout Virginia.

Permits are valid for a five year period, but the state has the authority to make the permit limits more stringent even before the permit expires. In addition, the SWCB can prevent "backsliding" of effluent quality in situations where an facility has shown it can achieve better effluent quality than is required by past permits.

The SWCB is considering promulgating general permit discharge regulations for molluscan shellfish and crustacea processing establishments (VR 680-14-10). The general permit would apply to dischargers that, "(i) involve the same or similar types of operations;

Table 2. Federal effluent guidelines and standards conventional blue crab processing subcategory[1] 40 CFR 408 - Subpart B.

	EXISTING SOURCE			NEW SOURCE		
	Direct Discharge		Indirect Discharge	Direct Discharge		Indirect Discharge
	Max[3]	Avg[4]		Max[3]	Avg[4]	40 CFR 403
BOD_5	no limit	no limit	no limit	0.30	0.15	
TSS	2.2	0.74	Do.	0.90	0.45	
O&G	0.60	0.20	Do.	0.13	0.065	
pH	(2)	(2)	no limit	(2)	(2)	

Note: All units, except pH, are lb/1,000 lb raw seafood processed

(1) Applies to existing facilities manually processing more than 3000 lbs of raw seafood on any day during the calendar year and all new sources

(2) Within the range 6.0-9.0

(3) Maximum for any one day

(4) Average of daily values for 30 consecutive days

(5) Set by POTW, within approved Pretreatment Program

Table 3. Federal effluent guidelines and standards mechanized blue crab processing subcategory(1), 40 CFR 408 - Subpart C.

	EXISTING SOURCE			NEW SOURCE		
	Direct Discharge		Indirect Discharge	Direct Discharge		Indirect Discharge
	Max(3)	Avg(4)		Max(3)	Avg(4)	40 CFR 403
BOD_5	no limit	no limit	no limit	5.0	2.5	
TSS	36	12.0	Do.	13	6.3	
O&G	13	4.2	Do.	2.6	1.3	
pH	(2)	(2)	no limit	(2)	(2)	

All units, except pH, are lb/1,000 lb raw seafood processed
(1) Applies to all facilities in which mechanical picking or separation of crab meat from the shell is utilized
(2) Within the range 6.0-9.0
(3) Maximum for any one day
(4) Average of daily values for 30 consecutive days
(5) Set by POTW, within an approved Pretreatment Program

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(ii) discharge the same or similar types of waste; (iii) require the same effluent limitations or operating conditions; and (iv) require the same or similar monitoring." The general permit is intended for small processing facilities and would assign the same effluent limits for all effected plants. The SWCB is also currently considering setting effluent ammonia discharge limits in addition to the parameters already regulated.

2.3 Blue Crab Processing Overview

The typical steps for processing live blue crabs and the wastes generated by each process are shown in Figure 1. The unrefrigerated live crabs are usually delivered by boat or truck to processing plants. The crabs are weighed, then dumped into large stainless steel baskets. During the winter dredging season, the crabs are directed through a tumble spray washer prior to being dumped into baskets. This washing step is essential for dredged crabs because they are covered with sand and grit from being buried in sand.

After washing, the crabs are placed into horizontal or vertical retorts, and cooked by steaming for 7 to 23 minutes at 121°C and 15 psig (Phillips and Peeler, 1972). In this paper the cooker effluent will be called "retort effluent." The main objectives of cooking are to facilitate removal of meat from the shell, give the product a characteristic crab meat odor and flavor, and reduce microbial populations.

Following cooking, the crabs are moved to a room where they are air-cooled to ambient temperatures within thirty minutes. Before the cooled crabs are moved to the cooler (33°F to 40°F), they must be cool enough so that steam is no longer rising from them. If cooked crabs were moved immediately to the cooler, steam rising from the crabs would condense on the ceiling of the cooler and drip back down on the crabs. This could potentially contaminate the cooked crabs (Ulmer, 1964; Wentz, 1985).

Two different processes are used to prepare the crabs for removal of the meat (picking): a wet process and a dry process. In the wet process, crabs are backed (carapace removed) and declawed first. The crab bodies are then washed by hand or machine and the meat is removed immediately, or the bodies may be refrigerated overnight. The dry process, most commonly used in the Chesapeake Bay area, does not include the washing step. Each picker backs, declaws, and removes all meat from each crab (Cockey, 1980).

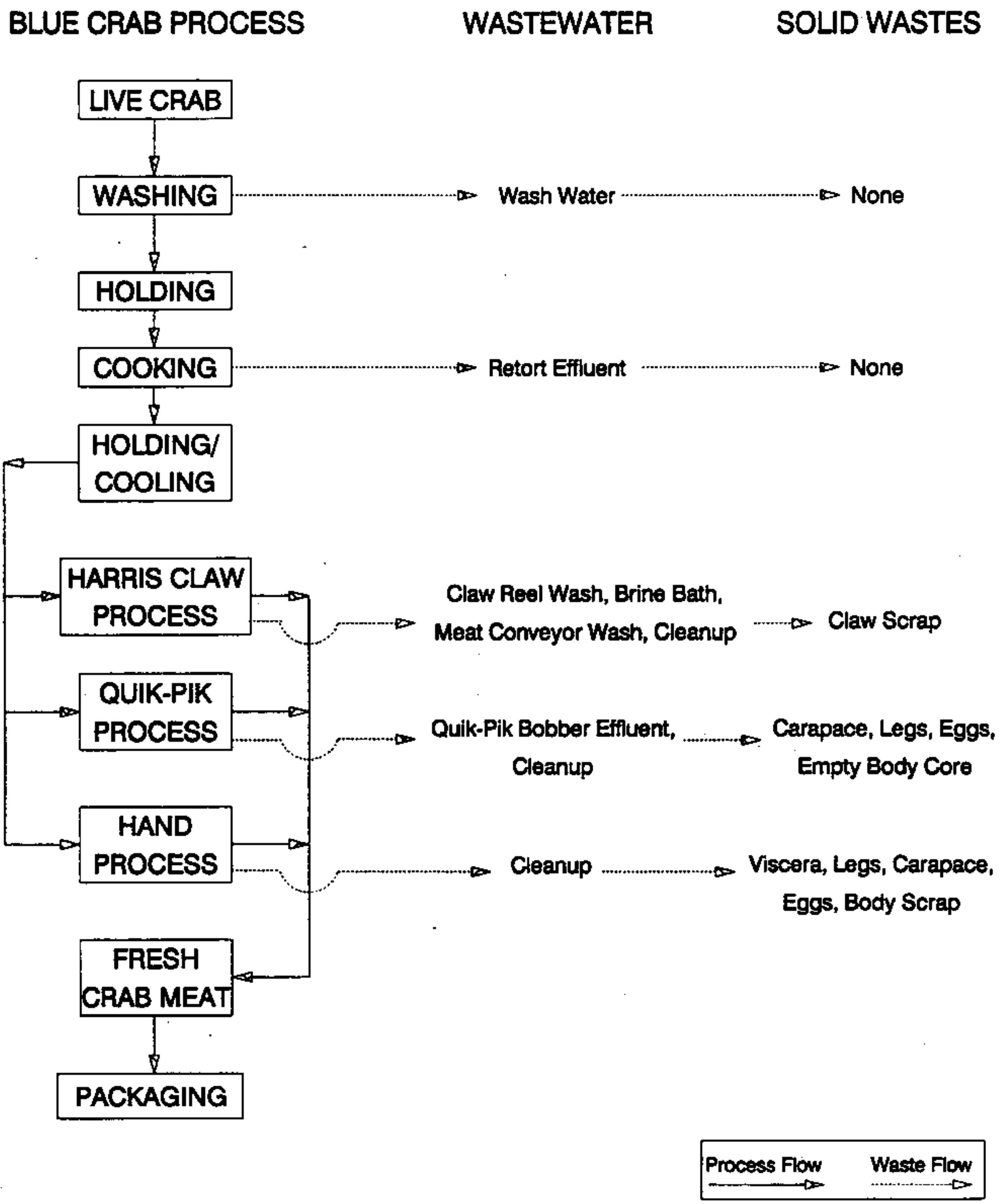


Figure 1. Schematic of blue crab process product and waste flows.

Picking meat from the crab is labor intensive and is sometimes supplemented in a few large plants by mechanized picking. The Harris Claw machine combines hammer mills with brine-tank flotation to remove the meat from claws. The Quik-Pik machine uses high speed vibration to remove body meat from the crab (Cockey, 1980; Wenstrom, 1978). The Quik-Pik process may also include a bobber which mechanically cuts off the legs, removes the carapace and then removes the viscera from the remaining core using spray jets and brushes. The bobber produces a concentrated effluent. The wastewater produced by the Quik-Pik process will be called "Quik-Pik bobber liquid."

After the picking process, the meat goes through a deboning step in which any remaining shell fragments are removed from the meat by hand. The deboned meat is then packed fresh, frozen or pasteurized.

Several different effluent streams were produced in the Harris Claw process. These streams were characterized separately and are named as follows: claw wash reel, shell liquid, brine bath and claw meat conveyor wash. At the start of the process, the claws are usually washed in a tumble spray process resulting in the "claw wash reel" effluent. The claws are then shattered as they pass through a hammer mill and the resulting meat and shell fragments fall into a tank of approximately 70% saturated brine solution. Due to the differing specific gravities between the meat, shell fragments and brine solution, the meat floats to the top of the tank while the shell fragments settle to the bottom. The shell pieces are carried out of the tank by a conveyor where they empty into a perforated receptacle. The shell conveyor carries a large quantity of the brine solution out with the shell fragments and is responsible for the "shell liquid" effluent. The perforations in the receptacle allow the liquid to run out leaving only the shell.

The meat that floats to the top of the tank is carried off by another conveyor belt where it is washed with potable water through a spray nozzle to reduce the salt content. Meat on the conveyor is also checked by hand to ensure removal of any remaining shell bits. This "claw meat wash water" provided a third effluent stream from the Harris Claw process. The final effluent from this process is the brine bath. The brine tank is filled only once or twice during a processing day, but is continually being refilled as solution spills over in the tank or is carried out by the conveyor belts.

2.4 Wastewater Characterization

Little published data exists on the characterization of wastewater discharged from conventional and mechanized blue crab processes. The first significant characterization study was completed in the early 1970s by the EPA in an effort to determine effluent quality for the development of effluent limitations under the FWPCA (Environmental Protection Agency, 1974). In fact, this study represented the only detailed study of mechanical blue crab processing presented in the literature. The results of the study are printed in the *Development Document for Effluent Limitations Guidelines and New Source Performance Standards for the Catfish, Crab, Shrimp and Tuna Segment of the Canned and Preserved Seafood Processing Point Source Category* (EPA, 1974). A detailed discussion of this document is appropriate because it is the basis for the development of federal BPT/BCT effluent limitations and guidelines for the blue crab processing industry.

In this study, two conventional and two mechanized processing plants in the Chesapeake Bay were examined. General conclusions made from this study were as follows:

- conventional blue crab processing uses only 1/10th the water of other crab operations (i.e. - mechanized blue crab, Dungeness, Tanner and King crab processing)
- mechanized blue crab processing produced double the BOD_5 of other crab species
- mechanized blue crab processing generated wastewater in significantly greater quantity and BOD_5 than conventional blue crab processing.

For the reasons listed above, the mechanized and conventional blue crab processing industries were given separate subcategories and standards in the FWPCA.

Results from the study on conventional blue crab processing facilities revealed the following: an average of 5,700 pounds of raw seafood processed each day, an average effluent flow of 660 gal/day distributed as approximately 60% ice making cooling water, 23% cleanup water and 17% cooker effluent, and an average of 142 gallons of water was used per 1,000 pounds of raw crab processed. The combined effluent quality averages were as follows: BOD_5 = 4,400 mg/L, TSS = 620 mg/L, O&G = 220 mg/L, TKN-N = 760 mg/L and NH_3-N = 50 mg/L.

In contrast, the mechanized blue crab processing facilities (an average of 10,600 lbs raw seafood/day processed) exhibited an average effluent flow of 46,500 gal/day distributed as approximately 90.5% machine picking water, 7.7% washdown water, 1.1% ice making cooling water, 0.5% brine tank and 0.2% cooker effluent. An average of 4,415 gallons of water was used per 1,000 pounds of raw crab processed. The combined effluent quality averages were as follows: BOD_5 = 600 mg/L, TSS = 330 mg/L, O&G = 150 mg/L, TKN-N = 760 mg/L and NH_3-N = 5.4 mg/L. In addition, chloride concentrations ranged from 100,000-200,000 mg/L in the brine separation tanks. The brine tanks held 275 gallons of wastewater and were dumped each shift.

Table 4 shows a detailed comparison between the mechanized and conventional blue crab processes in the EPA study. The mechanized processes used approximately 30 times more water than the conventional operation to produce the same amount of crab product. The mechanized process also exhibited a greater flow variation. The effluent pollutant concentrations were more dilute in the mechanized processes but pollutant loadings were higher since significantly larger volumes of water were discharged.

John Riley of the EPA noted that 264 gallons of wastewater were produced per 1,000 pounds of live crab versus 7,530 gallons per 1,000 pounds in the mechanized process (Riley, 1980).

Characterization of conventional blue crab processing wastewater was done by others (Table 4). Brinsfield et al. (1977), characterized six conventional plants which exhibited the following averages and standard deviations (based on 74 samples): production rate = 4,400 ± 1,560 lbs live crab/day, BOD_5 = 2.67 ± 7.81 lbs/1,000 lbs, TSS = 1.92 ± 4.26 lbs/1,000 lbs, O&G = 0.04 ± 0.07 lbs/1,000 lbs, TKN-N = 0.27 ± 0.66 lbs/1,000 lbs, NH_3-N = 0.04 ± 0.13 lbs/1,000 lbs and phosphorus = 0.04 ± 0.08 lbs/1,000 lbs (all units are per 1,000 pounds of live crab processed). A brine solution was used to separate meat from shell in the operation and chloride concentrations ranged from 100,00-200,000 mg/L. Rubin (1983), collected six samples from a conventional plant and found the following averages and standard deviations: flow = 3,000 gal/day, COD = 968 ± 119 mg/L, TKN-N = 1330 ± 113 mg/L, NH_3-N = 119 ± 11 mg/L and total phosphorous = 69 ± 9 mg/L. After in-plant changes, the average daily flow was reduced to 600 gal/day. No data were available on the corresponding change in effluent pollutant concentration.

Table 4. Conventional and mechanized blue crab wastewater pollutant characterization found in published literature.

	CONVENTIONAL PROCESSES					MECHANICAL PROCESSES	
	EPA Development Document 1974		Brinsfield et al. 1977	Overcash et al. 1980	Rubin 1983	EPA Development Document 1974	
Type	Average	Range	Average	Average	Average	Average	Range
No. of samples	18	18	74	-	6	7	7
No. of plants studied	2	2	6	-	1	2	2
Production (lb/day)	5700	-	4400+/-1560	-	-	-	-
Flow (gal/day)	665	(630-700)	-	-	3000	46500	(20000-73000)
Flow (gal/1000 lb)	142	(128-158)	765+/-69	-	-	4415	(3480-5350)
BOD-5 (mg/L)	4400	-	-	-	-	600	-
BOD-5 (lb/1000 lb)	5.2	(4.8-5.5)	2.67+/-7.81	-	-	22	(22-23)
COD (mg/L)	6300	-	-	1100	968+/-119	980	-
COD (lb/1000 lb)	7.5	(7.2-7.8)	-	-	-	36	(29-42)
TSS (mg/L)	620		-	-	-	330	-
TSS (lb/1000 lb)	0.74	(0.70-0.78)	1.92+/-4.26	-	-	12	-
O&G (mg/L)	220	-	-	-	-	150	-
O&G (lb/1000 lb)	0.26	(0.21-0.30)	0.04+/-0.07	-	-	5.6	(4.3-6.9)
TKN-N (mg/L)	760	-	-	55	1330+/-113	98	-
TKN-N (lb/1000 lb)	0.90	(0.80-1.0)	0.27+/-0.66	-	-	3.6	(2.7-4.4)
NH3-N (mg/L)	50	-	-	-	119+/-11	5.4	-
NH3-N (lb/1000 lb)	0.06	-	0.04+/-0.13	-	-	0.02	(0.16-0.24)
Phosophorus (mg/L)	-	-	-	10	69+/-9	-	-
Phosophorus (lb/1000 lb)	-	-	0.04+/-0.08	-	-	-	-
pH	7.5	(7.2-7.9)	-	-	-	7.0	(6.9-7.2)

Overcash (1980) found conventional plant effluent concentrations of 1,100 mg/L for COD, 55 mg/L for TKN-N, and 10 mg/L for total phosphorous.

An extensive wastewater characterization of the Dungeness crab processing industry was conducted by Soderquist et al. (1972). The EPA (1974) determined that significant differences existed in the processing methods and effluent quality between blue crab and Dungeness crab. For these reasons, separate discharge requirements were developed for each industry.

In the Dungeness crab study, three conventional processing plants were monitored for a three month period with seven to fourteen flow-proportioned samples taken from each operation. The combined effluent quantity and quality from the most intensively studied plant was as follows (production rate = 12,000-16,000 lbs live crab/shift): flow ratio = 1920 gal/1,000 lbs, BOD_5 = 6.8-13.5 lbs/1,000 lbs, COD = 11.5-22.9 lbs/1,000 lbs, TSS = 1.7-5.0 lbs/1,000 lbs, TKN-N = 1.4-2.8 lbs/1,000 lbs, NH_3-N = 0.13-0.24 lbs/1,000 lbs and total phosphorus = 0.08-0.33 lbs/1,000 lbs (all units per 1,000 pounds of live crab processed). The study concluded that the major sources of pollutant loads, in order of influence, resulted from the "brining step," cooking, and butchering/bleeding, respectively. The brining step was the major source of COD, suspended solids, total nitrogen, total phosphorus and chloride. The cooking step was the major contributor of ammonia and the second largest contributor of COD, suspended solids and total nitrogen in the combined effluent stream.

Characterization of blue crab cooker effluent has been conducted by a few researchers because of its relatively high pollutant concentration and potential for by-product recovery (Table 5). Chao et al. (1983) characterized the following cooker effluent ranges: flow ratio = 25-50 gal/1,000 lbs live crab cooked, BOD_5 = 10,000-14,000 mg/L, COD = 20,000-25,000 mg/L, TSS = 700-1,000 mg/L, and NH_3-N = 200-250 mg/L.

A study of cooker effluent from six plants showed an average 9,000 mg/L of BOD_5 and 1,500 mg/L of TSS (Wheaton et al., 1981).

In a third study, examination of cooker effluent from ten separate cooks at 121°C produced wastewater with an average of 16,557 mg/L BOD_5 and 55,568 mg/L COD (Hanover et al., 1975).

Table 5. Blue Crab Cooker Effluent Pollutant Characterization Found in Published Literature

Type	Chao et al. 1983 Range	Wheaton et al. 1980 Average	Hanover et al. 1974 Average
No. of samples	-	-	10
No. of plants studied	-	6	-
Flow (gal/1000 lb)	25-50	-	-
BOD_5 (mg/L)	10000-14000	9000	16557
COD (mg/L)	20000-25000	-	555688
TSS (mg/L	700-1000	1500	-
NH_3-N (mg/L)	200-250	-	-
pH	7.0-7.5	-	-

2.5 Wastewater Treatment

The NPDES regulates the discharge of blue crab processing wastewater into waters of the United States. As described in section 2.1., NPDES permits set limits on the allowable discharge concentrations and/or mass loadings of BOD_5, TSS, O&G and sometimes nutrients. A common method of attaining the requirements of these permits is through "end-of-pipe" physical, chemical and/or biological wastewater treatment. Review of the literature revealed that suspended solids removal and biological treatment methods constituted the majority of the treatment studies concerning blue crab processing wastewater.

The EPA described effluent discharge limits for the conventional and mechanized blue crab processing industry and the treatment technologies that were expected to achieve those limits in the *Development Document for Effluent Limitations Guidelines and New Source Performance Standards for the Catfish, Crab, Shrimp and Tuna Segment of the Canned and Preserved Seafood Processing Point Source Category* (EPA, 1974). To achieve the 1977 BPT guidelines for TSS and O&G, screens and simple grease traps were expected to be required for both conventional and mechanized processes. To achieve the 1983 BCT guidelines which included BOD_5 limits, aerated lagoon technology was anticipated to be required for both processes. In addition, mechanized processes would require in-plant modifications to meet limits.

The USEPA report predicted that in theory, aerated lagoons would treat conventional blue crab processing effluent to 125 mg/L BOD_5 and 375 mg/L TSS while extended aeration would produce effluent quality of 100 mg/L BOD_5 and 100 mg/L TSS. For mechanical processing effluent, aerated lagoons were expected to achieve an effluent BOD_5 of 80 mg/L and TSS of 200 mg/L, while extended aeration would produce an effluent quality of 60 mg/L BOD_5 and 60 mg/L TSS.

Zachritz and Malone (1991) estimated the removals of BOD_5, TSS and ammonia from conventional blue crab processing wastewater. Their estimates were based on theoretical waste minimization practices and typical process treatment efficiencies listed in *Wastewater Engineering: Treatment/Disposal/Reuse* by Metcalf and Eddy, (1979). The estimated treatment efficiencies were based on effluents other than blue crab processing wastewater. They estimated an effluent quality of 5 mg/L BOD_5, 5 mg/L TSS and 12 mg/L NH_3-N could be achieved by a combination of waste minimization (solids separation and washwater reduction), primary treatment (simple screens and sedimentation), secondary treatment (aerobic lagoon or extended aeration or DAF), and polishing (rock-reed filter followed by slow sand filtration with nitrification, or rapid sand filters followed by selective ion exchange).

2.5.1 Suspended Solids Removal

Wheaton et al. (1981) studied the ability of mechanical screening, settling, flocculation with alum and chitosan, centrifugation, and foam fractionation to meet conventional blue crab processing wastewater discharge limits. None of the technologies were able to meet BOD_5 or TSS permit limits for conventional blue crab process wastewater.

Mechanical screening through 3 μm filter paper reduced total suspended solids by about 30% and BOD about 15%. Filtering through 0.45 μm filter paper removed only slightly more BOD and TSS. Settling yielded no significant removal of BOD or solids because particles were small with low density.

Coagulation and flocculation with alum (2.5-4.0 gm/L) and chitosan (0.2-0.4 gm/L) were ineffective at reducing BOD or suspended solids.

In addition, neither centrifugation at 15,000 g's for 30 minutes or foam fractionation adequately removed BOD or suspended solids to required levels.

Johnson and Gallanger (1984) studied the removal of TSS from Tanner crab processing wastewater using the coagulants, ferric sulfate and chitosan. Jar test studies revealed an optimum dose of 10 mg/L of chitosan or 500 mg/L of ferric sulfate removed greater than 90% of the TSS. The ferric sulfate addition consistently produced a better effluent than chitosan addition, but resulted in a greater pH drop. The study also investigated the use of hydrocyclones to remove the TSS. A pilot-scale hydrocyclone achieved a 71% average reduction in TSS concentration in tanner crab process wastewater.

The use of ultrafiltration to treat blue crab cooker effluent and recover by-products was studied by Chao et al. (1983). Using a membrane filter cartridge with a 50,000-dalton molecular weight cutoff, the effluent parameters were reduced as follows: COD reduced by 65%, BOD_5 reduced by 67%, and TSS reduced by 99%.

Physical/chemical treatment alternatives for shrimp processing wastewater were examined by Hudson and Pohland (1975). Coagulation and flocculation were effective for removal of suspended solids, but not for removal of organics. Activated carbon treatment of clarified wastewater was effective for removal of organics, but was not considered practical because large quantities of carbon were required for treatment of the high strength wastewater.

The use of dissolved air flotation (DAF) for the treatment of shrimp processing wastewater was examined by Szabo et al. (1979). The researchers concluded that the effective use of DAF required knowledgeable operation and control. Chemicals for pH control and for coagulation and flocculation of the suspended solids were also required. The DAF treatment was not able to achieve the effluent BOD_5 and TSS loadings required by BAT guidelines for shrimp processing wastewater.

2.5.2. Biological Treatment

Wheaton et al. (1981) conducted bench-scale anaerobic and aerobic treatability tests of conventional blue drab processing wastewater. Four weeks of non-stirred anaerobic digestion resulted in a 60% BOD reduction at 10-15°C and a 20% BOD reduction at 5-10°C.

Aerobic treatment of conventional blue crab processing wastewater resulted in 70% BOD reduction after three weeks at 1-5°C and six days at 25-30°C. TSS concentrations were not significantly reduced in this study. Foaming resulting from aeration was a problem. Influent BOD (21 samples) averaged 795 mg/L and the effluent BOD (25 samples) averaged 266 mg/L.

Wheaton et al. (1984) performed a pilot-scale, aerobic, biological treatability study of conventional blue crab processing wastewater using a 7,600 gallon vinyl-lined swimming pool. Blue crab processing effluent was periodically pumped from a sump to the treatment tank for a period of one year. The hydraulic retention time in the basin was approximately 2 days. The influent BOD_5 averaged 753 mg/L and the effluent BOD_5 averaged 278 mg/L (32 samples). The influent and effluent TSS concentrations were not significantly different.

Geiger et al. (1985) examined aerobic treatment of blue crab processing wastewater with a more sophisticated bench-scale study. A 5% dilution of retort effluent with an approximate COD of 1,050 mg/L was used to simulate combined blue crab processing plant effluent. Aeration studies of 6, 12, 18 and 24 hours were carried out in 18 liter reactors at approximately 20°C. The retort waters were added to 6 liters of activated sludge with MLSS concentrations between 3,000 to 7,000 mg/L. The authors concluded that under these conditions, approximately 90% of the COD was removed in the 16 and 24 hour retention times. Also, settling of suspended solids was poor, especially at the lower retention times because of their small size and low density.

Aerobic and anaerobic treatment alternatives for shrimp processing wastewater were examined by Pohland and Hudson (1976). Aerobic treatment of presettled wastewater was effective with an optimum hydraulic retention time of 2 hours for a non-recycle, continuous-flow, stirred tank reactor (CSTR) and 1 hour for CSTRs with recycle. Anaerobic treatment studies indicated inhibitory influences were present, but the influences could not be determined.

Anaerobic reactors have been found to be very effective in the removal of organics from many different types of wastewater, but their successful operation is dependent upon the maintenance of a number of important operational parameters. Given these facts, an examination of the processes involved in anaerobic treatments and their necessary operating conditions is vital to the design and use of any treatment system based on anaerobic digestion.

Both the anaerobic and aerobic digestion of organics are accomplished through the use of bacteria which use the organic material as food for growth. As in all biological processes, these bacteria will seek to minimize energy expenditures while attempting to achieve maximum cell growth. In aerobic systems, the free dissolved oxygen is utilized as the terminal electron receptor in an efficient oxidizing process which results in approximately 70% of the consumed substrate (i.e. BOD or COD) being used for cell growth. In research involving the aerobic treatment of crab processing wastes, the technology was found to achieve BOD_5 removals insufficient to meet discharge limits (Wheaton et al., 1984). In anaerobic conditions, where no free dissolved oxygen is present, other electron receptors are utilized in a less efficient reducing process. Specifically sulfur, if present, and more usually carbon from the organic material, will become electron receptors and be reduced. In addition, some carbon from the organics will be oxidized to form carbon dioxide and volatile acids. The end-product (i.e. methane) of the process still has a large amount of energy (i.e. potential to accept electrons) which is not directed towards the growth of cells. One result of this inefficiency is that while the production of excess sludge, measured as volatile suspended solids, can be expected to average 0.5 kg VSS/kg COD in aerobic systems, anaerobic digestion processes would average only about 0.1 kg VSS/kg COD and would thus require a lesser amount of nutrients for cell growth (Eckenfelder et al., 1988).

Anaerobic digestion has been described as a three-stage process involving three distinct groups of bacteria (McCarty, 1981). The first stage, known as hydrolysis and fermentation, involves the conversion of complex organics into intermediate products such as fatty acids, alcohols, and carbon dioxide. In the next stage, bacteria utilize these intermediate products to form acetic acid, hydrogen, and carbon dioxide through the processes of acetogenesis and dehydrogenation. The actual stabilization of the waste is accomplished in a final stage known as methanogenesis. Here biogas gas, composed mainly of methane and carbon dioxide, is formed as a result of acetate decarboxylation and carbon dioxide reduction (McCarty, 1964 and 1981). The methanogens responsible for the processes in this final stage are the key to successful digestion of organics, but they are sensitive to a variety of environmental factors which should be studied to ensure stable operation of the digestion process.

Of particular importance to the growth of methanogens is the maintenance of a stable pH level. Various studies have sought to determine the optimum pH level for the operation of anaerobic digestion processes. While these studies have all shown an optimum pH of about 7, the range over which the optimization of growth is said to extend has varied greatly. Clark and Speece (1970) found no inhibition over the pH range of 6 to 8, while McCarty (1964) stated an optimum range of 6.7 to 7.3, with methanogens functioning well in a pH range of between 6.6 and 7.6. If the pH falls below this range, the bacteria will not be able to complete the conversion of the intermediate acids to methane and carbon dioxide. This incomplete conversion will result in a buildup of these intermediates to a point where they will, in turn, begin to lower the pH even further and aggravate the problem. While some research has found methanogenic activity occurring at pH levels as low as 5 (Clark and Speece, 1970), it has also been suggested that acidic conditions resulting from pH levels of less than 6.2 are toxic to most methanogens and should thus be strongly avoided (McCarty, 1964). The presence of moderate amounts of sulfates lead to the recommendation that pH levels be maintained between 7 and 8 (Eckenfelder et al., 1988). In order to guard against such upsets, *Standard Methods for the Examination of Water and Wastewater* (APHA, AWWA, WPCF, 1992, hereinafter referred to as *Standard Methods*) has recommended that alkalinity in anaerobic reactors be maintained at a concentration of at least 2000 mg/L, while Eckenfelder et al. (1988) suggest that the alkalinity levels be maintained in the 2500 to 5000 mg/L range for successful digestion.

An additional factor that is important to the success of anaerobic digestion is the operating temperature of the reactor. Methanogens can be separated into mesophiles and thermophiles when discussing the effects of temperature on reactivity. Thermophilic processes primarily exhibit optimum performance in the range of temperatures from 45°C to 65°C. While thermophilic digestion has been found to occur at two to four times the rate of mesophilic processes, the additional heat input required to maintain the higher operational temperatures needed by these methanogens may offset any such gains (Zinder, 1988). Mesophilic reactions have been found to be most productive at temperatures ranging from 30°C to 40°C (McCarty, 1964). The upper limit of this range is very important because it has been observed that certain mesophilic strains do not survive in temperatures just above 40°C.

Another parameter important to the stable operation of anaerobic digestion processes is cation concentration in the reactor. Certain cations have been shown to be toxic to methane producing bacteria at high concentrations, while exhibiting a stimulatory effect when present at lower levels (McCarty, 1964). These effects are summarized in Table 6, as reported by McCarty.

It has also been suggested that the ability of anaerobic processes to withstand cation toxicity is related to the pH level. Specifically, the maintenance of a neutral pH was found to increase reactor resistance to this toxicity, when compared to that shown by a reactor operated at a pH of 8 (McCarty and McKinney, 1961).

Table 6. Stimulatory and inhibitory concentrations of alkali and alkaline-earth cations (McCarty, 1964).

Cation	Stimulatory Concentration (mg/L)	Moderately Inhibitory Concentration (mg/L)	Strongly Inhibitory Concentration (mg/L)
Sodium	100 - 200	3500 - 5500	8000
Potassium	200 - 400	2500 - 4500	12000
Calcium	100 - 200	2500 - 4500	8000
Magnesium	75 - 150	1000 - 1500	3000

In addition, the toxicity of high concentrations of some cations can be alleviated or even negated by the presence of low concentrations (generally in the range of 0.01 M to 0.05 M for monovalent cations and 0.005 M to 0.05 M for divalent cations) of one or more additional cations acting as antagonists (Kugelman and Chin, 1971). Selected data from studies performed by Kugelman and McCarty (1965) on the effect of cations on methanogenesis are presented in Table 7. The antagonistic effect is aided substantially by the presence of multiple cations at these low concentrations, to the point where it is possible for metabolism to gain a stimulatory effect from the cation presence and progress at a rate that is higher than if no cations were present. This multiple antagonism has been observed even in the presence of very high concentrations of the toxic cation. Conversely, studies have also shown that in the presence of toxic levels of one cation, slightly higher concentrations of the same antagonistic cations have resulted in synergism, exacerbating the toxic effect (Kugelman and McCarty, 1965).

Ammonia is produced, in the form of ammonium ions, as an end product of the anaerobic deaminization of proteins. While its presence can aid in the buffering of the digestion process against decreases in pH, it can, in higher concentrations, have a toxic effect upon the process. This toxic effect can take place as the result of either of two sets of circumstances.

First, the presence of ammonium ions in concentrations greater than 3000 mg/L have been found to be toxic to anaerobic digestion processes (McCarty, 1964). In addition, McCarty and McKinney (1961) found that free ammonia gas, when present in concentrations greater than about 150 mg/L, exhibits inhibitory effects on the anaerobic digestion process. This translates to inhibition when overall ammonia levels are between 1500 and 3000 mg/L and pH is greater than 7.4 to 7.6 (McCarty, 1964).

In order for bacteria to be able to grow, certain nutrients must be made available for their utilization. Primary among these nutrients is nitrogen, which is used in new cell construction, and phosphorus. McCarty (1964) states that the nitrogen requirement, based on an average chemical formula for biological cells, in anaerobic systems is approximately 11% of the weight of volatile solids produced. Phosphorus requirements for the digestion process

Table 7. Antagonistic relationships in dual cation systems.

Toxic Cation	Conc. M	% Control Reaction Rate No Antag.	Antag. Cation	Range of Peak Antag. M	% Control Reaction Rate Peak Antag.
Na	0.3	54	K	0.002-0.06	72
K	0.25	20	Na	0.01-0.05	75
K	0.25	20	Ca	0.005-0.05	59
K	0.25	20	Mg	0.005-0.05	64
NH_4	0.15	80	Na	0.002-0.05	105
Mg	0.1	34	Na	0.005-0.035	103
Ca	0.1	55	K	0.002-0.06	73

are listed as being between 15% (Speece, 1983) and 20% (McCarty, 1964) by weight of the nitrogen levels.

In addition to these primary nutrients, sulfide, nickel, iron, and cobalt are also seen as necessary, in small amounts, to the successful anaerobic digestion of organics. The need for sulfide can be attributed to the relatively high sulfur content (2.6%) found in methanogens, and its presence at concentrations ranging from 1 to 25 mg/L has been found to be stimulatory. The nickel requirement of methanogens appears to be related to the presence of a coenzyme, apparently unique to methanogens, designated as F_{430}. Finally, research has shown that additions of iron and cobalt have also improved the treatment efficiency of anaerobic systems (Speece, 1983).

Solids retention time (SRT) is the average retention time of microorganisms in an anaerobic system and as such is an important parameter in the operation of an anaerobic reactor. In order to have a sustained anaerobic digestion reaction, the rate at which active solids are wasted (1/SRT) must not exceed the rate at which cell growth occurs. While a minimum SRT value of 4.2 days has been used successfully for the anaerobic digestion of acetic, propionic, and butyric acids at a temperature of 35°C (Lawrence and McCarty, 1969), a value of 20 days is suggested as a minimum for anaerobic system design purposes (Eckenfelder et al., 1988).

In a conventional continuously stirred tank reactor (CSTR) without recycle the SRT is equal to the hydraulic retention time (HRT) of the reactor, but in the anaerobic contact process, the SRT can be varied separately from the HRT through the settling and recycling of solids. This SRT is calculated as the total weight of active solids in the system divided by the total weight of active solids wasted per day (Lawrence and McCarty, 1969).

The production of methane (CH_4) is a direct result of the anaerobic digestion process and can therefore be used to track the performance of the reaction. In addition, the methane produced by the reaction may be able to be used to help meet any heating energy requirements of an anaerobic reactor.

McCarty (1964) stated that theoretically, the complete digestion of one pound of COD should yield 5.62 ft^3 of methane (in metric units: 1 kg of COD = 0.348 m^3 of CH_4) at standard temperature and pressure (STP). This value is based on the following formula:

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \qquad (1)$$

where 1 mole of CH_4 (16 grams) has an oxygen (O_2) demand of 2 moles (64 grams). In order to convert this value for reactor operation at 35°C, Charles' Law must be employed:

$$V_2 = V_1 * (T_2 / T_1) \qquad (2)$$

where,

V_2 = gas production volume at 35°C in liters
V_1 = gas production volume at STP in liters
T_2 = reactor temperature in Kelvin
T_1 = STP temperature in Kelvin (273 K).

The use of this equation yields a relationship of 0.393 L of CH_4 produced for every gram of COD digested at 35°C.

Anaerobic reactors are divided into two main groups, conventional (low-rate) and modern (high-rate) systems. Each of these groups has its own benefits and drawbacks, and consists of a variety of individual reactor types.

The most basic of the low-rate reactors is the anaerobic lagoon. In this format, wastewater is dumped into a large stagnant pond and is allowed to undergo natural fermentation without mixing or temperature control. While energy and knowledge requirements are minimal, reactor volume is very large.

Another low-rate system which has seen a great deal of use in the treatment of wastewater is the CSTR. This system utilizes a sealed tank through which the waste is passed, while the reactor contents are stirred and temperature is maintained at a level appropriate for the growth of the sludge. While performance stability is improved and expertise and energy requirements remain low, a large reactor volume is still required, and solids washout can be a problem.

The anaerobic contact process system incorporates a settling chamber, either external or internal to the reactor, in order to increase solids retention times and prevent the washout problem. In addition, some contact process units employ a degasifier to further aid in the settling of the sludge.

High-rate systems of various types have recently found increasing application to the treatment of industrial and domestic wastes. These systems employ an immobilized biomass in order to achieve higher loading rates without the problem of solids washout. This immobilization is usually accomplished through the use of either a granular sludge, as in an upflow anaerobic sludge blanket (UASB), an attached biomass, as in the trickling filter and fixed film reactors, or the entrapment of the biomass, as in the anaerobic filter process.

The main advantage to these systems is in the reduced reactor volumes resulting from their higher loading rates. Drawbacks associated with these reactors can include higher energy, monitoring, and control costs, lower process stability in the presence of inhibitory materials, and an increased need for specialized knowledge (Lettinga, 1984).

2.5.3 Ammonia Removal by Air Stripping

There has been a great deal of discussion regarding the toxic effects of ammonia on aquatic life. Simultaneously, concern about the impact of ammonia concentrations in wastewaters discharged to the environment has led to the search for possible wastewater treatment methods to minimize the effects of such discharges.

As stated previously, ammonia occurs in two forms in nature, and while its presence as the ammonium ion (NH_4^+) can add to the total oxygen demand on the water, it is the presence of free ammonia (NH_3) that is the main toxicity problem. Studies have reported the onset of acute toxicity effects in aquatic organisms at levels ranging from 0.01 mg/L to over 2.0 mg/L of molecular ammonia-nitrogen (Culp, 1978). The two forms of ammonia exist in an equilibrium reaction which is affected primarily by temperature and pH levels. As temperature or pH decreases, the amount of ammonia present in its molecular form (NH_3) decreases until at pH 7 and normal air temperatures, until almost all of the ammonia is present as NH_4^+ (Culp and Culp, 1971).

Studies involving the process of air stripping have shown it to be an effective, low cost tool in the removal of ammonia from both domestic and industrial wastewaters with observed removal rates ranging from 85% to over 95%. These same studies have also shown that the process is sensitive to changes in various operating parameters which may result in decreased treatment. Specific data from these studies will be presented in the following subsections.

Primary among the parameters affecting operating performance of air stripping towers in the removal of ammonia is wastewater pH. It has been shown that ammonia removal increases steadily as the pH of the wastewater influent is raised from 8 to 11, although no significant increases in treatment were seen as the pH was then raised from 11 to 12 (Kuhn, 1956). This study was performed with municipal wastewater, and it utilized a 7 ft tower packed with 0.5 in Raschig rings. Air-to-water ratios in the experiment varied between 40 and 500 ft^3/gal. Another study has shown effective ammonia removal (92% to 98%) can be achieved at pH levels above 9.7 (Slechta and Culp, 1967; as cited by Culp and Culp, 1971). This research treated a chemically coagulated and filtered secondary effluent, with ammonia levels of 25 - 35 mg/L as N, to test ammonia stripping technology. The tower used in these trials was 5 ft in diameter with a 2 ft packing depth. Redwood slats were employed as the tower packing. Air-to-water ratios between 380 and 3040 ft^3/gal were tested with optimal treatment achieved at approximately 750 ft^3/gal. Adjustments of pH necessary for studies of this process have been accomplished through the use of either sodium hydroxide (NaOH) or lime ($Ca(OH)_2$).

While the use of lime has often been found to incur lower chemical costs, such use has sometimes resulted in scaling problems. It has thus been suggested that provisions be made to allow for the cleaning of any such accumulations within the tower.

Other parameters which have been shown to affect process efficiency are the ambient air temperature at which the stripping tower is operated and the temperature of the wastewater to be treated. One problem found in areas where air temperatures dip below 0°C is an accumulation of ice which makes tower operations impractical. Additionally, since the solubility of ammonia in water increases as temperature decreases, any drop in temperature can cause a decrease in the effectiveness of ammonia removal by air stripping (Culp, 1978).

This second effect is most readily shown through the temperature dependence of the Henry's Constant for ammonia which is calculated by the following formula (Culp <u>et al.</u>, 1986):

$$\log H = -\Delta H^\circ/RT + C \qquad (3)$$

where:

H = Henry's Constant, $m^3 \cdot atm/mol$
ΔH° = Change in Enthalpy due to dissolution of component in water, 3750 kcal/kmol for ammonia
R = Universal Gas Constant, 1.987 kcal/kmol•°K

T = Absolute Temperature, °K
C = Constant specific to contaminant, 6.31 for ammonia.

Use of this formula gives a Henry's constant for ammonia which ranges from 0.57 atm at a liquid temperature of 15°C to a value of 5.33 atm at a wastewater temperature of 65°C. Since a higher Henry's constant means that the contaminant in question has a higher equilibrium concentration in air, it can be seen that increased wastewater temperatures will allow the ammonia to be more easily stripped.

Due to the high volumes of air, relative to the amount of liquid treated, necessary for the stripping of ammonia, the temperature of the wastewater leaving an air stripping tower is a function of both the influent temperature and, more importantly, the air temperature. In support of this, studies have shown substantial decreases in removal efficiency as air, and subsequently effluent, temperature is lowered from 22°C to 5°C (O'Farrell et al., 1973). In a similar fashion, extreme changes in influent temperatures can also be seen to affect removal efficiencies possible through air stripping (Joyce, 1993).

An additional factor shown to have a substantial effect on the removal rates possible through the use of this technology is the air to water ratio utilized in the operation of the stripping tower. Increased air to water ratios have been shown to increase the levels of removal possible through this technology and are seen as a possible remedy to efficiency problems brought about by decreased air temperatures. In order to achieve optimum removal it has been suggested that air to water ratios of 300 to 500 ft^3/gal be employed during warm weather (> 20°C) tower operation, while up to 800 ft^3/gal may be needed under cold weather (< 10°C) conditions (Culp, 1978).

Together with the utilization of sufficient air to water ratios, it has been observed that the maintenance of hydraulic loading rates at levels which will allow droplet formation and prevent water sheeting is important to proper tower operation (Culp, 1978). While the exact loading rate can be seen as a function of both depth and type of packing, research has shown that, in 20 and 24 ft cross- current towers using redwood slats as packing, sheeting can be eliminated, and ammonia removal maximized, at liquid loading rates of less than 3.0 gpm/ft^2 (South Tahoe Public District, 1971).

Tower design has also been cited as an important factor in the level of ammonia removal accomplished in air stripping studies. The two major types of air stripping towers employed in the aforementioned studies are countercurrent and crosscurrent. In countercurrent towers the entire airflow enters the tower at the bottom and flows up through the falling wastewater, while the crosscurrent design pulls air in through the sides of the tower over the entire height of the packing. It has been suggested that countercurrent towers are more effective in the removal of ammonia and less resistant to scaling problems than those of crosscurrent design (Culp, 1978).

The depth and type of packing material are other aspects of tower design that have been shown to have great effects on the removal efficiencies possible with this technology. It has been demonstrated that an increase in the packing depth of treatment towers has led to increases in observed ammonia removal rates achieved by these towers with maximum removals achieved at a depth of 24 ft (Slechta and Culp, 1967).

The calculation of packing depth is achieved through the use of the following set of equations (Gostin, 1993). Due to the proprietary nature of certain information, certain constants are presented without explanation.

$$\text{Packing Height} = (3 * \text{HTU}) * \text{NTU} * 1.1 \tag{4}$$

$$\text{HTU} = A * (L/M_L)^{0.33} * (M_L/\rho_L D_L)^{0.5} * (T/286)^{-4.255} \tag{5}$$

Where:

HTU = Height of Transfer Unit in ft
A = Constant, 0.0042 for Ammonia
L = Liquid Loading Rate in lbs/hr/ft^2
M_L = Viscosity of Water in lbs/hr-ft

$$M_L = 4.3231 * (T/273)^{-7} \tag{6}$$

ρ_L = Density of Water, 62.4 lbs/ft^3
T = Temperature of Water in °K
D_L = Diffusivity of Ammonia in Water in ft^2/hr

$$D_L = 6.3635 * 10^{-4} * (T/273)^{8} * (1/V_c^{0.6288}) \tag{7}$$

V_c = Critical Molar Volume of Ammonia, 72.5 cm^3/g-mole

$$NTU = \ln \frac{(C_1 - (1331.2/(H_c * AW)) * (C_1 - C_2))/C_2}{(1 - (1331.2/(H_c * AW))} \tag{8}$$

Where:

NTU = Number of Transfer Units
C_1 = Influent Ammonia Concentration in mg/L
C_2 = Effluent Ammonia Concentration desired in mg/L
AW = Air to Water Ratio in cfm/cfm.

Finally, since the success of this process is dependent upon the repeated formation and rupture of water droplets, it is important that the packing material be able to accomplish these functions while remaining resistant to the effects of pH and scaling (Culp, 1978).

3.0 METHODS AND MATERIALS

Methods and materials used to characterize process wastewater from three blue crab processing plants in Virginia are presented in this chapter. Also presented are the methods and materials used to perform treatability studies and corresponding analytical procedures.

3.1 Plant Process Characterization

The wastewater streams originating from the various operations in three blue crab processing plants were characterized. The three plants studied are all located in Virginia. For the purposes of this paper, the processing facilities will be designated plant #1, #2, and #3, respectively. Two trips each were made to plants #1 and #3. Three trips were made to plant #2.

In plant #1, crabs are processed by hand picking and the Harris Claw machine. Plant #2 uses both of these operations but also uses the Quik-Pik machine. Plant #3 uses only hand picking. In the plants #1 and #2, the following processes were characterized: cooking, hand picking, Quik-Pik process (plant #2 only), Harris Claw process, and clean-up. In the third plant, only the retort effluent was characterized.

Effluent samples were analyzed for the following: pH, temperature, chemical oxygen demand (COD), 5-day biochemical oxygen demand (BOD_5), total suspended solids (TSS), volatile suspended solids (VSS), total Kjeldahl nitrogen (TKN-N), ammonia nitrogen (NH_3-N), total phosphorus (TP), oil and grease (O&G), and chloride (Cl). The criteria used in selecting these tests was to include those that are now regulated, those that may be regulated in the near future, and those that might impact treatment works or the environment.

Samples of each effluent were collected in the following manner. First, a grab sample of a given effluent was collected in a clean, five gallon bucket when a process was operating at full speed. The contents of the bucket were mixed thoroughly and then transferred into 500 ml polyethylene bottles. At this time, the temperature and pH of the sample were taken. Half of the samples were then acidified to a pH of 2.0 or below for preservation. All samples were immediately placed on ice in a barrier container. Two exceptions to the method of collection were the retort water from the cookers and the brine bath from the Harris Claw process. For the retort water, all the fluids from a cooking cycle were collected in a 55 gallon receptacle. The contents of the drum were mixed thoroughly with a paddle to homogenize the sample. Samples were then transferred to polyethylene bottles by a hand pump. The entire effluent stream from the cook was collected after it was determined that the characteristics of the retort water changed substantially during the cooking cycle and a single grab sample would not be representative.

The samples from the brine bath were collected at the end of the Harris Claw process cycle just before the contents of the brine tank were dumped. The brine solution was recirculated for an entire processing period and therefore was most concentrated at the end of the cycle.

After samples were collected, they were packed on ice and transported by car to VPI & SU's Environmental Engineering Analysis Laboratory in Blacksburg, Virginia. The samples were then placed in a 4°C refrigerator until all analyses were performed. All tests on non-acidified samples were completed within 48 hours of collection and all tests on acidified samples were completed within 28 days as prescribed by EPA Methods for Chemical Analysis for Water and Wastes (EPA, 1983).

3.2 Treatability Studies

Three treatability studies were performed on each individual effluent stream from the various processes in the processing facilities including settling, filtration, and coagulation with pH adjustment. Anaerobic biological treatment studies were performed on the cooker effluent and a theoretical "combined" plant effluent. Also, anaerobic biological treatment of these effluents after pretreatment with coagulation by pH adjustment was examined.

3.2.1 Settling and Filtration Tests

Settling and filtration tests on effluent samples were performed. Values for COD, BOD_5, TSS, VSS, TKN-N, NH_3-N, and total phosphorus were determined for each effluent type before settling, after settling and after filtration. To obtain settled values, collected samples were mixed and then allowed to settle at room temperature (approximately 23°C) for one hour. The supernatant from a sample was then analyzed for the above constituents. To obtained filtered values, samples were filtered through Whatman grade 934AH glass-fiber filters. Analyses were performed on the filtrate.

3.2.2 Determination of Process Flows and Volumes

In the process characterizations, effluent flow rates and volumes were determined by the following methods. First, the water line which fed the Harris Claw process at plant #1 had a flow totalization meter. Omega FTB series turbine flow totalization meters were placed on several individual feed lines of the Harris Claw and Quik-Pik process at plant # 2. The effluent volumes produced over a given period of time could be read directly from the meters.

Some effluent streams, such as the shell liquid effluent, could not be read directly from totalization meters. In this case, the time required for the effluent to fill a calibrated five gallon bucket was recorded. When the bucket method was used, several samples were collected during process operations and the results averaged to improve the accuracy of the measurement.

The volume of the Harris Claw brine tanks was determined by measuring their dimensions. This volume was multiplied by the number of times the brine tanks were dumped and refilled during the day. Some solution was removed from the brine tanks by the meat and the shell fragment conveyor belts. This volume was accounted for in the measurements of those respective effluent streams.

Finally, the volume of the retort water was determined in two separate ways. At plants #1 and #2, cooker water flowed through pipes which discharged at the edge of the docks. The cooker water was collected by placing a five gallon bucket over the end of the pipe, and then repeatedly transferring the contents of the bucket to a large calibrated receptacle, until the cooking cycle was complete. At plant #3, the pipe exiting the cooker was attached to a fire hose. The fire hose was connected directly to a 55 gallon industrial drum where all the fluids from a cook were collected. A dip stick was then placed in the drum to determine the depth occupied by the cooker effluent. The total volume was then calculated from the measured dimensions.

3.2.3 Coagulation with pH Adjustment

Coagulation with pH adjustment was carried out to investigate the removal of COD, BOD_5, TSS, VSS, TP, TKN-N, and, NH_3-N. The volume and percent moisture of the resulting sludge was also determined. The authors noted that when the pH of the effluent samples was decreased to pH 6.0 or below, a coagulation of the contents occurred.

To test treatability by acidification, jar tests were completed on highly concentrated processing effluents from plant #2 including the retort water, the Harris Claw brine bath and the Harris Claw reel wash water.

In the jar tests, one liter of the given effluent (19-22°C) was placed in each of six, one liter, circular jars. It should be noted that the samples had been previously allowed to settled at room temperature for one hour so as to remove all solids susceptible to gravity settling. In this way, the removal efficiency resulting from coagulation of colloidal particles due to pH adjustment could be determined.

In the first five jars, the pH of the samples were adjusted to 6.0, 5.0, 4.0, 3.0, and 2.0, respectively, with concentrated sulfuric acid. The final jar acted as a control and the pH was not adjusted. The six jars were then placed in a standard gang mixing jar test apparatus and flocculated with single-blade paddles at 20 rpm for 15 minutes. The mixing was then stopped and the solids allowed to settle for 30 minutes (60 minutes for the retort water). After settling, the supernatant was carefully separated from the sludge layer with a 100 ml syringe and transferred to 500 ml sample bottles to be analyzed. The volume and percent moisture of the resulting sludge was then determined.

The pH adjustment study described above was carried out on a second set of samples from plant #2 one month later. The procedure was identical with the following exceptions. Only three pH ranges (other than the control) were examined. The pH of the retort water was adjusted to 4.0, 3.0, and 2.0. The pH of the Harris claw brine bath and the Harris claw reel wash water were adjusted to 5.0, 4.0, and 3.0.

In a separate study, a five gallon volume of cooker water with an initial pH of 6.44 and a temperature of 21°C was acidified to pH 3.5 with concentrated sulfuric acid. The solution was stirred slowly by hand with a metal paddle for 15 minutes to induce flocculation and then allowed to settle for one hour. The supernatant was carefully decanted from the corresponding sludge with a beaker.

3.2.4 Sludge Analysis by Food Science and Technology

A two liter volume each of the initial retort water, settled retort water, acidified retort water and the resulting sludge were sent to VPI&SU's Food Science and Technology Analysis Laboratory in Blacksburg, Virginia, for analysis. The samples were analyzed for the following parameters: protein, carbohydrate (acid digestible fiber), fat, ash and percent water. All tests were performed in accordance with *Official Methods of Analysis by the Association of Official Analytical Chemists, Inc.* (15th edition, 1990) with the following exceptions and notations:

- Protein was determined by potentiometric endpoint determination (Kjeldahl) using 0.25 gm dry sample digested with 6 gm catalyst and 12 ml H_2SO_4 for 30 minutes. The catalyst was comprised of 9.9 gm of K_2SO_4, 80 mg selenium, and 410 mg $CuSO_4$. Then, 75 ml distilled water and 50 ml 38% NaOH (w/w) was added. The sample was titrated with 0.1 N HCl using bromcresol green as the indicator with green as the endpoint color.
- Fat was determined by the Goldfisch Soxhlet Method (Randall, 1974). One gram of dry sample was extracted for 4.0 hours in boiling petroleum ether.
- Ash was determined by burning sample at 550°C for 6 hours and then checking every hour until no change in weight was detected.
- Percent water was determined by freeze drying the sample for 48 hours at -60 to -80°C.

3.2.5 Anaerobic Biological Treatment

3.2.5.1 Short-Term Anaerobic Tests

Anaerobic biological treatment of retort water and a theoretical "combined" plant effluent both before and after pH adjustment effluent were examined. A given volume of each of these effluents was fed to a two-liter, completely-mixed, reactor every 24 hours. All reactors and gas collection equipment were maintained at 35°C in an incubator. A schematic of the reactor design is shown in Figure 2.

The reactors, designated A, B, C, and D, treated the following process wastewaters: A - retort water, B - retort water after coagulation by pH adjustment, C - combined effluent, and D - combined effluent after coagulation by pH adjustment. Reactors B and D were

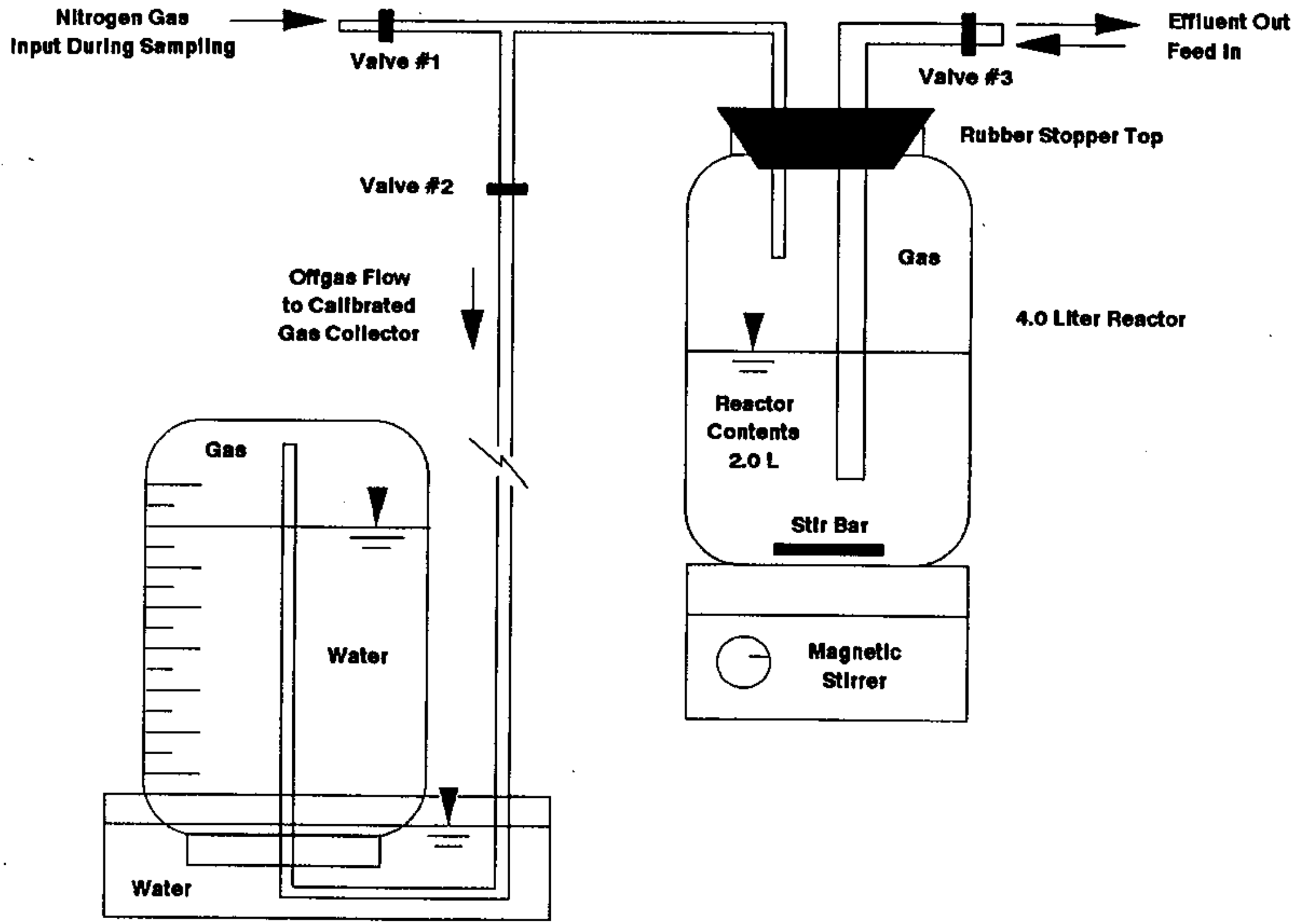


Figure 2. Experimental arrangement used in the anaerobic reactor studies.

operated to compare the relative degradability of the wastewaters after coagulation by pH adjustment.

The combined plant effluent contained retort water, Harris claw reel wash water, Harris claw brine bath and Harris claw meat conveyor wash water in a 1:1:1:3 volumetric ratio. The combined effluent was formulated to represent the typical daily process wastewater discharged from plant #2, as determined in the preceding characterization study.

To obtain feed B, the initial pH of a five gallon portion of feed A was decreased from 8.1 to 4.0 with concentrated sulfuric acid. The contents were stirred slowly for fifteen minutes with a metal paddle to induce flocculation. The solids were then allowed to settle for one hour. The supernatant was decanted and the pH was adjusted to 7.0 with 6N sodium hydroxide. Feed D was made from a five gallon volume of feed C in an identical manner. All feed was kept at 4°C until use. At that time, the necessary volume of feed was warmed to 35°C before being transferred to a reactor. The total COD, TSS, and VSS of the feeds were monitored to track degradation during storage.

The four reactors were operated at approximately 4,000 mg/L of mixed liquor volatile suspended solids (MLVSS). To achieve this MLVSS concentration, the reactors were seeded with 0.4 liters of anaerobically stabilized municipal sludge from Peppers Ferry Sewage Authority. A volume of 1.6 liters of warm tap water was added to this to bring the final volume to 2.0 liters. The tap water was purged with pure nitrogen gas to remove dissolved oxygen prior to mixing with the sludge. To buffer against upsets during start up, 4.0 grams of calcium carbonate was added to each reactor.

The appropriate volume to feed and waste from each reactor every 24 hours was based on food to microorganism ratio (F/M). Units for F/M values were lb COD/lb MLVSS/day. The feed volume was calculated as follows:

$$\text{feed volume/day} = \frac{(\text{F/M})(\text{reactor volume})(\text{MLVSS})}{(\text{total COD of feed})} \qquad (9)$$

where:
feed volume = ml
F/M = 1/day
reactor volume = ml
MLVSS = mg/L
total COD = mg/L.

For the first ten days, systems A and B operated at a F/M of 0.3 and systems C and D operated at an F/M of 0.2. During this time, indications of reactor failure appeared. The pH and total alkalinity of each system dropped and the soluble COD steadily increased in the effluent. The decision was made to operate reactors A and B at a F/M of 0.15 and reactors C and D at a F/M of 0.05. After the loadings were reduced, conditions in the systems stabilized.

In order to maintain a constant 2.0 liter reactor volume without losing biomass during wasting, stirrers were turned off for 30 minutes and the biomass was allowed to settle. The appropriate volume of effluent was wasted from the reactors each day through a glass outlet tube. To accomplish effluent removal, valve #2 was closed as shown in Figure 2. Valve #1 was then opened and pressurized nitrogen gas was forced through the inlet tube in the top of each reactor. Value #3 on the outlet tube was then opened. The pressure produced in the reactors forced the effluent through the outlet tube and into a collection flask. After the proper volume of effluent was wasted, the nitrogen gas was cut off and valves #1 and #3 were closed while valve #2 was reopened. This procedure prevented free oxygen from entering the system during wasting and feeding.

The systems were operated for a total of 55 days. Reactors A and B were operated at a F/M of 0.15 from day 11 to day 31. The soluble COD in the effluents of these reactors decreased during this period. As a result, the F/M was increased to 0.25 for day 32 to day 55. Reactors C and D were operated at a F/M of 0.05 from day 11 to day 55.

The mean cell residence times (MCRT) for reactors A and B were 153 days and 149 days, respectively, before day 31, but changed to 136 days and 132 days, after day 31. The MCRTs for reactors C and D were 109 days and 101 days, respectively. The hydraulic retention time (HRT) for all reactors was 18.2 days except for reactors A and B after day 31 where it decreased to 12.5 days.

The temperature and total gas volume produced by each reactor were monitored daily. Every two days, pH, MLSS, MLVSS, and alkalinity were measured in each reactor. The following tests were done on the feeds and the effluents from each system at intervals necessary to track performance: TSS, VSS, total BOD_5, total COD, soluble COD, TKN-N, NH_3-N, TP, calcium, magnesium, potassium, and sodium. Also monitored were the following four volatile acids: acetic acid, propionic acid, isobutyric acid and n-butyric acid.

3.2.5.2 Long Term Anaerobic Tests

Anaerobic treatability studies were performed to determine the effectiveness of this method in achieving a reduction in the organic matter of the process waters.

The reactor vessel was made from a 4 l high density polyethylene (HDPE) bottle. The reactor top was plugged with a size 13 rubber stopper, through which two 1/2 in glass tubes were fitted. This allowed feeding, capture of biogas, and sampling of effluent, while maintaining an anaerobic environment in the reactor. The effluent/feeding tube was connected to a 1 ft length of 1/2 in Tygon® tubing which was clamped off at all times except during feeding and sampling. The biogas was passed though Tygon® tubing to the collection system. This consisted of a calibrated 4 L HDPE bottle which was filled with water and inverted in a water bath. The biogas tubing was passed into this bottle, and its volume was measured via water displacement. Figure 3 shows a schematic of the reactor setup.

All wastewaters treated in this experiment were taken from the Graham and Rollins processing facility located in Hampton, VA. Retort wastewater was obtained by collecting the process flow in a 5 gal HDPE bucket held under the retort waste outlet. The wastewaters from the Harris Claw reel wash, brine bath, and conveyor wash operations were collected in similar manners. Once obtained, the feeds were stored at 4°C in HDPE containers until used. Total COD of the feed was monitored during storage to check for degradation.

Four reactors were operated over the course of the study. Reactors A and B treated retort water and were maintained at food to microorganism (F/M) ratios, stated in units of lb COD/lb MLVSS/day, of 0.35 and 0.25, respectively. Reactor A was operated at an F/M ratio of 0.40 from day 1 through day 40, but the ratio was decreased to 0.35 after reduced treatment was observed. Reactors C and D, meanwhile, were fed a mixed wastewater, the composition of which was based on daily production flows determined for the processing facility. These reactors were maintained at F/M ratios of 0.10 and 0.07, respectively. The wastewaters from the Harris Claw reel wash, brine bath, and conveyor wash operations were combined with the retort process water in a 2.0:4.9:8.6:3.4 ratio to make this mixed wastewater. Reactor A had been operated at F/M ratios of 0.15 and 0.25 over a total of 55 days prior to the beginning of this study. Similarly, reactor C was operated at an F/M ratio

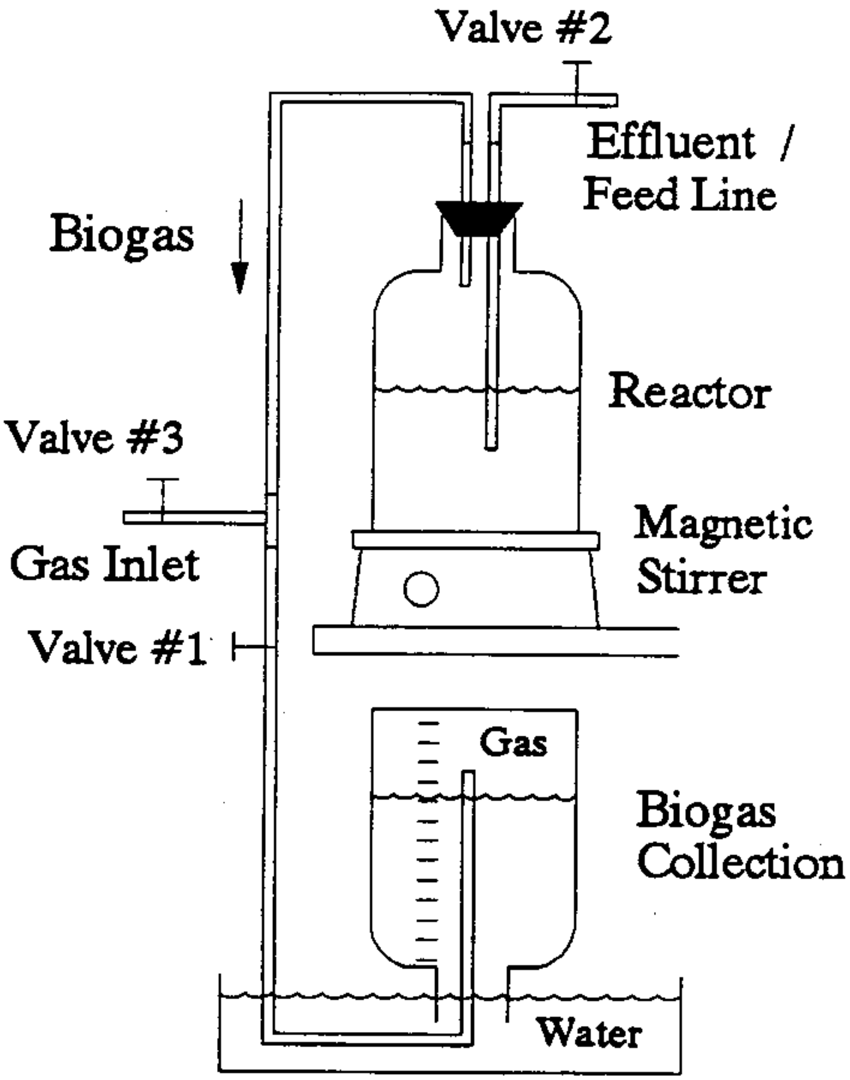


Figure 3. Anaerobic reactor schematic.

of 0.05 for a period of 55 days prior to the beginning of this study. For these reactors, day 1 is the first day on which the F/M ratio was increased to the levels tested in this study. In the case of reactors B and D, day 1 is the day on which the reactors were started.

The F/M ratios were used to determine the volumes that were wasted from and fed to the reactors on a daily basis. These volumes were calculated as follows:

$$\text{feed volume/day} = \frac{(\text{F/M})(\text{reactor volume})(\text{MLVSS})}{(\text{total COD of feed})} \tag{10}$$

where:
feed volume = mls
reactor volume = mls
MLVSS = mg/L
total COD = mg/L

In an attempt to achieve a start-up mixed liquor volatile suspended solids (MLVSS) level of between 3,000 and 4,000 mg/L, each reactor was seeded with 0.4 liters of anaerobically stabilized municipal sludge. This was obtained from the Peppers Ferry Sewage Authority treatment plant. This was added to 1.6 liters of warm tap water, which had been purged with pure nitrogen in order to remove any dissolved oxygen, for a total reactor volume of 2.0 liters. Calcium carbonate, in the amount of 4.0 grams, was added to each reactor to guard against upset during start-up.

In order to maintain a relatively constant temperature, the reactors were kept in an incubating cabinet at approximately 35°C. This temperature was chosen in order to maintain the reactors in the mesophilic range recommended for anaerobic digestion (McCarty, 1964). The reactors were maintained in a continuously-mixed state through the use of magnetic stirrers. These stirrers were turned off only for a period of 45 minutes to an hour prior to the daily wasting and feeding of the reactors to minimize the loss of MLVSS. This settling period was increased to 1 hour for reactors C and D after the MLVSS concentrations in each began to decrease.

After the settling period, valve 2, referring to the schematic in Figure 3, was closed and valves 1 and 3 were then opened. Pressurized nitrogen gas was then fed through the inlet tube and was used to force the appropriate amount of waste out of the reactor through the effluent tube where it was collected in a graduated cylinder. The nitrogen gas was then disconnected. The amount of feed needed to maintain the desired F/M ratio was then poured into the reactor through the effluent tube, and valves 1 and 3 were closed. Finally, the stirrer was turned on and valve 2 was opened.

Due to changes in the MLVSS concentrations of the reactors and in the strength of the wastewater used as feed, the hydraulic retention times (HRT) of the reactors varied considerably over the course of their operation. Average HRTs for reactors A, B, C, and D were found to be 12.2, 18.7, 36.3, and 32.4 days, respectively.

Reactor performance was tracked by monitoring temperature, biogas production volume, pH, alkalinity, and MLVSS. In addition, concentration levels of the following factors were determined for both the feeds and the reactor effluents at intervals necessary to define reactor health and treatment effectiveness: total and soluble BOD_5, total and soluble COD, total and volatile suspended solids (TSS and VSS, respectively), Total Kjeldahl and Ammonia Nitrogen (TKN-N and NH_3-N, respectively), Total Phosphorus (TP), and the cations, calcium, magnesium, potassium, and sodium.

3.2.6 Air Stripping Experiment

These experiments were carried out in order to assess the effectiveness of this technology in achieving the removal of the ammonia present in the retort process effluent. Figure 4 shows a schematic of the air stripping tower design.

The tower was constructed from three sections of 5/8 in thick PVC pipe with an inside diameter of 18 in. PVC piping was chosen for its heat and chemical resistance properties. The base of the tower was constructed from a 2.5 ft section of the pipe which was secured to a 20 in square Plexiglas base with a thickness of 5/8 in. The joint was sealed with a silicone-based caulk to prevent leakage. A shelf of 1 in chicken wire was inserted into a 6.5 ft pipe section and was held in place with a PVC support ring. By using this type of packing support, adequate strength was supplied, while back pressure was kept to a minimum. The 6.5 ft section was then inserted into the flanged top of the first section and filled to a 6 ft depth with 2.3 in LANPAC® packing material. This depth was chosen based on calculations using equations (3) through (8) and was based on an influent ammonia concentration of 160 mg/L and a desired 95% removal rate, together with a liquid flow rate of 1.2 GPM, an air flow rate of 500 cfm, and a minimum water temperature of 15°C. Calculations of this packing depth are presented in Appendix B. A 1 ft section of the pipe

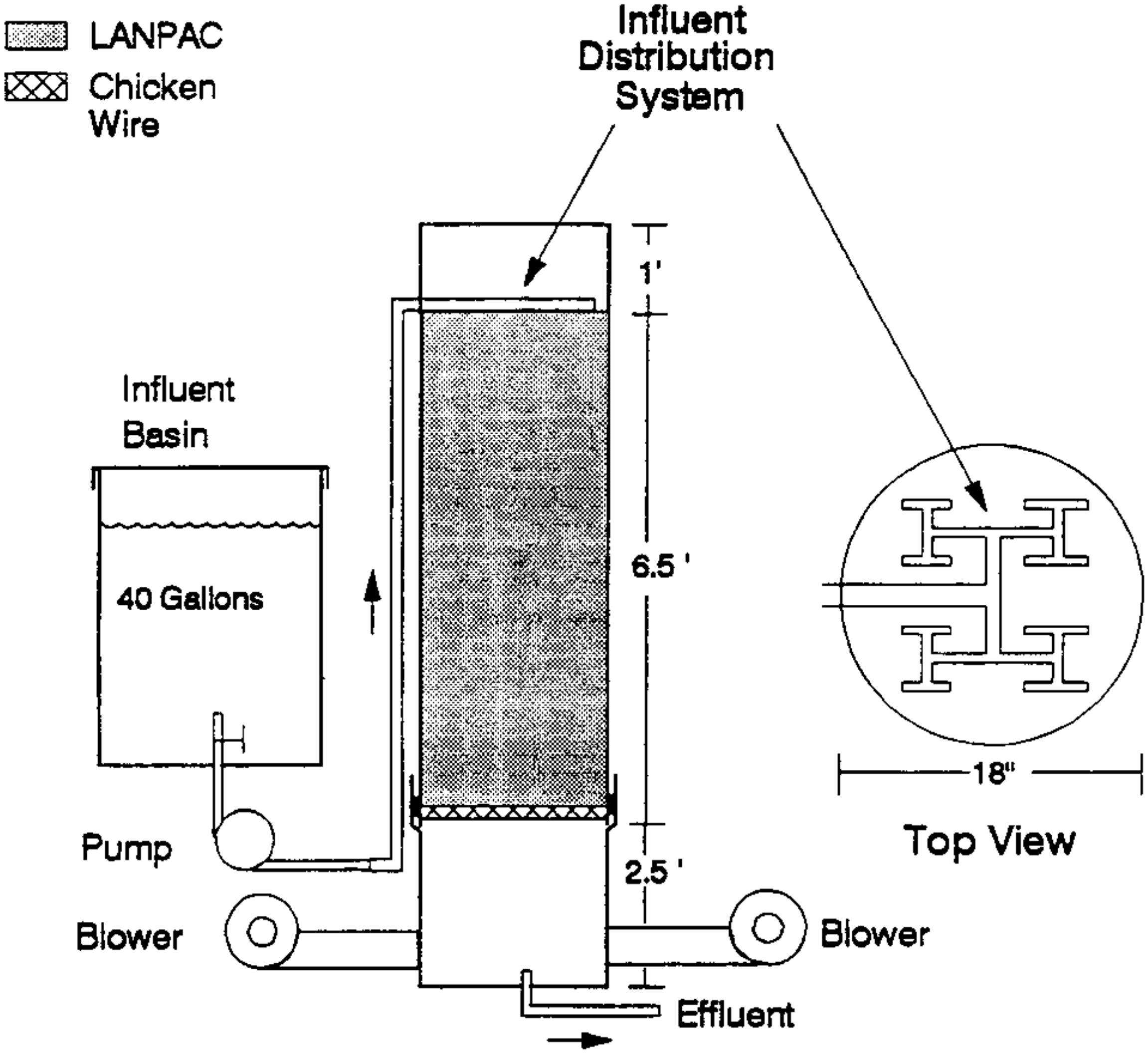


Figure 4. Air stripping tower schematic.

was then attached to the top of the tower in order to prevent splashing when the blowers were in operation.

The wastewater was pumped from a HDPE holding basin by a centrifugal pump resistant to temperatures up to 93°C and pH levels up to 11. In order to prevent damage from higher pH levels, the pump was flushed with tap water after use. This pump was powered by a 0.04 hp motor. A combination of high temperature 0.75 in diameter PVC piping and Tygon® tubing was used to transport the water from the basin to the high temperature PVC distribution system at the top of the 6.5 ft section.

The tower was of a countercurrent design with the air supplied by two two-speed blowers which were both operated at their high speed setting and generated a combined air flow-rate of approximately 500 cfm. These were attached to the base of the tower via 4.25 in by 8.5 in galvanized steel ducts inserted into ports in the side of the bottom section. The treated wastewater left the tower via two 0.75 in ports located in the side of the base section.

The wastewater used in the testing of the tower was collected in a 5 gal. HDPE bucket held under the retort process waste outlet. Although every attempt was made to recover the entire retort waste stream, a small percentage was lost. It is believed, however, that a representative sampling of the stream was recovered.

This water was then transferred to the holding basin where pH adjustments necessary for the different tests were done using 50% NaOH w/w (weight ratio). The pH of the waste was checked using a Fisher Model 160, portable pH meter with a glass electrode probe. After the pH of the waste in the basin was adjusted, the basin was covered in order to minimize ammonia loss before treatment.

After the waste was brought to the appropriate pH level, ranging from 10 to 12.2 depending on the trial, the pump was primed and the tower was allowed to operate until a consistent effluent flow-rate was observed (approx. 10 minutes). Flow-rate was determined by measuring the time required for the tower effluent to fill a calibrated bucket. The three flow-rates considered in the trials were 0.6, 1.2, and 2.2 gal/min.

Once a constant flow rate was observed, grab samples of both influent and effluent were taken at 5 minute intervals. Influent and effluent temperature readings were also taken at these intervals in order to determine any cooling benefits of the air stripping process. In order to track the tower performance over the course of each trial, individual effluent samples were taken within 30 seconds of the corresponding influent sample. In addition, samples of the retort effluent were taken prior to pH adjustment to ensure that the tower influent ammonia levels were consistent with the retort effluent levels. All samples were taken and stored in 500 ml HDPE bottles and were acidified to pH < 2 through the addition of 15% sulfuric fuming acid. The samples were transported to Virginia Tech Environmental Engineering Laboratory where ammonia levels were determined according to the *Standard Methods* (1992) procedure stated in the previous section.

The efficacy of the treatment scheme was determined through the comparison of the tower effluent and influent (retort wastewater after pH adjustment). To ensure that the main portion of the ammonia removal occurred as a result of the air stripping process and not the pH adjustment itself, occasional samples were also taken of the retort process waste prior to the pH adjustment and compared to the tower effluent.

In order to the examine the potential benefits of operating two towers in series, two trials were conducted. In the first, conducted at pH 12.2, the tower effluent from one test was collected, combined in a 3:1 ratio with a portion of that test's influent, and used as the influent for the tower. This combination was attempted in order to test the potential benefits of increasing the temperature of the effluent before recycling it through the tower. In the second, performed at pH 11.0, the effluent was collected and run back through the tower without any mixing. Samples from these tests were taken, preserved, and analyzed, as described above.

The trials were geared toward determining the effects that changes in three variables would have on the tower performance. These factors were influent temperature, pH, and the air-to-water ratio at which the tower was operated. This last parameter was adjusted by varying the input liquid flow rate. The values of the operating parameters for each test are summarized in Table 8. The trials are presented in the order in which they were performed.

In all tests the air flow rate was held steady at approximately 495 - 500 CFM. This flow-rate was calculated by multiplying the average air speed through the tower, measured across the tower diameter, by the area of the tower opening. The average air speed was measured with a TSI Model 1650, air velocity meter.

Table 8. Influent variables for air stripping tests.

TRIAL	pH	Liquid Flow Rate GPM	Air-To-Water Ratio ft^3/gal	Influent Temperature °C
1	11.0	2.2	225	56
2	11.0	1.2	412	58
3	10.0	1.2	412	59
4	11.1	1.2	412	65
5	12.2	0.6	825	58
6	12.2	0.6	825	20
7	12.2	0.6	825	50
8	11.0	0.6	825	58
9	11.0	0.6	825	20
10	10.0	0.6	825	58

3.3 Analytical Procedures

All samples were analyzed in accordance with *Standard Methods* with the following exceptions and notations:

- BOD_5 samples were not seeded with the exception of those used in the acidification treatability study. With the acidified samples, the pH was adjusted to approximately 7.0. Blanks and samples were then seeded with primary effluent from a municipal activated sludge facility. Quality control samples made from a standard glucose-glutamic acid solution were included.
- Chloride was determined by means of ion chromatography (Dionex model 2010i chromatograph) with a cross-linked polystyrene/ divinyl benzene column, flow rate of 2.0 ml/min and pressure of 1,000 psi.
- Total phosphorus was determined using the ascorbic acid method with persulfate digestion. Absorbance values were determined with a Beckman DU-6 spectrophotometer at a wavelength of 880 nm.
- Volatile acids were determined with a Tracor 560 gas chromatograph equipped with flame ionization detection and a polar Carbowax 20M column. Nitrogen was used as a carrier gas with a flow rate of 3.8 ml/min. The injection port and detector were set at 200°C and the oven temperature was 120°C. The run time was 6 minutes and was isothermal. All samples were acidified with concentrated formic acid at 0.01 ml/ml.
- Calcium, magnesium, potassium and sodium were determined with a Perkin-Elmer 703 Flame Atomic Absorption Spectrophotometer. Samples were acidified with concentrated nitric acid. Standard lanthanum chloride solution was added to calcium and magnesium samples and standard cesium chloride solution was added to potassium and sodium samples to minimize ionization of the tested metals in the flame.
- Alkalinity of anaerobic reactor contents was determined by titration of 25 ml samples with 0.10 N H_2SO_4 to a pH of 4.3.

4.0 RESULTS AND DISCUSSION

The results of wastewater characterizations for three blue crab processing facilities in Virginia are presented in this chapter along with the results of treatability studies on the process effluents. Treatment alternatives investigated include gravity settling, filtration, coagulation by pH adjustment and anaerobic biological treatment. Analysis of sludge resulting from coagulation of retort effluent was also examined for its potential as a useful byproduct.

4.1 Plant Process Characterization

The results of the effluent characterizations from three blue crab processing plants are shown in Table 9 and Table 10. The characterization studies were done between July 1990 and July 1991. Overall, plant characterizations were done in plant #1 and plant #2. Retort water was also characterized at plant #3. The given flows and volumes correspond to the pounds of finished crab meat product processed during that day of production. The exception is for retort water, which is represented in terms of pounds of live crab per cook.

During this study, the total daily process volumes never exceeded 20,000 gallons per day (gpd) at plant #1 and never exceeded 10,000 gpd at plant #2. When no mechanized processes (Harris Claw and/or Quik-Pik operations) operated during a day, the total effluent volume was typically around 2,000 gpd. The daily process volumes were not measured at plant #3.

The temperatures of the various process effluents, other than the cooker discharge, were in an acceptable range (16-30°C). However, the temperature of the retort water ranged from 64-91°C. The high temperature is of concern when discharging to receiving waters. Some processors currently have NPDES permits which place limits on the discharge temperature of their retort water (32°C). The high temperature also complicates biological treatment methods since a cooling step would be required prior to discharge into a biological system.

The pH of all effluents, with two exceptions, remained within regulatory discharge limits of 6.0-9.0. On two occasions in the hand pick room of plant #1, pH values of 9.2 and 9.6 were measured and probably resulted from the use of a caustic floor cleaning agent. This relatively high effluent pH would likely be neutralized during mixing with other process wastewater before discharge.

Table 9. Characterization of typical daily effluents from three blue crab processing plants.

Process		(lbs)	(gpm)	(gal)	°C	
Type	Plant	Product	Flow	Volume	Temperature	pH
Harris Claw Reel Wash	2	2,380	1.8	650	30	7.5
	2	1,535	2.3	540	-	8.4
	2	760	-	390	-	-
Shell Waste Effluent	1	-	0.3	-	22	8.4
	1	1,550	0.5	135	16	8.2
	2	2,380	1.1	400	30	8.1
	2	1,535	2.2	520	-	7.9
	2	760	-	370	-	-
Harris Claw Brine Bath	1	-	-	230	20	8.3
	1	1,550	-	230	19	8.2
	2	2,380	-	220	30	-
	2	1,535	-	220	-	7.9
	2	760	-	220	-	-
Claw Meat Conveyor Wash	1	-	25.4	-	22	8.6
	1	1,550	33.5	9,715	19	8.3
	2	2,380	8.2	2,950	30	8.2
	2	1,535	8.8	2,070	-	8.2
	2	760	-	1,800	-	-
Claw Room Cleanup	1	-	7.8	-	22	8.6
	1	1,550	-	520	18	8.5
	2	2,380	-	-	-	-
	2	1,535	1.3	120	-	8.5
	2	760	-	360	-	-
Quik-Pik Bobber Effluent	2	-	2.3	-	30	7.0
	2	750	2.5	960	-	7.7
	2	490	-	280	-	-
Quik-Pik Room Cleanup	2	-	-	-	30	7.4
	2	750	-	-	-	7.9
	2	490	-	-	-	-
Retort Effluent	1	* 1,800	-	@ 70	91	8.6
	1	* 1,800	-	@ 72	74	7.0
	2	* 1,050	-	-	64	6.6
	2	* 1,050	-	@ 47	74	7.1
	3	* 1,300	-	@ 50	-	-
	3	* 1,300	-	@ 50	-	-
Hand Pick Room Cleanup	1	-	-	-	-	9.6
	1	-	-	280	18	9.2
	2	920	-	220	-	-

The * denotes pounds of live crab per cook, all other weights are of final product.

The @ denotes gallons produced per cook.

Table 10. Characterization of typical effluents from three blue crab processing plants.

Process Type	Plant	(mg/L)								
		COD	BOD-5	TSS	VSS	Cl	O&G	TKN-N	NH3-N	TP
Harris Claw Reel Wash	2	9,145	>3,740	3,970	3,400	620	-	-	-	-
	2	2,940	2,770	1,270	1,100	-	-	400	14	42
	2	15,410	>7,800	10,880	9,500	-	-	2,400	100	135
Shell Waste Effluent	1	15,690	15,350	18,340	5,160	112,170	10	-	-	-
	1	23,280	15,120	25,100	5,420	-	-	2,260	330	190
	2	13,005	>9,250	-	-	100,160	-	-	-	-
	2	17,260	13,500	13,740	2,660	-	-	2,240	125	170
	2	39,680	16,240	33,415	16,390	-	-	3,390	115	280
Harris Claw Brine Bath	1	14,980	7,000	17,460	4,390	143,990	6	-	-	-
	1	21,510	15,000	31,540	8,320	-	-	3,060	190	270
	2	-	7,805	-	-	135,020	-	-	-	-
	2	17,460	14,000	14,970	4,270	-	-	2,330	110	230
	2	15,870	8,925	12,670	3,740	-	-	1,775	80	160
Claw Meat Conveyor Wash	1	570	265	445	410	3,100	-	-	-	-
	1	650	640	660	540	-	-	150	< 10	7
	2	1,790	1,020	585	355	15,275	-	-	-	-
	2	2,520	2,040	930	480	-	-	370	22	25
	2	2,625	1,790	1,170	775	-	-	390	17	27
Claw Room Cleanup	1	3,920	1,260	1,375	630	26,770	-	-	-	-
	1	900	-	3,050	2,790	-	-	210	< 10	20
	2	-	-	-	-	-	-	-	-	-
	2	550	420	590	500	-	-	160	< 10	8
	2	-	-	-	-	-	-	-	-	-
Quik-Pik Bobber Effluent	2	9,755	7,000	5,130	4,490	590	-	-	-	-
	2	16,460	12,690	5,440	5,040	-	260	1,030	28	234
	2	29,655	16,980	12,000	10,860	-	-	2,000	53	321
Quik-Pik Room Cleanup	2	-	-	1,100	880	510	-	-	-	-
	2	620	410	410	370	-	-	50	< 10	7
	2	-	-	-	-	-	-	-	-	-
Retort Effluent	1	32,940	27,350	1,790	1,550	6,770	22	-	-	-
	1	35,240	-	6,200	4,710	-	10	3,940	160	185
	2	29,000	28,500	1,460	1,305	5,100	-	-	-	-
	2	21,510	17,380	1,010	910	-	50	2,240	70	102
	3	31,040	18,780	653	535	-	-	-	-	-
	3	23,920	13,720	1,980	1,640	-	-	2,510	130	160
Hand Pick Room Cleanup	1	2,750	1,510	2,610	1,975	370	-	-	-	-
	1	4,980	3,075	1,660	1,080	-	-	180	< 10	110
	2	3,680	2,480	1,240	1,000	-	-	250	10	125

Tables 10 and 11 show that blue crab processing typically produces relatively concentrated wastewaters. Similar results have been found by other researchers (Brinsfield, 1977; Carawan et al., 1979; Chao et al., 1983; Hanover et al., 1975; Horn and Pohland, 1973; Johnson and Gallanger, 1984; Wheaton et al., 1984). For example, effluent concentrations were as follows: BOD_5 = 265-28,500 mg/L, TSS = 410-33,400 mg/L, TKN-N = 50-3,390 mg/L, NH_3-N = < 10-330 mg/L and TP = 7-320 mg/L. The high BOD_5 and TSS values are of concern because they are currently regulated by the Virginia SWCB. The high ammonia values are a problem because it is expected that they will soon be regulated. High TKN-N and TP concentrations are also of concern to processors discharging to POTWs because new surcharges for these constituents will be implemented in July 1992. In the Hampton Roads Sanitation District (1992 rate schedule), industries will be required to pay surcharges for phosphorus in excess of 6 mg/L at a rate of $114.00 per hundred pounds and for TKN-N in excess of 35 mg/L at a rate of $31.13 per hundred pounds. Currently, BOD_5 and TSS surcharges in excess of 250 mg/L are $20.65 and $23.55 per hundred pounds, respectively.

The COD values were only slightly larger than the BOD_5 values in most cases and the VSS typically constituted a very high percentage of the TSS. The BOD_5 was 32-98% of the COD, and the VSS was 19-93% of the TSS. The results imply that the wastewater consisted largely of highly degradable organic matter.

The Harris Claw process displayed a great variability in effluent quality between and within the plants. The variability of this process may be related to time of year and location in which a particular batch of crabs was harvested, and whether they were put through the tumble spray washer upon arrival.

Only plant #2 actively used the Harris Claw reel washer during the site visits. This effluent was highly concentrated with organic matter. The COD of this wastewater ranged from 2,940-15,410 mg/L and the VSS was 85-88% of the TSS.

The Harris Claw shell liquid and brine bath contained extremely high chloride concentrations (100,000-144,000 mg/L). The brine bath was kept approximately 70% saturated with sodium chloride to maintain efficient separation of the meat and shell fragments. On one site visit to plant #2, it was noted that 3,440 pounds of salt were used to produce 2,250 pounds of claw meat. The high chloride concentrations are of concern

Table 11. Typical effluent loadings from three blue crab processing plants.

Process		(lbs/1,000 lbs product)								
Type	Plant	COD	BOD-5	TSS	VSS	Cl	O&G	TKN-N	NH3-N	TP
Harris Claw	2	20.8	>9	9.0	7.7	1.4	-	-	-	-
Reel Wash	2	8.6	8.1	3.7	3.2	-	-	1.2	0.04	0.12
	2	66.0	>33	46.6	40.7	-	-	10.3	0.43	0.58
Shell Waste	1	16.9	11.0	18.2	3.9	-	-	1.6	0.24	0.14
Effluent	2	18.2	>13	-	-	140	-	-	-	-
	2	48.8	38.1	38.8	7.5	-	-	6.3	0.35	0.48
	2	161.1	65.9	135.7	66.5	-	-	13.8	0.47	1.14
Harris Claw	1	26.6	18.6	39.0	10.3	-	-	3.8	0.24	0.33
Brine Bath	2	-	6.0	-	-	104	-	-	-	-
	2	20.9	16.7	17.9	5.1	-	-	2.8	0.13	0.27
	2	38.3	21.5	30.6	9.0	-	-	4.3	0.19	0.39
Claw Meat	1	34.0	33.5	34.5	28.2	-	-	7.8	< .52	0.37
Conveyor Wash	2	18.5	10.5	6.0	3.7	158	-	-	-	-
	2	28.3	22.9	10.5	5.4	-	-	4.2	0.25	0.28
	2	51.9	35.4	23.1	15.3	-	-	7.7	0.34	0.53
Claw Room	1	2.5	-	8.5	7.8	-	-	0.6	< .03	0.06
Cleanup	2	0.4	0.3	0.4	0.3	-	-	0.1	< .01	0.01
Quik-Pik	2	175.7	135.5	58.1	53.8	-	2.78	11.0	0.30	2.50
Bobber Effluent	2	141.3	80.9	57.2	51.8	-	-	9.5	0.25	1.53
Retort **	1	10.7	8.9	0.6	0.5	2.2	0.007	-	-	-
Effluent	1	11.8	-	2.1	1.6	-	0.003	1.3	0.05	0.06
	2	8.0	6.5	0.4	0.3	-	0.019	0.8	0.03	0.04
	3	10.0	6.0	0.2	0.2	-	-	-	-	-
	3	7.7	4.4	0.6	0.5	-	-	0.8	0.04	0.05
Hand Pick Room Cleanup	2	7.3	4.9	2.5	2.0	-	-	0.5	0.02	0.25

** All units are lbs/1,000 lbs product except for retort water which are lbs/1,000 lbs of live crab.

because of the potential toxicological effects to microorganisms in biological treatment systems.

The shell waste effluent and the brine bath contained the highest TSSs of any effluent. The VSS concentration was also high since significant quantities of meat particles remain uncollected by the meat conveyor. These effluents contained a relatively large percentage of non-volatile suspended solids. The VSS concentration was only 19-30% of the TSS concentration (49% in one case). In comparison, the Harris Claw reel wash VSS concentration comprised 85-88% of the TSS concentration. The Harris Claw reel wash effluent differs in that no salt is added at that stage of the process and the claws are not yet shattered in the hammer mill. The non-volatile suspended solids in the shell waste effluent and brine bath resulted from inert shell components (crab scrap is 40-50% calcium carbonate) and precipitated salt.

The claw meat conveyor wash effluents displayed high variability. The differences result from plant #1's use of four to five times more conveyor wash water to process the same amount of product. However, the loadings per 1,000 pounds of product from each plant for this effluent were approximately the same because the high volume effluent was more dilute (Table 12). In the plants that used the Harris Claw process, the claw meat conveyor wash water produced a greater effluent volume (1,800-9,715 gpd) than any other process. Since this effluent is relatively dilute, it offers potential for use in diluting some of the more concentrated effluents for biological treatment (especially for plant #1). However, the character of this relatively dilute effluent substantially exceeds permit limits and still has chloride concentrations that could be inhibitory/toxic in the wrong environment (3,100-15,275 mg/L).

The Quik-Pik process was used in plant #2 only during days of very high production to support the hand pick operations. The Quik-Pik bobber produced the only substantial effluent stream from this process. The bobber effluent was extremely concentrated (BOD_5 = 7,000-16,980 mg/L, TSS = 5,130-12,000 mg/L, TKN-N = 1,030-2,000 mg/L and TP = 235-320 mg/L) and contributed 2-16% of the total plant effluent COD loadings when in use. The bobber effluent was also the only plant process with significant concentrations of oil and grease (260 mg/L). The Quik-Pik operations resulted in greater wastewater loading than

Table 12. Average loadings and volumes from each effluent per 1,000 pounds of final product.

Process Type	Plant	(lb/1000 lb) COD	(lb/1000 lb) TSS	(gal/1000 lb) Volume
Harris Claw	1	-	-	-
Reel Wash	2	31.8	19.8	380
Shell Waste	1	17.0	18.2	87
Effluent	2	83.7	26.0	330
Harris Claw	1	26.6	39.0	460
Brine Bath	2	28.6	29.2	450
Claw Meat	1	34.0	32.0	6,270
Conveyor Wash	2	32.9	13.2	1,625
Quik-Pik	1	-	-	-
Bobber Effluent	2	158.5	52.8	925
Retort	1	* 11.3	* 1.3	* 40
Effluent	2	* 9.6	* 0.8	* 40
Cleanup Processes		(lb/day)	(lb/day)	(gal/day)
Claw Room	1	2.5	8.5	520
Cleanup	2	1.5	4.0	360
Quik-Pik Room	1	-	-	-
Cleanup	2	1.5	4.0	360
Hand Pick	1	7.3	2.5	280
Room Cleanup	2	7.0	2.5	250

* Retort effluent results are per 1,000 pounds of live crab.

hand pick operations because the viscera of the crab are incorporated into the wastewater in the Quik-Pik operation whereas the viscera go to the solid waste stream during hand picking.

The retort water was studied in more detail than other processes because it was produced in all plants and it was typically responsible for a significant percentage of the COD waste loadings from each plant (39-99%). It should be noted that the values in Table 10 are per cook of live crab. Each of these plants will typically have 6 to 20 cooks per day. This corresponds to a daily retort water volume of 400-1,000 gallons. The retort effluent character and volume remained very consistent within and between each plant. BOD_5 concentrations were greatest in the retort water, ranging between about 14,000-30,000 mg/L. The TKN-N, NH_3-N and TP values were also very high. The suspended solids values, however, were typically lower than most other processes (TSS = 650-2,000 mg/L).

The character of the retort effluent changed over a cooking cycle. A typical cooking cycle lasted 20-30 minutes. The retorts took approximately 10 minutes to reach 15 psi; the crabs are then cooked for 10 minutes followed by a cool down period. As the cook cycle proceeded, the TSS decreased, the BOD_5 increased and the discharge changed from liquid to steam.

Representative effluent volumes and concentrations for the Harris Claw cleanup, the Quik-Pik cleanup and hand pick cleanup were the most difficult of any process to obtain. In some of these process rooms, as many as five different hoses were used during cleanup. The runoff from these hoses often flowed into several different drainage channels and the character of the cleanup water changed continuously during wash down. The quantity of water used for wash down also varied each processing day. This variation was probably due to such factors as the individual doing the wash down, how dirty the process room had become, and the particular plant studied. Typically, grab samples were taken near the beginning of cleanup and the quantity of water used was determined by flow meters or by timing the use of hoses for which the flow rate had been previously determined. Composite sampling methods should be performed to obtain more reliable results in the future. Cleanup wastewater contributed 3-8% of the COD waste load from plant #1 and 2-6% from plant #2. However, the cleanup wastewater contributed 5-44% of the wastewater volume from plant #1 and 11-26% of the wastewater volume from plant #2. The cleanup wastewater contributed the largest percentage to the total wastewater volume when the mechanized processes were not in operation.

Table 12 shows the average COD and TSS loadings and volumes from each effluent stream for plants #1 and #2. The loadings and volumes are presented in pounds or gallons per 1,000 pounds of final product except for retort effluent which is presented in pounds or gallons per 1,000 pounds of live crab. The process cleanup wastewater loadings and volumes were presented in pounds or gallons per day based on the assumption that the daily cleanup volume and loadings are independent of the amount of product processed. Plant #1 did not have Harris Claw reel wash, Quik-Pik bobber or Quik-Pik cleanup wastewater streams.

A comparison of the loadings and volumes produced per 1,000 pounds of product by "equivalent" wastewater streams at plant #1 and plant #2 show similar results with a few exceptions. First, almost four times more Harris Claw conveyor wash water was used at plant #1 to wash the same amount of product. No studies were performed to compare the quality of the resulting claw meat products from these two plants.

Second, the shell waste effluent COD loading and volume from the Harris Claw process were four to five time greater at plant #2. The increase was probably due to relatively different mechanical operation of the shell removal conveyor belts at each plant. The conveyor at plant #2 brought a larger wastewater volume out of the brine tank than at plant #1 even though the average COD concentrations in the shell waste effluent were very similar. The increased effluent volume resulted in a relatively larger effluent COD and TSS loading at plant #2.

The average COD loading from the Harris Claw process was 77 and 177 lb/1,000 lb of final product at plant #1 and plant #2, respectively. The COD loading on a per weight basis was greater at plant #2 because of the additional Harris Claw reel wash effluent and the larger shell waste effluent volume. Interestingly however, the average Harris Claw process effluent volumes were much greater at plant #1 due to the large volume of Harris Claw conveyor wash (6,590 and 2,790 gallons/1,000 lb of final product at plant #1 and plant #2, respectively).

Wastewater volume and strength from crab processing plants varied considerably during the year due to seasonal variations in the plant production rate. Average daily total effluent volumes at plant #1 based on two years of production data are shown in Figure 5. Average daily total COD and total TSS waste loads from plant #1 are shown in Figures 6 and 7.

The flows and loads were normalized over 30.5 days/month. That is, the results are presented as if the plant processed a constant amount every day of the month. On any given day, the process load and volume could be higher than the average. In the spring and fall, the plant operated about ten days per month, but the averages were calculated as if they operated 30.5 days/month. In the summer, a given daily process volume and load would be close to the average since the firm processed nearly every day during that period.

Peak wastewater flow occurred at plant #1 during August at 11,960 gal/day. Peak COD and TSS loads occurred in August at 254 and 173 lb/day, respectively. The plant did not operate from January through March.

The Harris Claw process and cleanup contributed 79-95% of the wastewater volume from June through October when it was used at least 17 days a month. The operation contributed 13-52% of the COD and 70-91% of the TSS during the same period. The remaining wastewater consisted of the retort effluent and the hand pick cleanup. The results indicate that the Harris Claw process contributed the majority of the wastewater volume and TSS load at plant #1, but the COD load results primarily from the retort effluent.

The total effluent COD concentration from plant #1 was calculated to range from 2,500-3,700 mg/L from June through October when the Harris Claw process was used heavily. The total effluent COD concentration increased to 5,500-16,500 mg/L during months when the Harris Claw process was used less than 14 days per month. The concentration increase resulted from the decreased use of the Harris Claw process and the corresponding decrease in dilution effects from the Harris Claw conveyor wash.

Average daily total effluent volumes based on three years of production data (four years for retort data) for plant #2 are shown in Figure 8. The average daily total COD and total TSS waste loads from plant #2 are shown in Figures 9 and 10. These volumes and loads were also normalized over 30.5 days/month. Plant #2 operated all year long with the exception of the Quik-Pik process which was typically suspended from January through May.

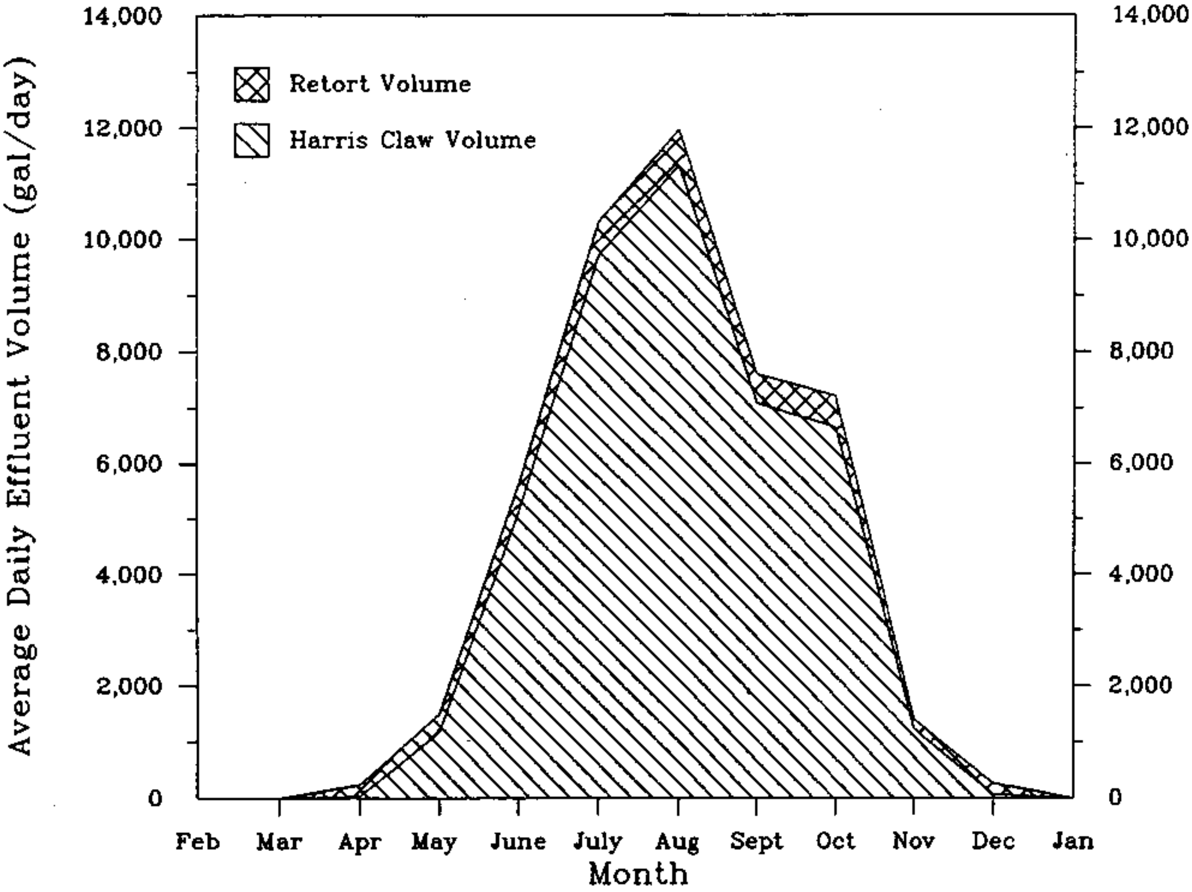


Figure 5. Average daily effluent volume from plant #1.

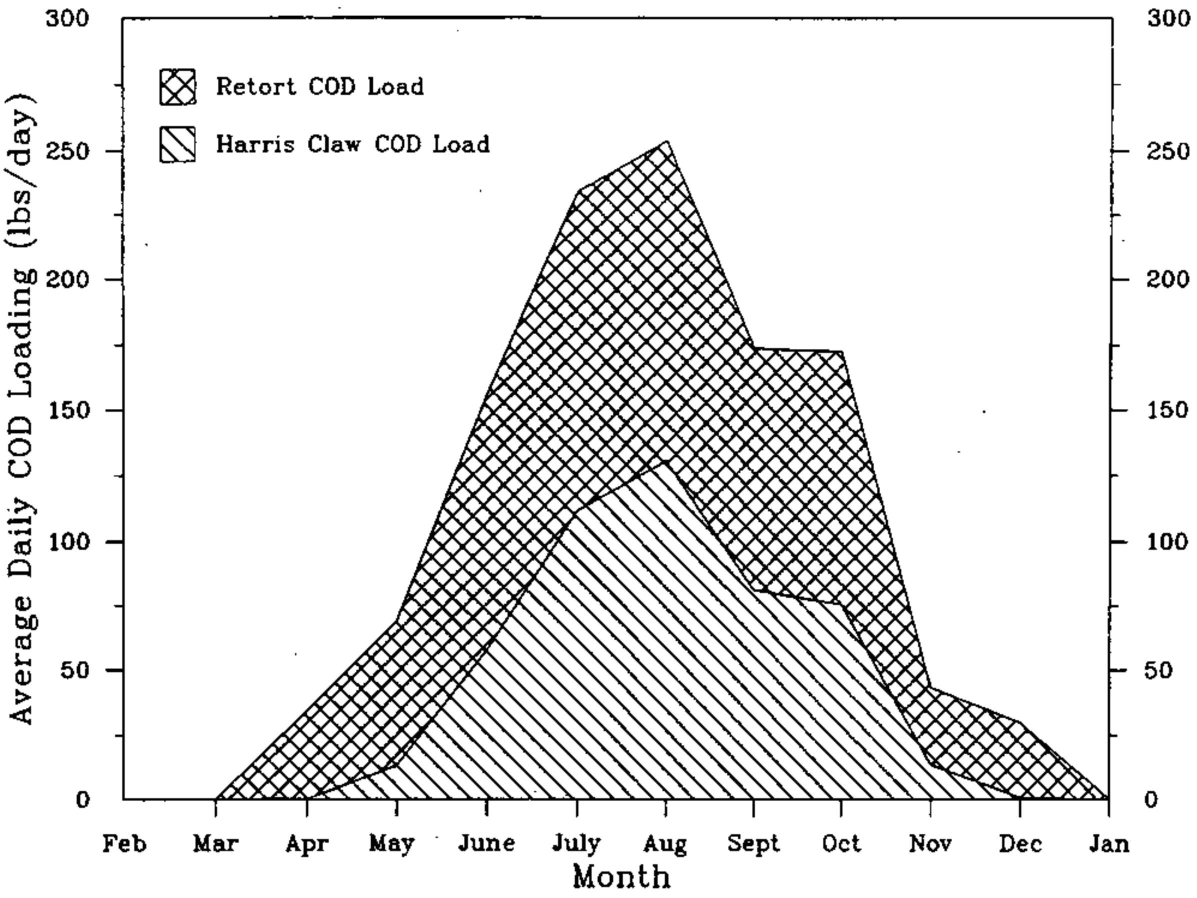


Figure 6. Average daily COD loading from plant #1.

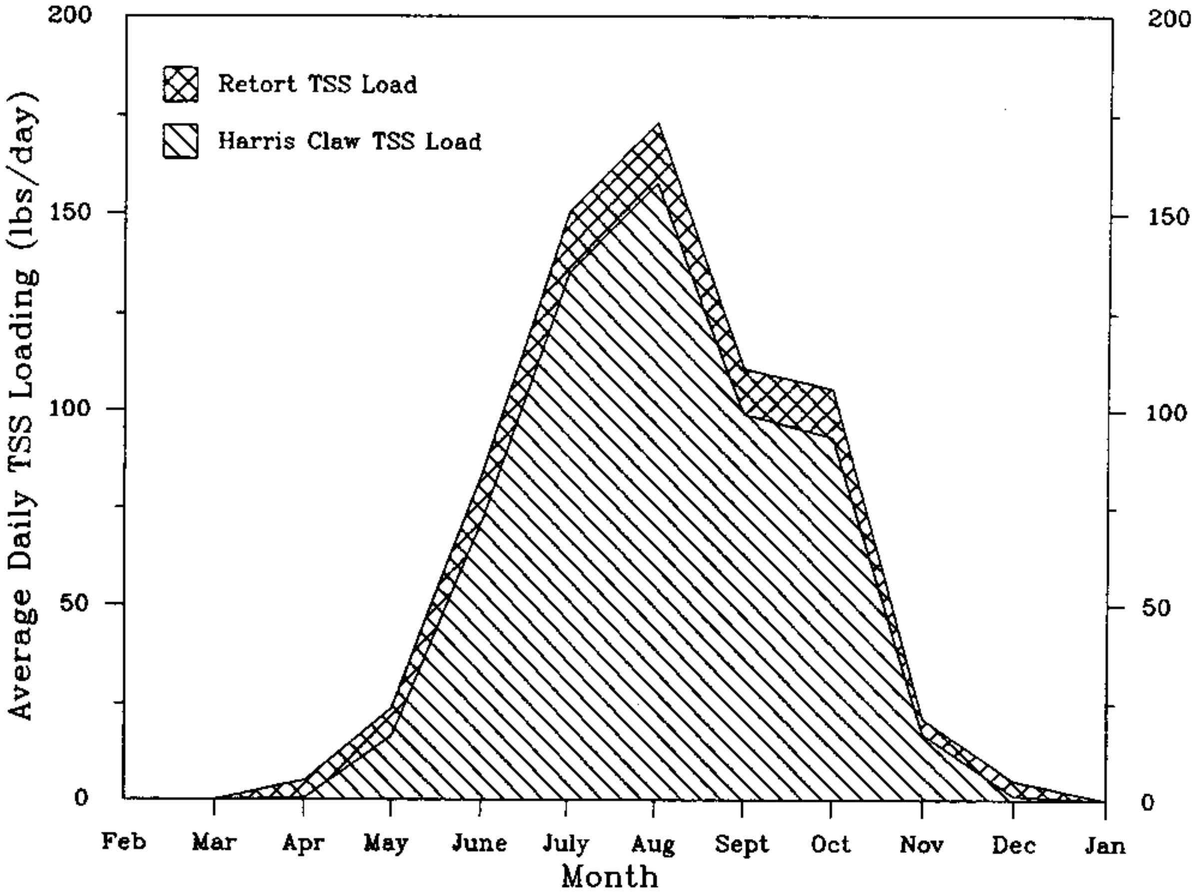


Figure 7. Average daily TSS loading from plant #1.

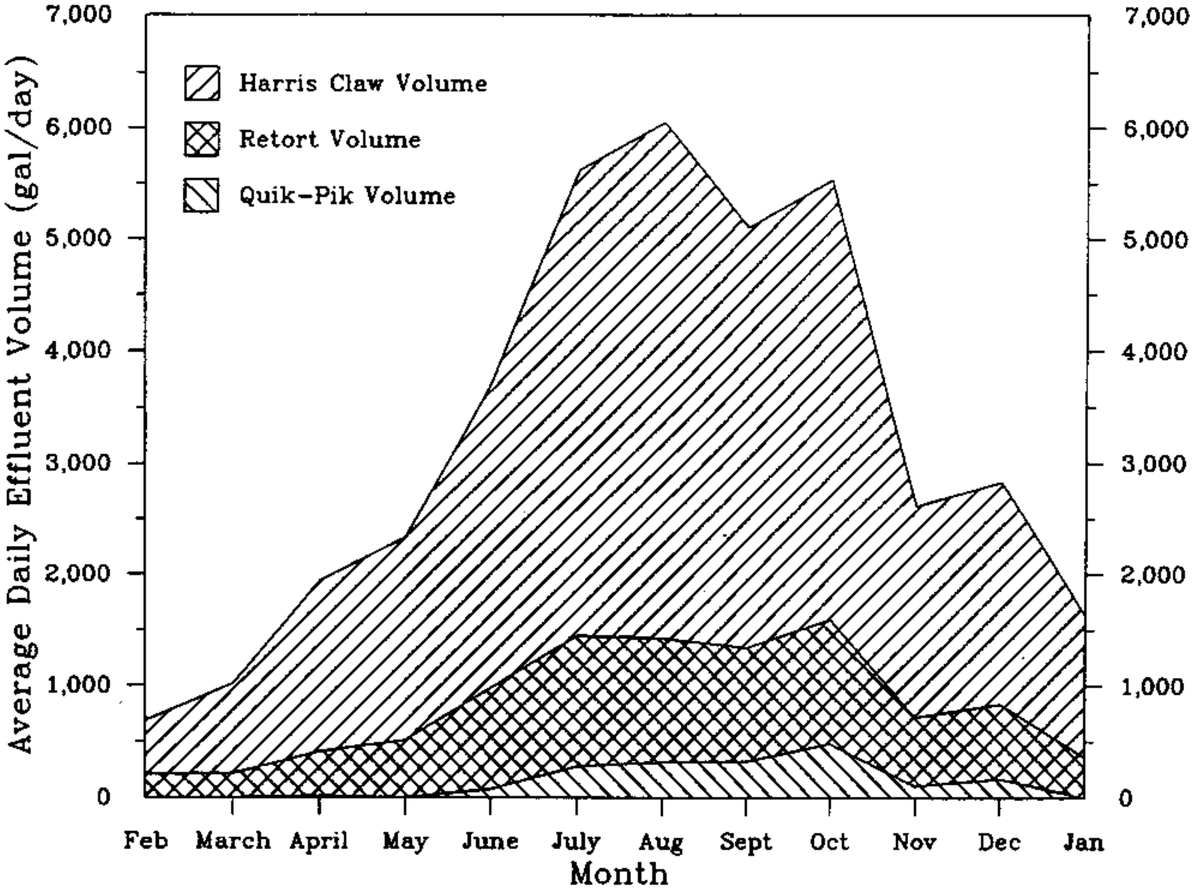


Figure 8. Average daily effluent volume from plant #2.

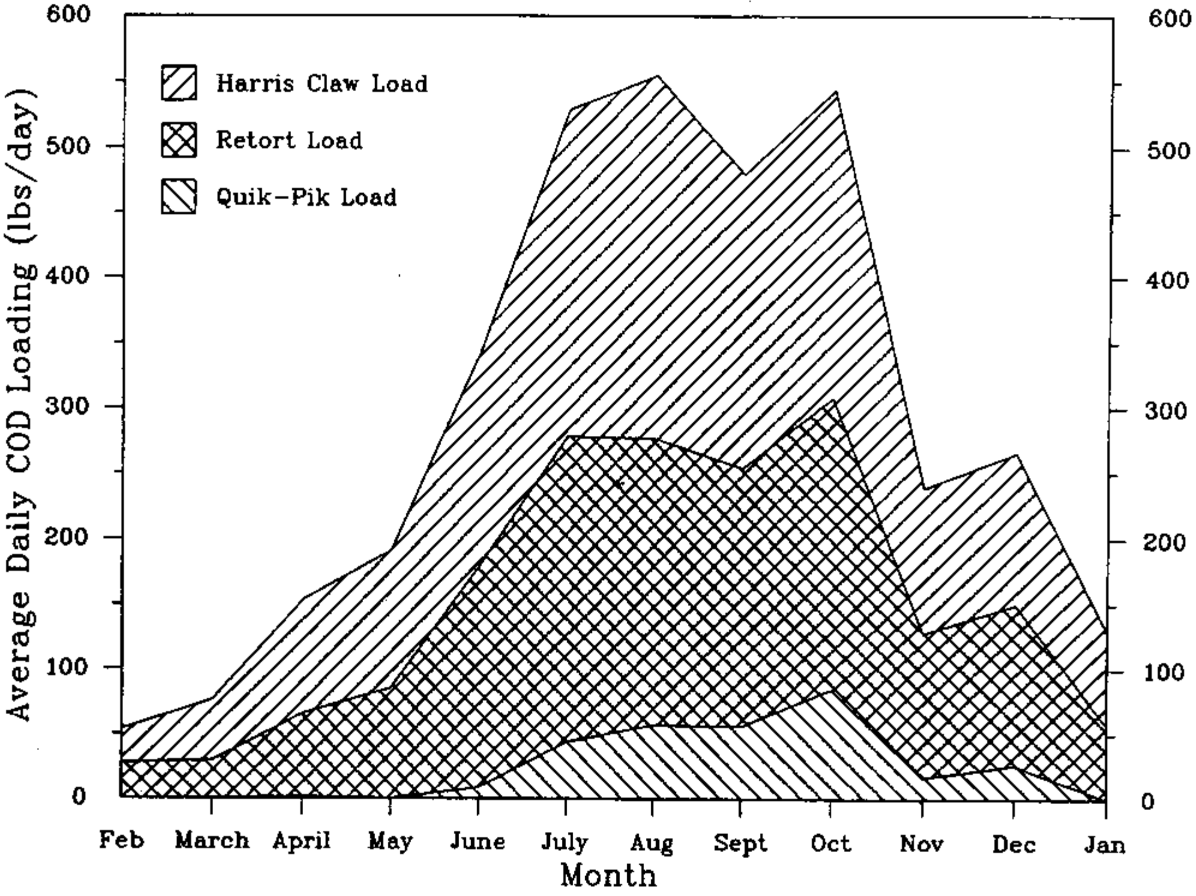


Figure 9. Average daily COD loading from plant #2.

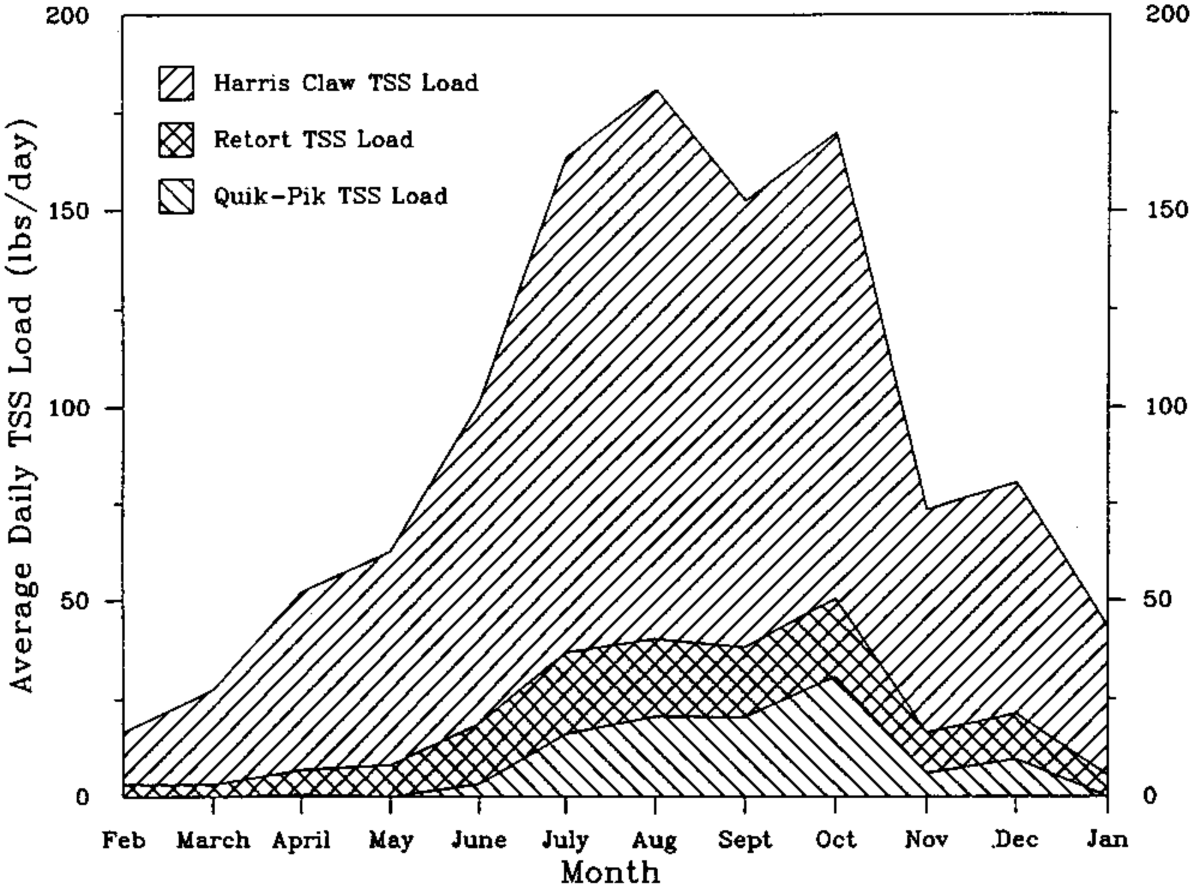


Figure 10. Average daily TSS loading from plant #2.

Peak average wastewater flow occurred in August at 6,050 gal/day and the minimum average flow was noted in February at 680 gal/day. The peak COD and TSS loads occurred in August at an average of 555 and 181 lb/day, respectively, and the minimum loads in February at an average of 54 and 17 lb/day, respectively.

At plant #2, the Harris Claw process and claw room cleanup contributed 70-79% of the wastewater volume, 43-61% of the COD load and 70-88% of the TSS load during the year. The retort effluent and hand pick cleanup contributed 20-30% of the wastewater volume, 32-49% of the COD load and 11-19% of the TSS load. Finally, the Quik-Pik process and cleanup contributed 0-9% of the wastewater volume, 0-15% of the COD load and 0-18% of the TSS load. The results indicate that the Harris Claw process contributed the majority of the wastewater volume and TSS loading, but only about half of the COD load. The more concentrated retort and Quik-Pik effluents add considerably to the COD load.

The total effluent COD concentration from plant #2 was calculated to range from 9,000-11,800 mg/L during the year. The total effluent concentration was typically higher at plant #2 than plant #1, and the effluent volume lower because a much larger volume of Harris Claw conveyor wash was used at plant #1. As expected, the average daily COD and TSS loadings were greater at plant #2 even though the Harris Claw process produces less of load on a per weight of product basis than plant #1. The reason for the greater loading at plant #2 is that more crab meat was processed at plant #2 on an average daily and yearly basis and the additional Quik-Pik process was used during the year.

4.2 Treatability Studies

The three blue crab processing plants currently have no liquid waste treatment systems other than screens in some drains and hand sinks to capture large solids. The facilities discharge all effluent overboard into water bodies adjacent to the plant. Some dilute cleanup effluents from plants #2 and #3 discharge to municipal treatment systems. Certainly, a rigorous study of pretreatment and treatment alternatives for these effluents must be investigated. Some treatment studies are presented herein, including gravity settling and filtration of the major process effluents, coagulation by pH adjustment of the more concentrated effluents and anaerobic biological treatment of retort water and combined effluent. The treatability studies were performed from July 1991 to April 1992.

4.2.1 Settling and Filtration Tests

Settling tests were done to see if sedimentation basins offer promise for reduction of solids, degradable organics and nutrients. Results of the settling and filtration tests are shown in Table 13. Settling appeared to work fairly well (TSS removals ranged from 40-74%) for effluents that contained high initial solids concentrations such as those generated by Harris Claw and Quik-Pik operations. However, final solids concentrations were still well above current limits (TSS values ranged from 410-12,000 mg/L after settling). One very important result is that settling removed little COD and BOD_5 and is not a viable method for reducing these parameters to meet limits. Only 3-36% of BOD_5, 2-58% of the TKN-N and 3-33% of the TP was removed by settling.

Filtration tests were performed to examine the maximum treatment that would be achieved by the removal of all suspended solids. Filtration of initial samples through glass-fiber filters removed 18-65% of the COD, 9-49% of the BOD_5, 9-62% of the TKN-N and 10-66% of the TP. Even after filtration, significant concentrations of BOD_5 and nutrients remained (after filtration: BOD_5 = 1,260-17,000 mg/L, TKN-N = 180-3,040 mg/L and TP = 22-270 mg/L).

The results from the settling and filtration studies indicate that substantial quantities of soluble material are in these effluent and sophisticated and expensive solids removal equipment may not be practical for meeting permit limits. The development of treatment systems for these effluents should focus on biological and/or chemical treatment methods.

4.2.2 Coagulation by pH Adjustment

Coagulation by pH adjustment of samples with high concentrations of soluble organics was evaluated on the basis of how well COD, BOD_5, TSS, VSS, TP, TKN-N and NH_3-N were removed. The results of two trials are shown in Table 14. It was determined that a pH range of 3.0 to 4.0 produced optimum removals in the retort effluent, brine bath and claw reel wash. It should be noted that all samples were settled prior to pH adjustment so that removals due to acidification alone could be determined. COD and BOD_5 reductions ranged from 12-37% and 6-30%, respectively. TSS and VSS reductions ranged from 76-93% and 76-92%, respectively. Over 90% removal of TSS were achieved for the Harris Claw Brine Bath. Removals of TKN-N, NH_3-N, and TP ranged from 0-24%. Coagulation

Table 13. Settling and filtration of effluent samples from a blue crab processing plant.

	(mg/L)							(ml/L)
DESCRIPTION	COD	BOD-5	TSS	VSS	TKN-N	NH3-N	TP	Set. Sol.
Harris Claw Reel Wash	15,410	>7,880	10,880	9,500	2,399	97	133	102
Settled	12,335	>3,900	2,850	2,450	1,006	110	111	-
Filtered	8,335	>3,900	-	-	916	96	71	-
Shell Waste Effluent	39,680	16,240	33,414	16,386	3,391	114	278	198
Settled	20,645	12,533	11,971	3,357	2,371	128	185	-
Filtered	17,135	9,553	-	-	2,172	120	140	-
Brine Bath	15,871	8,923	12,667	3,737	1,773	77	157	45
Settled	14,080	7,904	7,610	1,700	1,491	76	131	-
Filtered	11,467	6,867	-	-	1,305	74	93	-
Claw Meat Conveyor Wash	2,663	1,792	1,170	775	388	17	27	12
Settled	2,137	1,461	430	225	269	25	24	-
Filtered	2,059	1,322	-	-	252	19	22	-
Quik-Pik Bobber Effluent	29,655	16,980	12,000	10,860	1,998	53	321	72
Settled	22,938	13,725	5,360	5,240	1,435	46	221	-
Filtered	10,240	9,050	-	-	1,025	49	110	-
Retort Effluent (Trip #1)	31,040	18,780	653	534	-	-	-	1
Settled	25,858	18,181	624	497	3,069	165	298	-
Filtered	24,465	17,000	-	-	3,034	163	269	-
Retort Effluent (Trip #2)	23,921	13,720	1,980	1,640	2,511	131	157	-
Settled	21,463	13,000	780	645	2,455	130	153	-
Filtered	19,497	12,400	-	-	2,277	122	124	-
Hand Pick Cleanup	3,680	2,482	1,238	1,000	249	10	123	24
Settled	3,128	1,585	452	356	218	9	112	-
Filtered	2,508	1,262	-	-	179	9	91	-

Samples settled for 1 hour, then filtered through Whatman grade 934AH glass-fiber filters.

Table 14. Removals resulting from coagulation by pH adjustment.

	Concentration (mg/L)						
pH	TSS	VSS	COD	BOD-5	TP	TKN-N	NH3-N
TRIAL #1 - 10/23/91							
Type: Retort Effluent							
8.5	1,116	974	36,000	25,290	194	4,580	330
6.0	1,040	926	36,000	24,600	194	4,500	330
5.0	241	212	33,710	23,820	197	4,420	255
4.0	266	235	31,550	22,440	180	4,040	250
3.0	1,560	1,440	32,750	23,425	181	4,010	260
2.0	338	295	33,310	22,735	175	4,140	260
Type: Harris Claw Brine Bath							
8.2	9,195	1,445	13,925	6,950	144	1,845	90
6.0	8,895	1,435	13,625	6,920	141	1,815	90
5.0	2,810	353	12,725	6,870	140	1,600	90
4.0	690	90	10,980	5,910	136	1,475	88
3.0	945	119	11,090	6,340	134	1,500	90
2.0	1,030	125	11,080	6,200	131	1,465	88
Type: Harris Claw Reel Wash							
6.7	400	368	4,400	2,860	69	630	130
6.0	377	355	4,400	2,835	69	630	126
5.0	126	117	2,800	2,105	65	555	118
4.0	50	46	2,800	2,030	62	525	115
3.0	86	80	3,000	2,105	64	540	118
2.0	487	467	3,100	2,775	67	615	116
TRIAL #2 - 11/25/91							
Type: Retort Effluent							
5.0	722	606	7,860	4,725	36	856	33
4.0	340	300	5,790	4,115	34	773	36
3.0	122	94	4,965	3,860	36	717	33
2.0	121	96	5,540	4,020	38	773	30
Type: Harris Claw Brine Bath							
8.3	8,990	1,360	11,580	6,410	80	1,546	97
5.0	3,325	365	9,815	6,410	78	1,397	96
4.0	920	190	8,825	5,840	81	1,297	97
3.0	1,050	120	7,390	5,775	72	1,270	96
Type: Harris Claw Reel Wash							
7.7	1,170	1,050	5,955	3,815	62	580	30
5.0	93	87	4,960	2,750	59	516	32
4.0	58	49	4,260	2,615	60	518	31
3.0	83	78	3,722	2,660	58	483	29

Data in the first row of each effluent type is the control with no pH adjustment.

All tests were done with previously settled samples from plant #2.

may be an effective pretreatment method for some of these effluents, particularly the Harris Claw brine bath and shell waste effluents. The Quik-Pik bobber effluent was not studied, but may also be treated effectively with coagulation by pH adjustment because it had high initial TSS concentrations.

The results of this study were similar to the results from the settling and filtration experiments for the removal of suspended solids. Even with good removal of TSS, high BOD_5 and nutrients remained. The finished effluents were still well above permit limits (BOD_5 = 2,030-22,440 mg/L and TSS = 50-1,050 mg/L after acidification), but acidification shows promise as a pretreatment method to reduce sewer surcharges. Also, the sludge produced from the process may be marketable as a feed amendment, protein source or flavor compound.

4.2.3 Retort Effluent Sludge Analysis

A proximate analysis of supernatant and resulting sludge produced by pH adjustment of the retort effluent to pH 4.0 was performed. After removing all water, the retort effluent was comprised of 41% crude protein, 26% ash and 33% other material including fat, nitrogen free extract and acid digestible fiber on a dry weight basis. After coagulation, the supernatant consisted of 33% crude protein and 56% ash as the major components. The corresponding sludge was about 50% crude protein and 49% ash.

It is apparent from these results that crude proteins are the major constituent being removed from the retort effluent during pH adjustment because there was a decrease in percentage of protein in the supernatant on a dry weight basis. Also, the percentage ash on a dry weight basis increased in the supernatant, as would be expected, since less crude protein remains. However, even after coagulation, a significant quantity of crude protein remained in the supernatant (33% on a dry weight basis) which helps to explain the high BOD_5 and nutrient concentrations remaining in the effluent after nearly complete TSS removal.

The high percentage of crude protein in the sludge suggests that it may be a good animal feed. However, there is concern over the composition of the ash which also comprised a large percentage of the dried sludge. The composition of the sludge should therefore be analyzed for inorganic compounds that could render it useless as a feed.

4.2.4 Anaerobic Biological Treatment

4.2.4.1 Short-Term Anaerobic Tests

The performance of four 2.O L anaerobic reactors treating retort effluent and combined effluent, both before and after coagulation by pH adjustment, was monitored for 55 days. Alkalinity and MLVSS along with the total daily gas production and soluble, effluent COD for each reactor are shown in Figures 11-14. The F/M was increased from 0.15 to 0.25 on day 31 in reactors A and B because the soluble effluent COD was steadily decreasing. After the loading was increased, the soluble, effluent COD remained steady, ranging between 620-805 mg/L COD for reactor A and between 740-910 mg/L for reactor B. The soluble, effluent COD fluctuated between 360-895 mg/L COD for reactor C and between 550-1,070 mg/L in reactor D.

Efforts were made to maintain the MLVSS at 4,000 mg/L. The MLVSS in reactors A and B remained fairly constant after day 31 at approximately 3,750 mg/L. The MLVSS in reactors C and D gradually decreased to 3,300 mg/L and 3,000 mg/L, respectively. The decrease was probably due to both the low F/M of 0.05 and the high salt concentration. Low F/Ms typically result in slow and dispersed growth of biomass. Also, high salt concentrations inhibit coagulation and settling. The combination of low F/M and high salt concentrations created a situation in which more MLVSS was lost in the effluent than could be maintained in reactors C and D.

Average effluent VSS from reactors C and D were 334 mg/L and 382 mg/L, respectively, whereas the average effluent VSS concentrations from reactors A and B were 123 mg/L and 136 mg/L, respectively. The effluent from reactors A and B was transparent and yellow to green in color (average effluent TSS from reactors A and B were 156 and 194 mg/L, respectively) whereas the effluent from reactors C and D was opaque and gray to black in color due to high effluent TSS (average effluent TSS in reactors C and D were 802 and 1,030 mg/L, respectively).

It should be noted that the influent to reactors C and D had Na^+ concentrations of 24,000 mg/L and 21,400 mg/L, respectively (3,400 mg/L and 3,240 mg/L Na^+ in A and B influent). By day 55, the Na^+ concentration exceeded 9,000 mg/L in reactor C and 10,000 mg/L in reactor D. High salt concentrations are known to inhibit flocculation. Also, Na^+

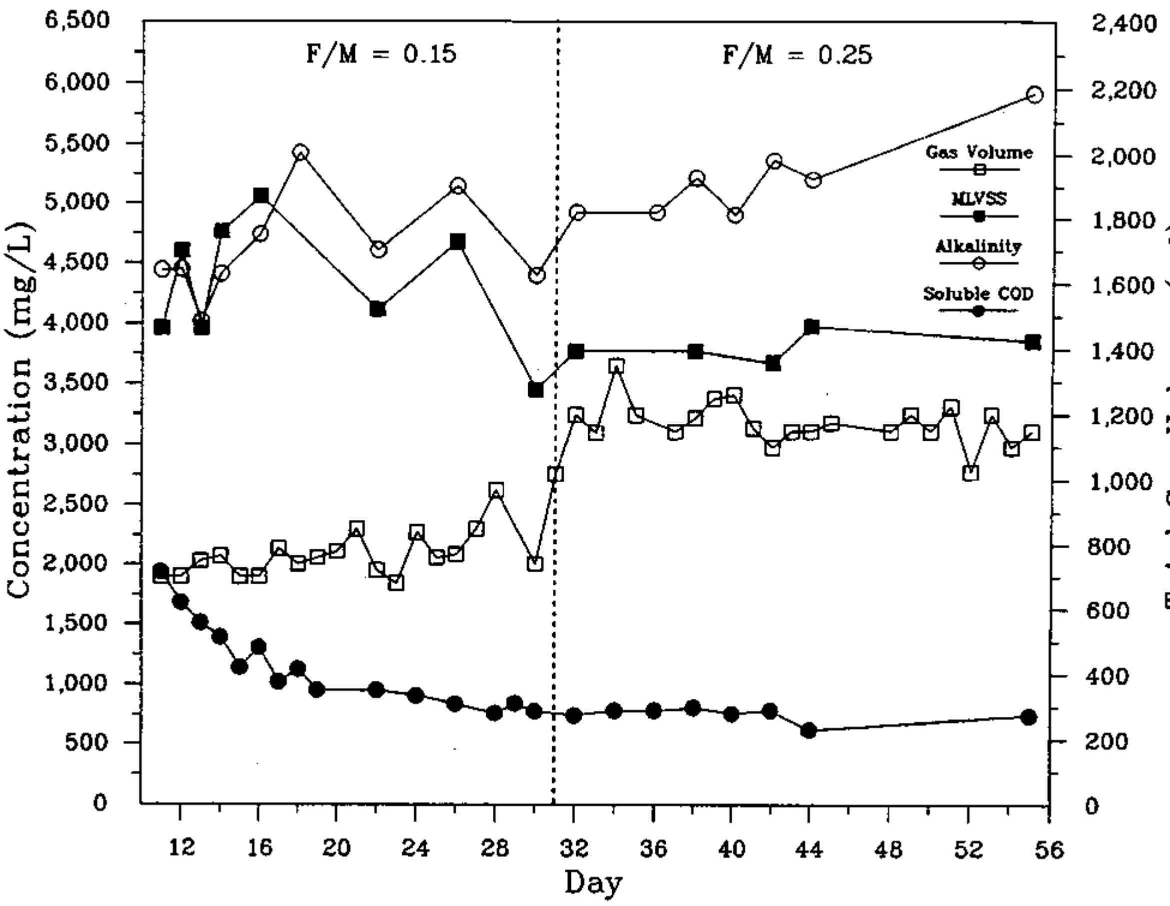


Figure 11. Anaerobic treatment of retort effluent (Reactor A).

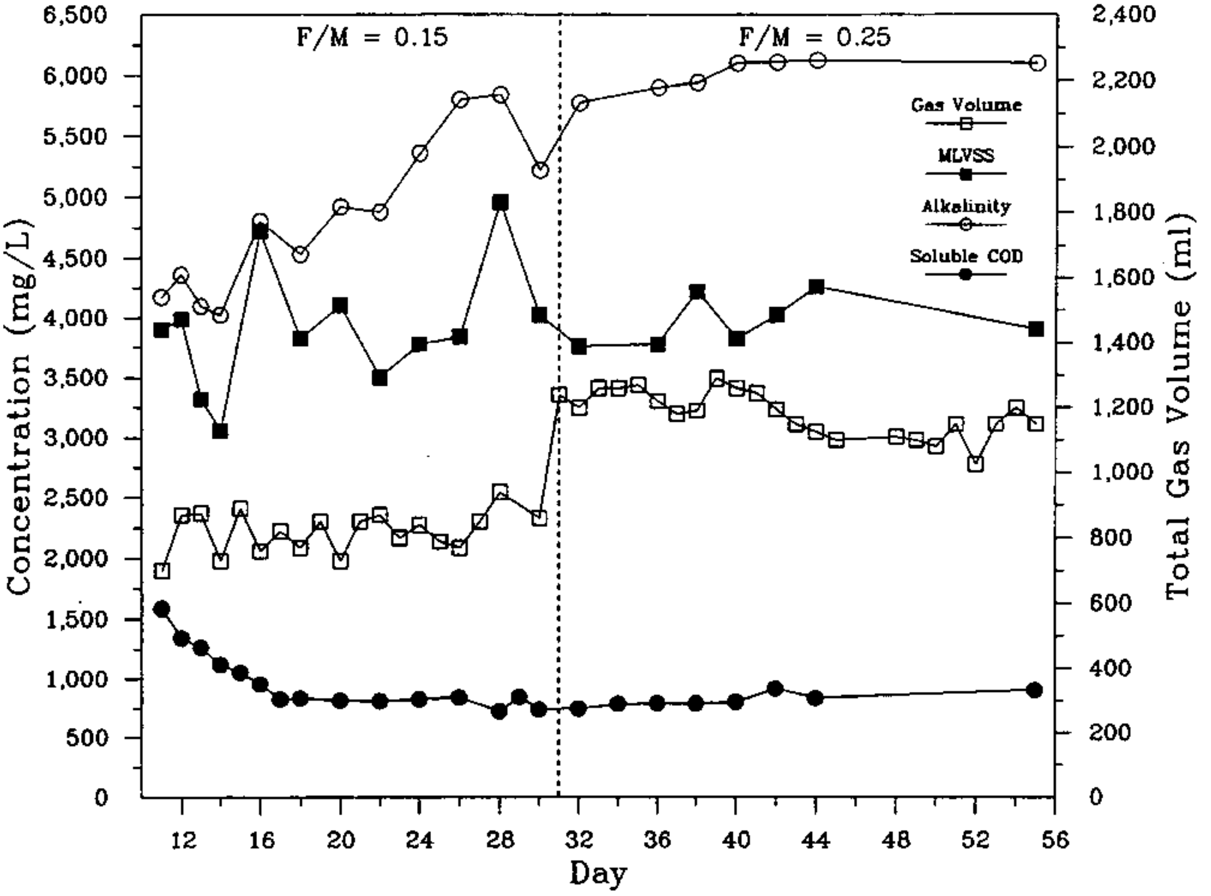


Figure 12. Anaerobic treatment of retort effluent after coagulation by pH adjustment (Reactor B).

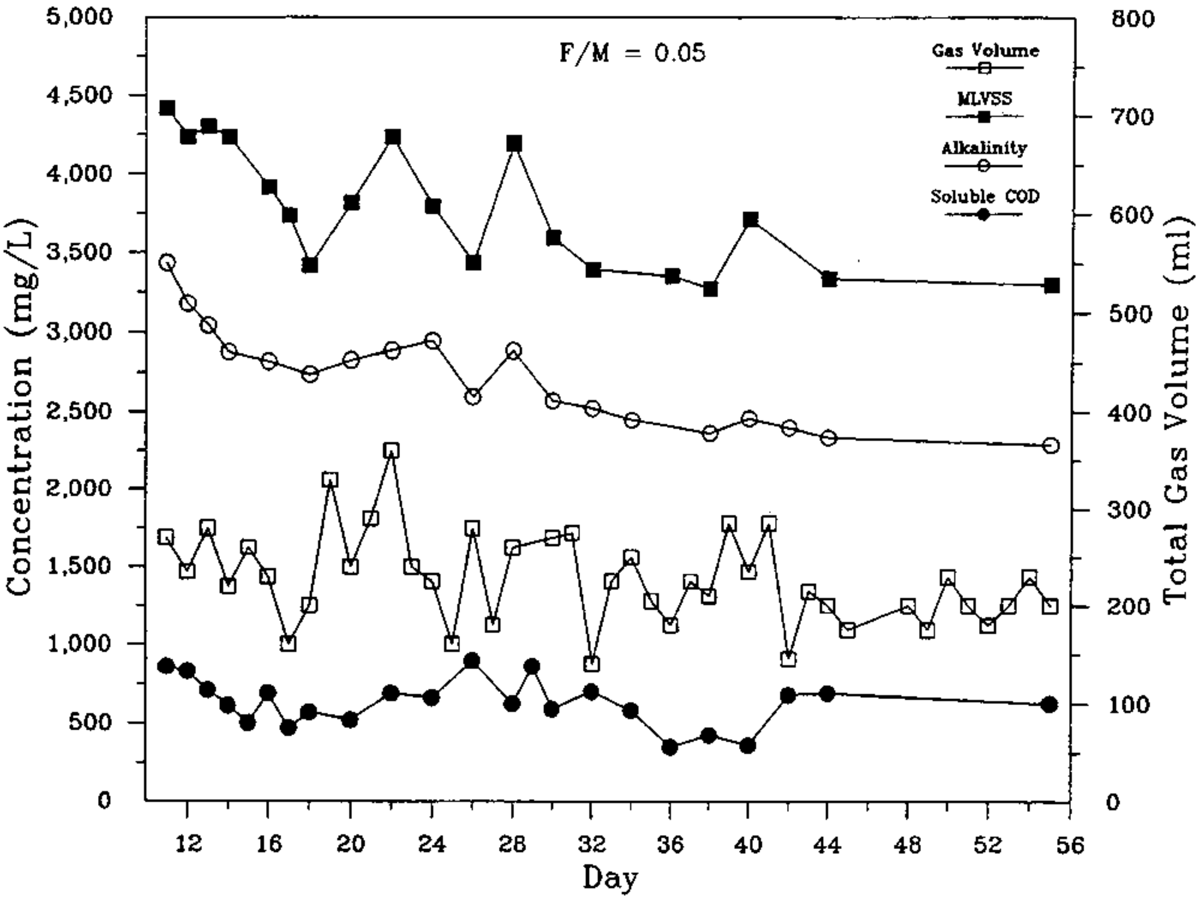


Figure 13. Anaerobic treatment of combined effluent (Reactor C).

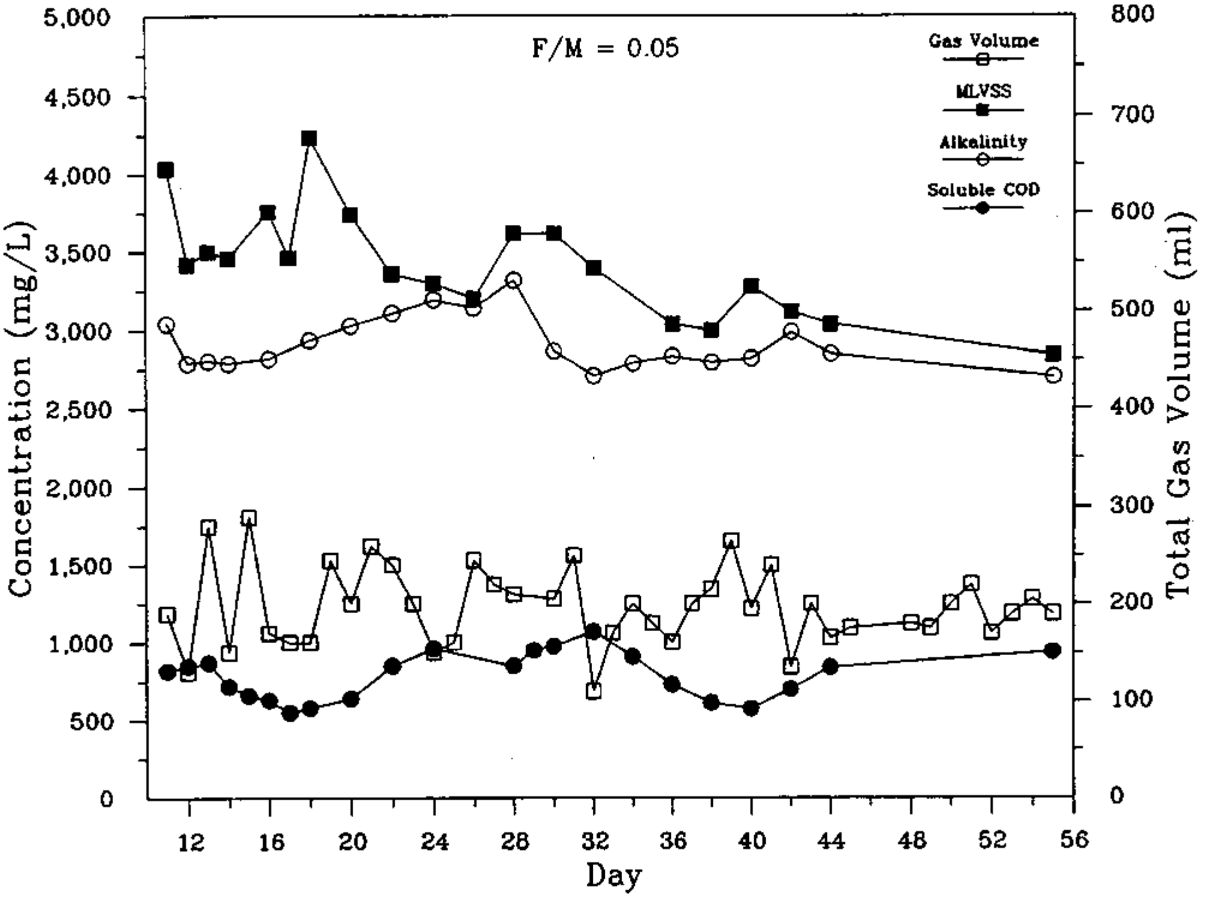


Figure 14. Anaerobic treatment of combined effluent after coagulation by pH adjustment (Reactor D).

concentrations have been shown to be inhibitory to methane fermentation in the range of 3,500-5,500 mg/L and strongly inhibitory above 8,000 mg/L (McCarty, 1964).

An average of 0.57-0.60 L total gas was produced per gram of total COD degraded in each reactor. McCarty and Young (1964) predicted that 0.348 liters of methane are produced per gram of COD at STP. Correcting for a reactor temperature of 35°C, and assuming 70% methane in the biogas, 0.56 L of total gas per gram of COD is predicted. The predicted gas production per gram of COD therefore closely corresponds with the levels seen in this study.

The TKN-N in feeds A and B was approximately 2,000 mg/L. NH_3-N was approximately 200 mg/L. The TKN-N in feeds C and D was approximately 800 mg/L and NH_3-N was approximately 90 mg/L. The TKN-N in the effluent from reactors A and B steadily increased from 800 mg/L on day 21 to 1,200 mg/L on day 55. Approximately 90% of the effluent TKN-N was in the form of NH_3-N. The effluent TKN-N in reactors C and D remained steady from day 21-55, ranging from 400-500 mg/L. NH_3-N comprised approximately 90% of this total. Organic nitrogen in the influent was highly available to microorganisms, as indicated by the high percentage of NH_3-N in the effluent. However, direct discharge of water containing high NH_3-N levels would not be permitted.

The TP in feed A, B, C, and D were 109, 94, 58 and 51 mg/L, respectively. On day 22 the TP in effluent A, B, C and D was 42, 49, 43 and 47 mg/L, respectively, and on day 55 was 52, 50, 34 and 40 mg/L, respectively. The results suggest that phosphorus was not a limiting nutrient in the systems.

The total alkalinity in the reactors appeared to be a function of both MLVSS and ammonia in the reactors. Fluctuations in total alkalinity in all four reactors mirrored fluctuations in ammonia with the exception of reactors A and B after about day 40. At that time the NH_3-N concentration had increased to approximately 1,000 mg/L in reactors A and B. The increase in alkalinity resulted from the increase in NH_3-N and bicarbonate (interaction of NH_3 and volatile acids with carbonate system components) in the system.

The total alkalinity in reactors A and B increased from 4,500 to 6,000 mg/L as $CaCO_3$ after day 40 even though the MLVSS remained at approximately 4,000 mg/L. The alkalinity in reactors C and D decreased from 3,500 mg/L to 2,700 mg/L as $CaCO_3$. This alkalinity decrease in C and D closely followed the decrease in MLVSS concentration. The ammonia levels in reactors C and D remained fairly steady and at much lower concentrations (NH_3-N = 350-400 mg/L) than in reactors A and B. Reactors A and B received a much greater TKN-N loading than reactors C and D which resulted in higher ammonia concentrations.

The concentration of four volatile acids was monitored in the reactors from day 11-55. On day 11, acetic acid, propionic acid, isobutyric acid and n-butyric acid were found in all effluents at concentrations ranging between 1-36, 0-230, 0-93 and 0-52 mg/L, respectively. The concentrations of these volatile acids gradually decreased until the systems stabilized at about day 25. During this stabilization period in all reactors, propionic and/or isobutyric acid were detected at greater concentrations than acetic acid. After day 25, acetic acid was always found in the highest concentration and all volatile acids in reactors A, B, and C effluents were below 10 mg/L. In reactor D, acetic acid fluctuated between 5-35 mg/L after day 25.

All reactors appeared to have stabilized because the volatile acid concentrations were low and steady. The systems appeared capable of degrading higher waste loads since the volatile acid concentrations were low. However, there was always a residual of volatile acids detected which provided the microorganisms with the intermediates necessary for metabolism. It was not determined what level of volatile acids were optimum for these systems. It should be noted that the acceptable volatile acid concentration is a function of alkalinity.

Soluble COD effluent values and gas production remained constant through day 55 in all reactors, indicating that virtually all of the organic material in the influent was degraded. The total BOD_5 and COD of the effluent from each reactor was substantial, however. Effluent BOD_5 and COD resulted from biomass in the effluent, as well as residual materials that were not degraded anaerobically. The average effluent total BOD_5 and total COD, respectively, were as follows: reactor A - 300 and 955 mg/L; reactor B - 375 and 1025 mg/L; reactor C - 195 and 900 mg/L; reactor D - 270 and 935 mg/L. These effluent BOD_5s may be permitted by regulatory agencies. One crab processor in Virginia has average and maximum BOD_5 permit limits of 300 and 600 mg/L, respectively. However, there is always a concern that those limits will become more stringent in the future. A second processor has average and maximum BOD_5 limits of 20 and 40 mg/L. It is clear that a polishing step would be required for these systems to further reduce effluent BOD_5, TSS and NH_3-N to meet current and expected future limits.

Using the conditions of this study (F/M = 0.05 for combined effluent, F/M = 0.25 for cooker effluent and MLVSS concentration = 4,000 mg/L), four different anaerobic treatment scenarios were postulated for plant #1 and plant #2 (Table 15) (Appendix C). It should be noted that the F/M values were achieved using feed which is typical of wastewater at plant #2. Since the COD concentration of the "combined" wastewater from plant #1 was expected to be less than plant #2 due to large dilution effects from the Harris Claw conveyor wash, the F/M values used in deriving the treatment scenario for plant #1 may not be valid. Still, a comparison between potential anaerobic biological treatment systems shows that blue crab processing facilities have different problems and requirements, and cannot all be treated in the same manner.

The first two scenarios for each plant consider treatment of all process effluents at a 0.05 F/M ratio. If the required reactor size was based on peak daily COD load, as show in scenario #1, a 152,500 gallon tank would be required for plant #1 and a 333,250 gallon tank would be required for plant #2. The corresponding HRTs would range from 13-610 days (high to low flow conditions) and 55-489 days for plants #1 and #2, respectively.

If the required reactor size was based on average daily COD load, as show in scenario #2, a 78,000 gallon tank would be required for plant #1 and a 178,340 gallon tank for plant #2. The corresponding HRTs range from 7-312 days and 29-261 days for plant #1 and plant #2, respectively.

The third and fourth scenarios consider treatment of cooker effluent only at a 0.25 F/M. At plant #1, based on peak daily COD loads, a 14,200 gallon tank (HRTs = 34-151 days) would be required to effectively treat cooker effluent, whereas a 8,650 gallon tank (HRTs = 21-92 days) results from a design based on average daily COD loads. At plant #2, based on peak daily COD loads, a 27,500 gallon tank (HRTs = 29-264 days) would be required to effectively treat cooker effluent, whereas a 14,900 gallon tank (HRTs = 16-143 days) results from a design based on average daily COD loads. The peak/average COD load ratio and the peak/average hydraulic load ratio ranged from 1.6-2.3 for both the combined effluent and the cooker effluent scenarios at plants #1 and #2.

Table 15. Design criteria for potential anaerobic treatment scenarios at blue crab processing plants #1 and #2.

	Combined Effluent *		Cooker Effluent **	
Scenario	#1	#2	#3	#4
PLANT #1				
Design Criteria	152,500 gallon reactor volume; based on peak load	78,000 gallon reactor volume; based on avg. load	14,200 gallon reactor volume; based on peak load	8,650 gallon reactor volume; based on avg. load
Max COD Load (lb/day)	254	254	118	118
Avg COD Load (lb/day)	130	130	72	72
Min COD Load (lb/day)	30	30	27	27
Max Flow (gal/day)	11,960	11,960	416	416
Avg Flow (gal/day)	5,132	5,132	254	254
Min Flow (gal/day)	250	250	94	94
Max Flow HRT (days)	13	7	34	21
Avg Flow HRT (days)	30	15	56	34
Min Flow HRT (days)	610	312	151	92
PLANT #2				
Design Criteria	333,250 gallon reactor volume; based on peak load	178,340 gallon reactor volume; based on avg. load	27,500 gallon reactor volume; based on peak load	14,900 gallon reactor volume; based on avg. load
Max COD Load (lb/day)	555	555	229	229
Avg COD Load (lb/day)	297	297	124	124
Min COD Load (lb/day)	54	54	25	25
Max Flow (gal/day)	6,048	6,048	954	954
Avg Flow (gal/day)	3,265	3,265	517	517
Min Flow (gal/day)	681	681	104	104
Max Flow HRT (days)	55	29	29	16
Avg Flow HRT (days)	102	55	53	29
Min Flow HRT (days)	489	261	264	143

Plant #1 and #2 COD load and flow averages calculated over 9 and 12 months of operation, respectively.

* F/M = 0.05 lb COD/lb MLVSS/day for scenario #1 and #2.

** F/M = 0.25 lb COD/lb MLVSS/day for scenario #3 and #4.

MLVSS concentration = 4,000 mg/L in all cases.

The reactor needed to treat combined waste at plant #2 would need to be approximately twice as large the reactor at plant #1 based on both peak and average conditions. The same holds true for reactors treating retort effluent alone.

With the exception of treatment of combined effluent during peak loads at plant #1 (7 and 13 days), the HRTs are fairly long (15-610 days). The HRTs were shorter at plant #1 than plant #2 because the combined wastewater was more dilute and voluminous. The Harris Claw conveyor wash at plant #1 was responsible for the majority of the dilution and volume effects. Given the long HRTs, reactor designs based on average daily COD loads should be able to maintain effluent quality during peak conditions. Additionally, a system using two smaller reactors can offer some advantages over one large tank, including the flexibility to shut one tank down during low flow or mechanical failure. However, a one reactor system could be operated at half volume during low loading conditions. Waste sludge from these reactors would be handled most efficiently by local POTWs due to the expense and involved operation and maintenance required of dewatering systems.

High NH_3-N and TSS concentrations from these systems were of concern. The NH_3-N concentrations in the system effluents ranged from 380-1,280 mg/L, greatly exceeding proposed limits. Acceptable ammonia concentrations in water are a function of pH, temperature and salinity. Recently adopted acute and chronic ammonia standards range from 3.3-16.0 and 0.5-2.4 mg/L NH_3-N, respectively, given the following conditions typical of ocean waters: pH = 7.8-8.2, temperature = 15-25°C and salinity = 30 gm/kg (Water Quality Standards Regulation, VR-680-21, 1992). Also, effluent TSS concentration was of concern for the combined effluent (TSS ranged from 680-1,050 mg/L) because the solids did not settle well, thereby necessitating additional treatment.

The reactors proposed for the blue crab processing plants considered may not be feasible due to economics and/or space limitations. Higher loading rates may be required to make anaerobic treatment viable.

4.2.4.2 Long Term Anaerobic Tests

The first two of the four reactors (A and B) used in this phase of the work were fed retort process wastewater and were operated for 216 and 94 days, respectively. Reactor A was fed at a 0.40 F/M ratio from day 1 through day 40, but after decreased treatment was observed, the reactor's F/M ratio was reduced to 0.35. Reactor B operated at a 0.25 F/M ratio over the course of its operation. Both reactors were able to achieve substantial reductions in the level of organics in the treated waste, but each eventually exhibited signs of failure and was shut down.

From day 48 through day 161, reactor A achieved COD removal rates ranging between 85% to 91% with an average of 87%. This removal translates to an effluent with a soluble COD largely in the range of 1550 to 2520 mg/L with an average of 2103 mg/L. Similarly, the reactor exhibited BOD_5 removal resulting in soluble effluent concentrations ranging between 1365 and 1497 mg/L over this time with an average removal of 88%, translating to an average effluent soluble BOD_5 of 1405 mg/L. Data on Effluent COD and BOD_5 concentrations are presented in Figure 15. Soluble material accounted for an average of 87% of the COD and 95% of the BOD_5 observed in reactor A effluent. These levels imply that the effluent solids contained a higher portion of inorganic material than would be suggested by the ratio found in solids from the secondary effluent of activated sludge systems (0.65 mg BOD_5/mg TSS). During this period of operation, the reactor produced biogas at an average rate of 1430 ml/day which is somewhat higher than the 1061 ml/day level that would correspond to the aforementioned removal rates (17.049 g/L COD * 0.180 L of feed/day * 88% removal * 0.393 L of gas produced/g of COD digested). This excess gas production can be attributed to the fact that total biogas production from anaerobic digestion can include byproducts such as carbon dioxide and hydrogen sulfide in addition to methane.

Figure 16 tracks gas production levels over the life of the reactor. Over this same period, effluent total suspended solids (TSS) varied between 400 and 520 mg/L with an average of 473 mg/L while MLVSS ranged from 3600 to 4800 mg/L with an average of 4042 mg/L. Solids retention time (SRT) varied between 96 and 248 days over the life of the reactor with an average value of 154.6 days. Hydraulic retention time (HRT) was maintained at an average of 12.2 days over the life of the reactor. Beginning on day 167, COD and BOD_5 levels steadily increased while biogas production decreased until the reactor was shut down on day 216. By this time the effluent COD had reached a level of over 9300 mg/L.

Reactor B similarly obtained substantial reduction in COD and BOD_5 levels from day 26 through day 67 of operation. Over this period of time, the reactor achieved COD removal rates of between 89% and 93% with an average of 91% and BOD_5 removal at rates

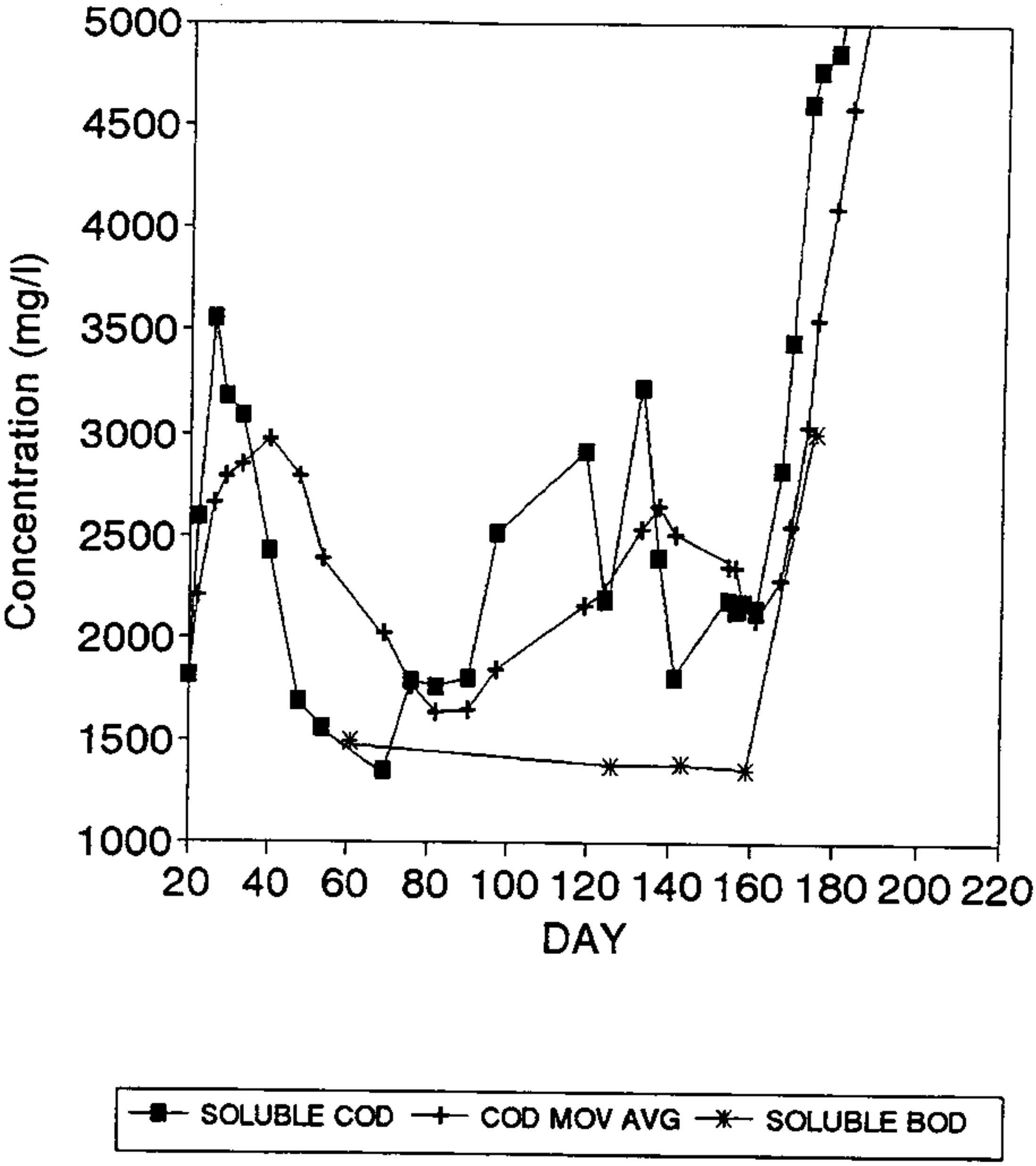


Figure 15. Effluent BOD_5 and COD levels for Reactor A.

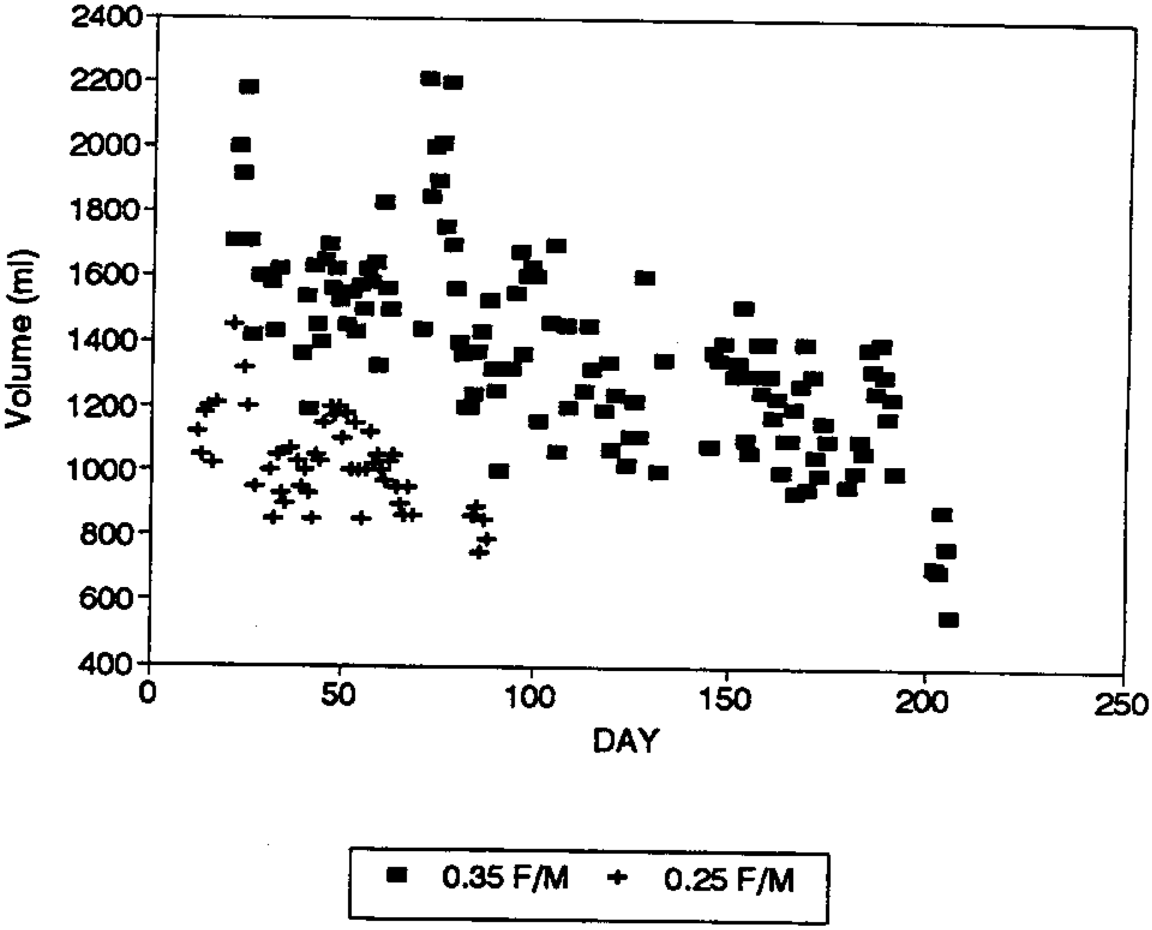


Figure 16. Daily biogas production levels for retort water Reactors A and B.

that remained between 93% and 94%. These removal rates translate to effluent COD levels ranging from 1120 to 1875 mg/L with an average of 1461 mg/L and BOD_5 concentrations that varied from 558 to 603 mg/L with an average of 578 mg/L. Soluble material in reactor B effluent accounted for an average of 87% of all observed COD and BOD_5. As with those found for reactor A, these levels imply that the effluent solids contained a higher portion of inorganic material than the 0.65 mg BOD_5/mg TSS suggested by the ratio found in solids from the secondary effluent of activated sludge systems (Perry et al., 1988). As a result of the digestion process, biogas production, tracked in Figure 16, averaged approximately 1014 ml/day, or approximately twice the predicted value of 517 ml/day (15.427 g/L COD * 0.094 L of feed/day * 91% removal * 0.393 L of gas produced/g of COD digested), during this operational period. Over the course of the previously mentioned time, effluent TSS levels were maintained in the range of 240 to 500 mg/L with an average of 408 mg/L, and the reactor's MLVSS concentration increased from just under 3000 mg/L on day 33 to over 3500 mg/L by day 64. SRT ranged between 215 and 448 days over the course of reactor operation and had an average value of 263.4 days. HRT was maintained at an average of 18.7 days over the life of the reactor. Shortly after day 67 this reactor began to exhibit a decreased effectiveness in the removal of organics. This decrease is characterized by the steady increase in both COD and BOD_5 levels shown in Figure 17. Together with the rise in COD and BOD_5 came a decrease in the reactor's biogas production. This deterioration continued until the reactor was shut down on day 94 of operation.

In an attempt to determine a cause for the failure of these reactors, various operating parameters were studied. At the time of failure, both alkalinity and MLVSS in reactor A were found to be at levels (6560 mg/L and 4740 mg/L, respectively) which were consistent with those found in earlier tests conducted when the reactor had been achieving the previously mentioned removal rates. Similarly, neither alkalinity nor MLVSS levels (5258 mg/L and 3340 mg/L respectively) in reactor B were found to be abnormally low at the time of the reactor's failure. Tests for total effluent phosphorus of both reactors showed concentrations (50 mg/L in A and 31 mg/L in B) which indicate levels sufficient to meet the nutrient needs of each reactor.

Ammonia toxicity has also been cited as a possible cause for disruption in anaerobic digestion processes and was therefore investigated. Ammonia nitrogen concentrations in

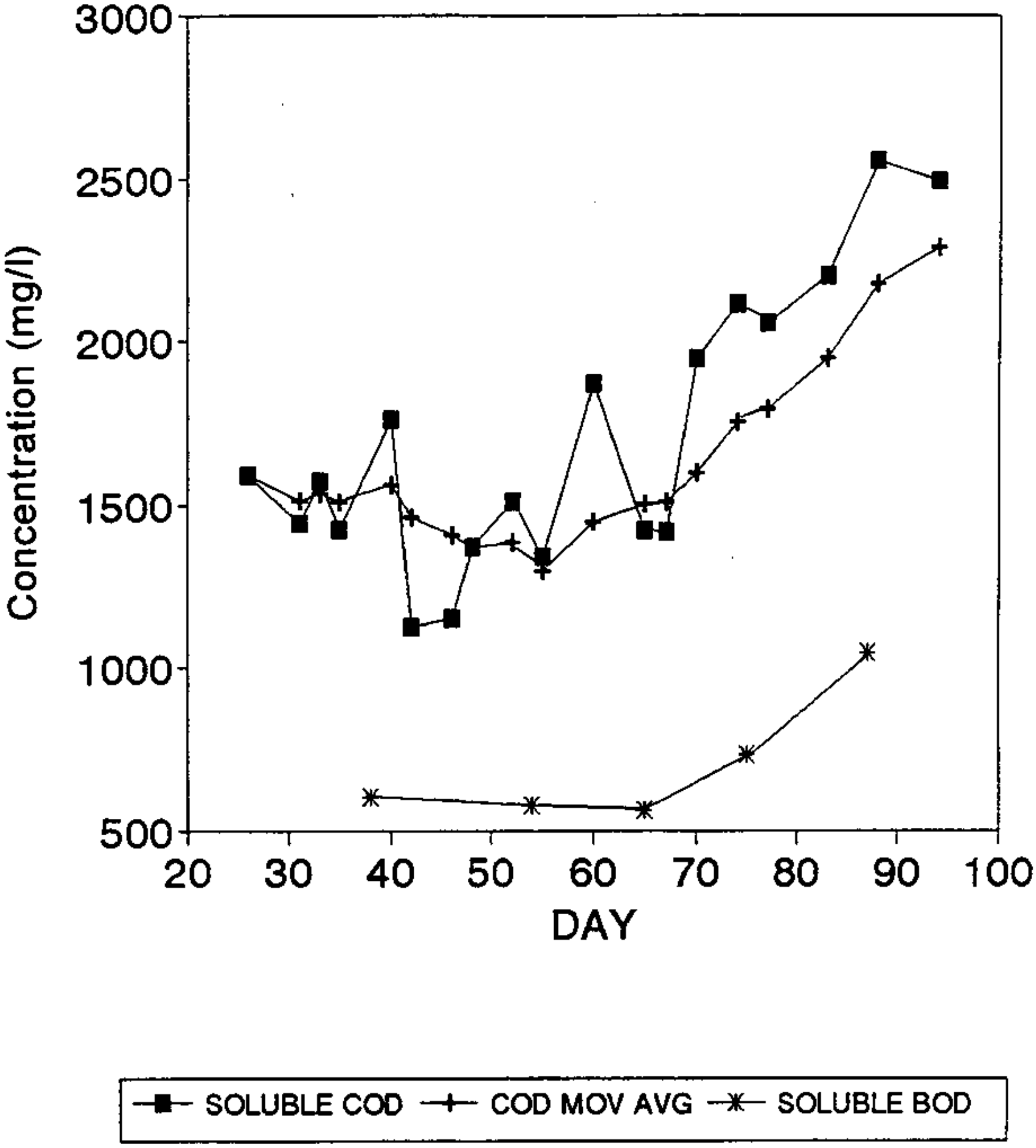


Figure 17. Effluent BOD_5 and COD levels for Reactor B.

reactor A at the time of failure remained in the 1680 to 1780 mg/L range through day 177. While ammonia levels between 1500 and 3000 mg/L have been shown to be inhibitory at pH levels above 7.4 to 7.6 (McCarty, 1964), the fact that the reactor was operating at a pH of 7.50 and that no decrease in pH, usually found during reactor failure (McCarty, 1964), was observed would tend to suggest that ammonia toxicity was not the primary cause of the failure. Figure 18 shows the effluent ammonia and TKN concentrations for reactor A over the course of its operation. While reactor B operated in a higher pH range of 7.62 to 7.75 during failure, the ammonia concentration never reached the inhibitory 1500 mg/l level during operation. Figure 19 tracks the effluent ammonia and TKN concentrations for reactor B over the course of its operation.

Finally, cation toxicity was explored as a possible cause for the failure of the reactors. At the time of failure, sodium, potassium, magnesium, and calcium levels within reactor A had reached 5080 mg/L, 526 mg/L, 224 mg/L, and 162 mg/L, respectively. Tests on effluent taken from reactor B during failure also revealed these four cations to be present in concentrations of 4175 mg/L, 495 mg/L, 210 mg/L, and 177 mg/L, respectively. In each case, sodium is present in levels which have been found to be moderately inhibitory (McCarty, 1964). Inhibition observed in these reactors may have been mitigated through the presence of cations (potassium, calcium, and magnesium) which are often seen as antagonists in situations of sodium toxicity (Kugelman and Chin, 1971). Even with the antagonistic effects of these additional cations, high sodium levels together with the high ammonia concentrations present are believed to be the cause for the unstable manner in which these reactors operated and the eventual cause of their failure. Data on the cation concentrations found in reactor B are presented in Figure 20. Further study of the toxicity effects of high concentrations of free ammonia, when found in conjunction with high light metal cation concentrations, is warranted.

The last two reactors (C and D) were operated at F/M ratios of 0.10 and 0.07, respectively. These two reactors were both fed a wastewater mix representative of the processing facility's overall wastewater output and were operated for periods of 287 and 94 days, respectively. As with the retort fed reactors described previously, these reactors showed signs of failure, and were subsequently shut down, after having earlier exhibited reductions in COD and BOD_5 levels of the waste.

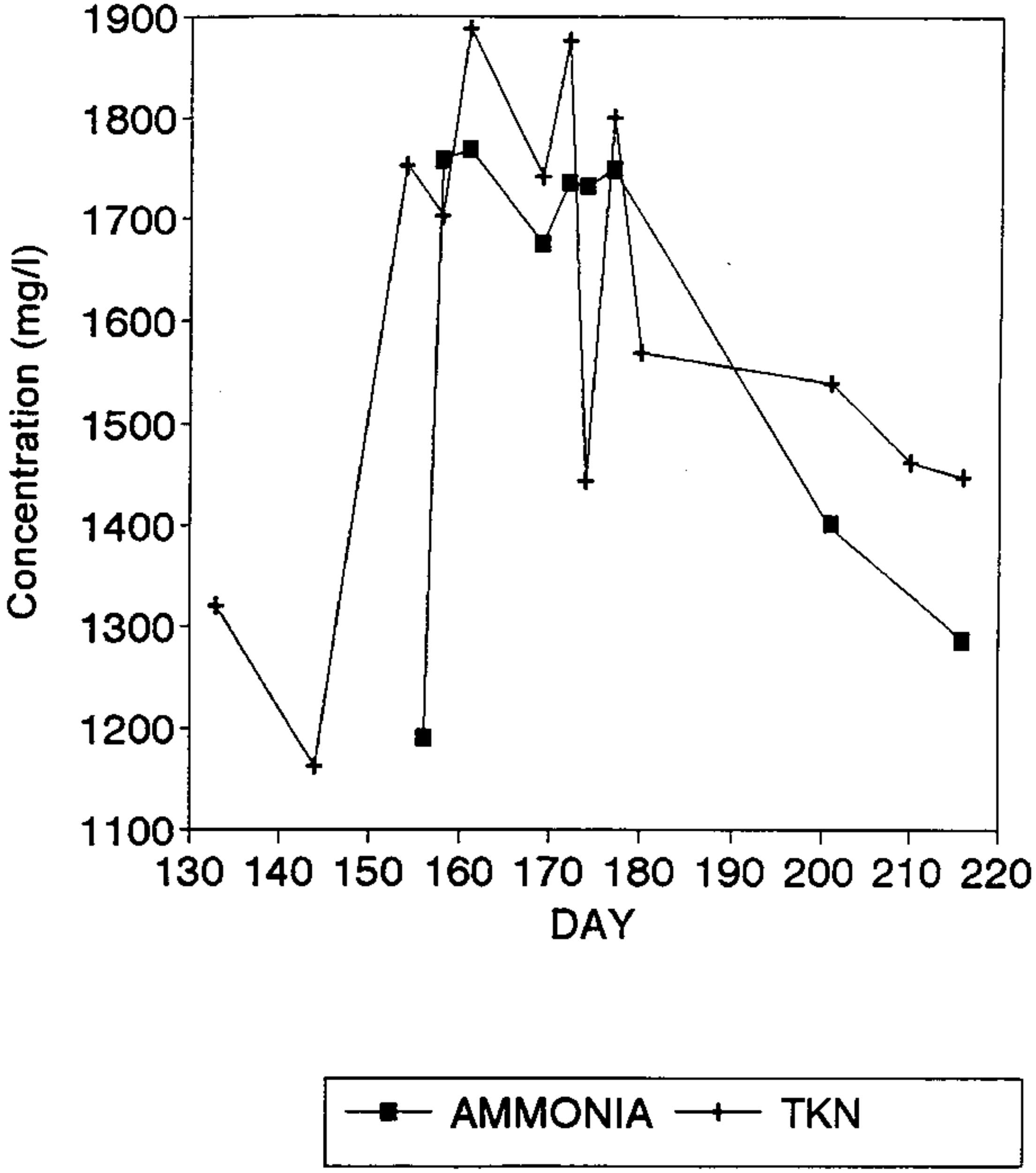


Figure 18. Effluent ammonia and TKN levels for Reactor A.

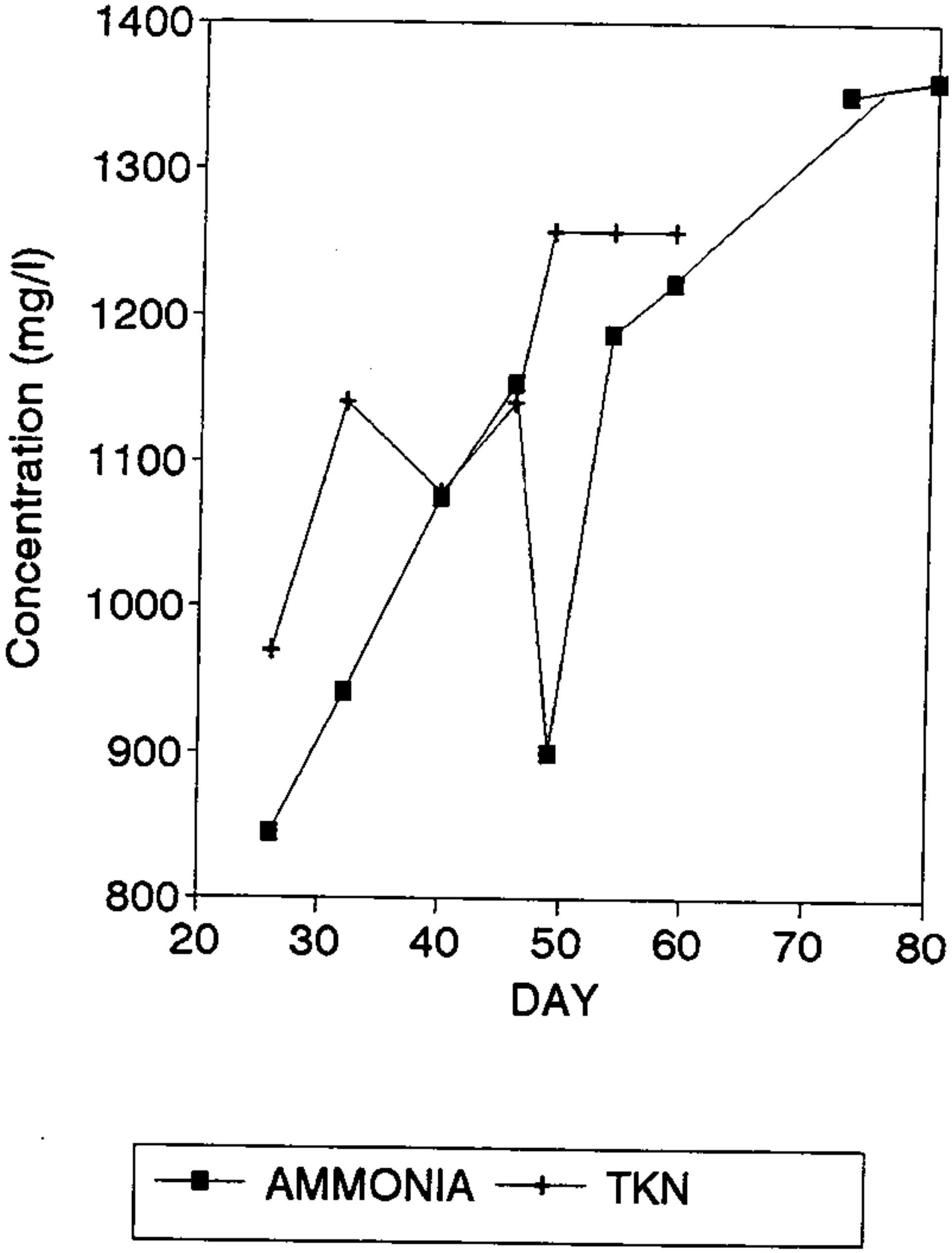


Figure 19. Effluent ammonia and TKN levels for Reactor B.

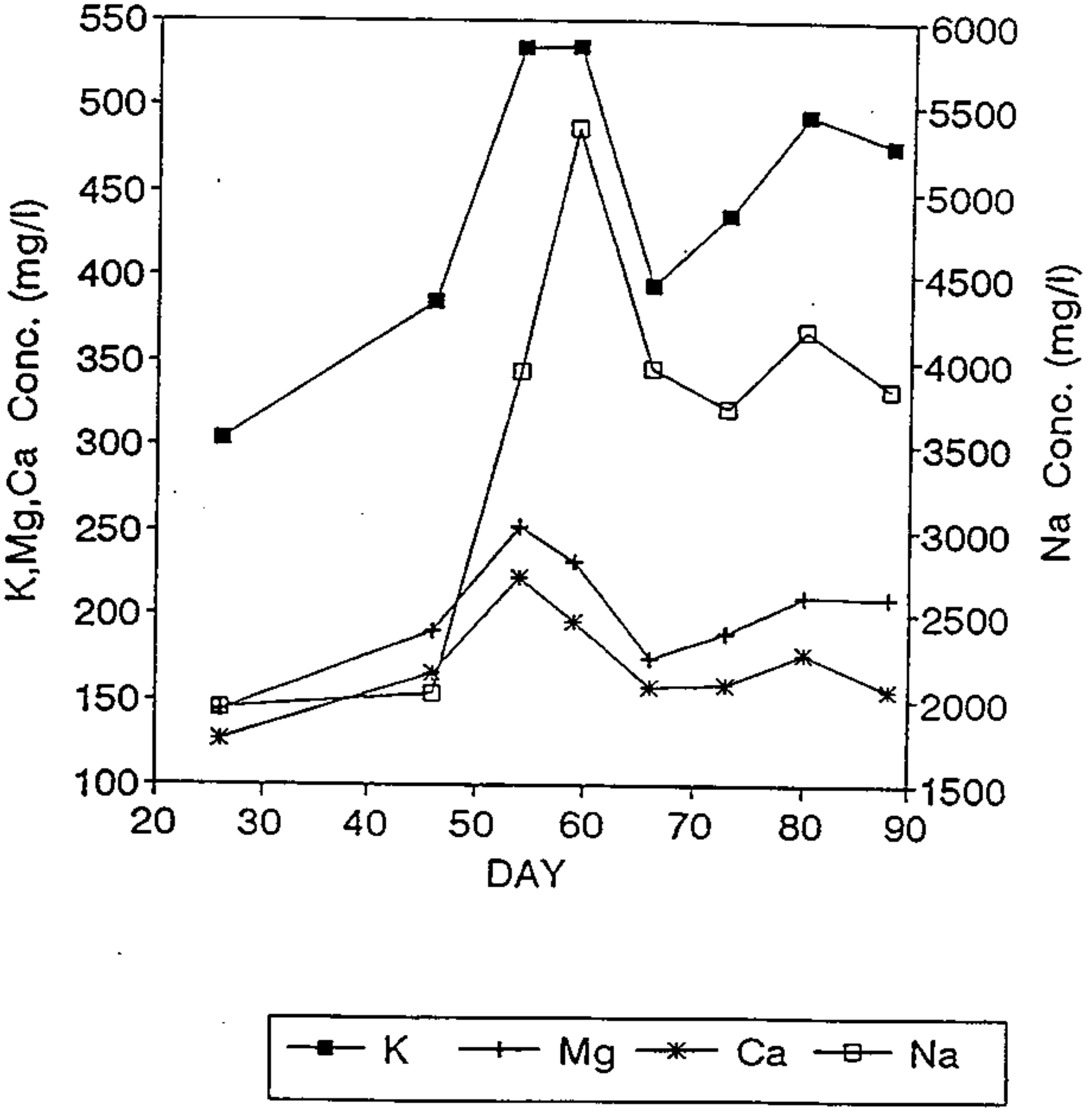


Figure 20. Cation concentrations for Reactor B.

For the period of time from day 22 through day 247, effluent soluble COD levels in reactor C remained between 1250 and 3310 mg/L with an average effluent COD of 2290 mg/L. These levels correspond to removal rates ranging from 55% to 83% with an average of 69%. This reactor also exhibited BOD_5 removal rates, over the same period, varying between 72% and 84% with an average of 79%. These rates translate to effluent soluble BOD_5 concentrations between 893 and 1596 mg/L with an average of 1180 mg/L. As a result of the organics removal achieved by the reactor over this period of time, biogas was produced at an average rate of 254 ml/day. This rate is 2 times the 127 ml/day rate predicted by observed average COD percent removal, and daily feed levels and strength (7.405 g/L COD * 0.063 L of feed/day * 69% removal * 0.393 L of gas produced/g of COD digested). As with the retort-fed reactors, excess gas production can be attributed, in large part, to the fact that total biogas production from anaerobic digestion can include byproducts such as carbon dioxide and hydrogen sulfide in addition to methane. Also, over this period of time it was observed that MLVSS of the reactor declined from 3240 mg/l on day 28 of operation to 1340 mg/L by day 235. Effluent TSS levels for this reactor ranged from 700 to 2180 mg/L over this same time with an average of 1577 mg/L. Solids retention time varied between 41 and 183 days over the life of the reactor with an average value of 98.6 days. HRT was maintained at an average of 36.3 days over the life of the reactor. The period from day 248 through day 276 was marked by an increase in average COD concentration to 2730 mg/L, and was followed by substantial increases in both soluble COD and BOD_5 levels. As a result, the reactor was shut down on day 287 after effluent COD had reached 4100 mg/L, and biogas production had fallen to less than 100 ml/day. Soluble material accounted for an average of 85% of all COD and 94% of all BOD_5 observed in reactor C effluent. As with those found for the retort water reactors, these levels imply that the effluent solids contained a higher portion of inorganic material than that suggested by the ratio found in solids from the secondary effluent of activated sludge systems (0.65 mg BOD_5/mg TSS; Peavy et al., 1985). The COD and BOD_5 concentrations over the life of the reactor are presented in Figure 21, while biogas production is tracked in Figure 22.

Reactor D operated with effluent soluble COD levels between 985 and 1445 mg/L from day 26 through day 46 with an average COD of 1281 mg/L. This translates to COD removal rates ranging from 79% to 86% with an average of 84%. Over this period, the

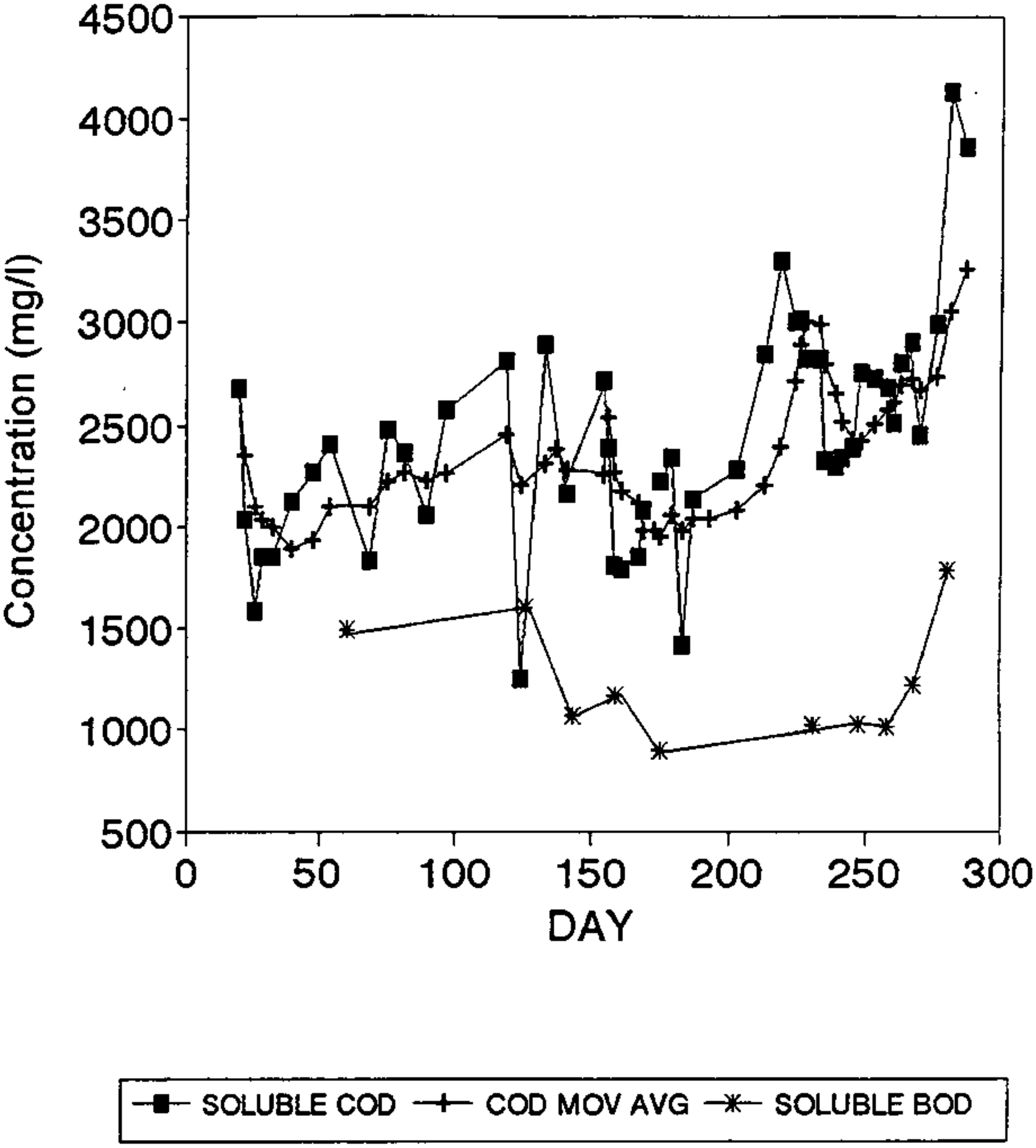


Figure 21. Effluent BOD_5 and COD levels for Reactor C.

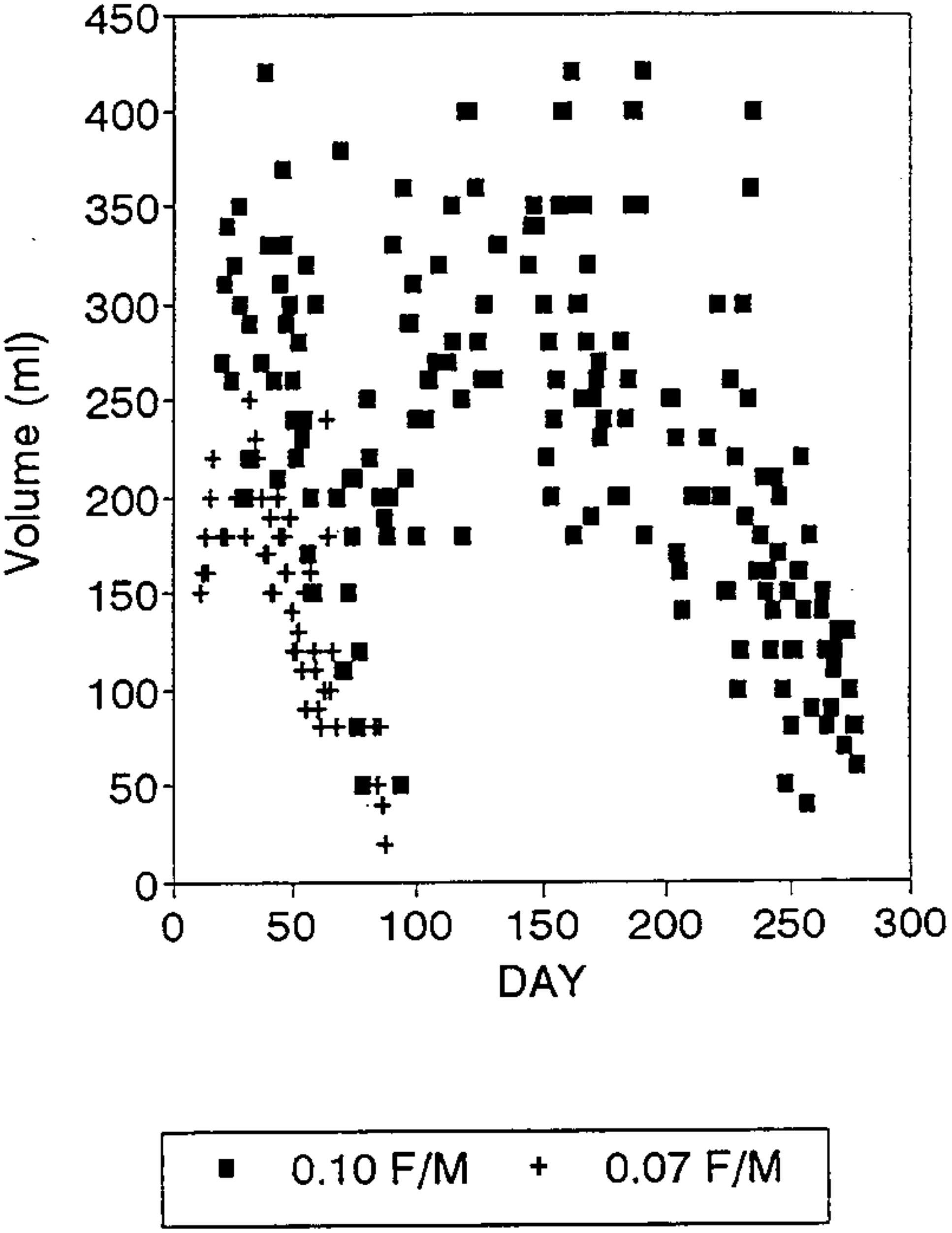


Figure 22. Biogas production levels for waste Reactors C and D.

biogas production from the reactor proceeded at the rate of 195 ml/day, but this rate decreased during failure to a level of less than 100 ml/day. The predicted rate of biogas production for this reactor was 155 ml/day (8.080 g/L COD * 0.058 L of feed/day * 84% removal * 0.393 L of gas produced/g of COD digested). During the observed reactor failure, effluent COD levels steadily increased after day 48 to over 4200 mg/L by shut down on day 94. Soluble BOD_5 levels remained between 833 and 891 mg/L from day 26 through day 65 with an average of 870 mg/L for a BOD_5 removal rate of 83%. After day 65, BOD_5 concentrations also began a rise, which continued until reactor shut down. Soluble material accounted for an average of 84% of all COD and 99% of all BOD_5 observed in reactor D effluent over the course of reactor operation. As with the preceding reactors, these levels imply that the effluent solids contained a higher portion of inorganic material than what could be expected given the ratio found in solids from the secondary effluent of activated sludge systems. Effluent TSS increased over the course of the reactor operation from 240 mg/L on day 26 to 2320 mg/L on day 83, and while MLVSS remained between 2780 and 2960 mg/L from day 26 to day 46, their levels fell to 2460 mg/L by day 85. After an initial reading of 686 days, solids retention time for the reactor decreased rapidly until it stabilized in the 102 - 112 day range. The average value of the SRT over the life of the reactor was 138.6 days. HRT was maintained at an average of 32.4 days over the life of the reactor. COD and BOD_5 levels over the life of reactor D are tracked and shown in Figure 23 while daily biogas production is shown in Figure 22.

As with the retort-fed reactors, an attempt was made to discover the cause of failure in these two reactors. Alkalinity and pH were not seen to be problems for either reactor, as each was found to be operating during failure at levels 3828 mg/L in a pH range of 7.50 to 7.65 for reactor C, and 3000 mg/L in a pH range of 7.45 to 7.55 for reactor D, consistent with those observed during the aforementioned period of organics removal. Loss of MLVSS was, as mentioned earlier, a problem for both mixed wastewater reactors; however, since failure of each occurred during periods of stability in this parameter and not during periods of decline, this loss is not seen as a primary cause of reactor failure in either reactor C or D. Effluent phosphorus concentrations found during operation and failure in each reactor (70 to 120 mg/L for reactor C and 55 to 120 mg/L for reactor D) indicate that it was present in sufficient quantities as not to be a limiting factor for either reactor.

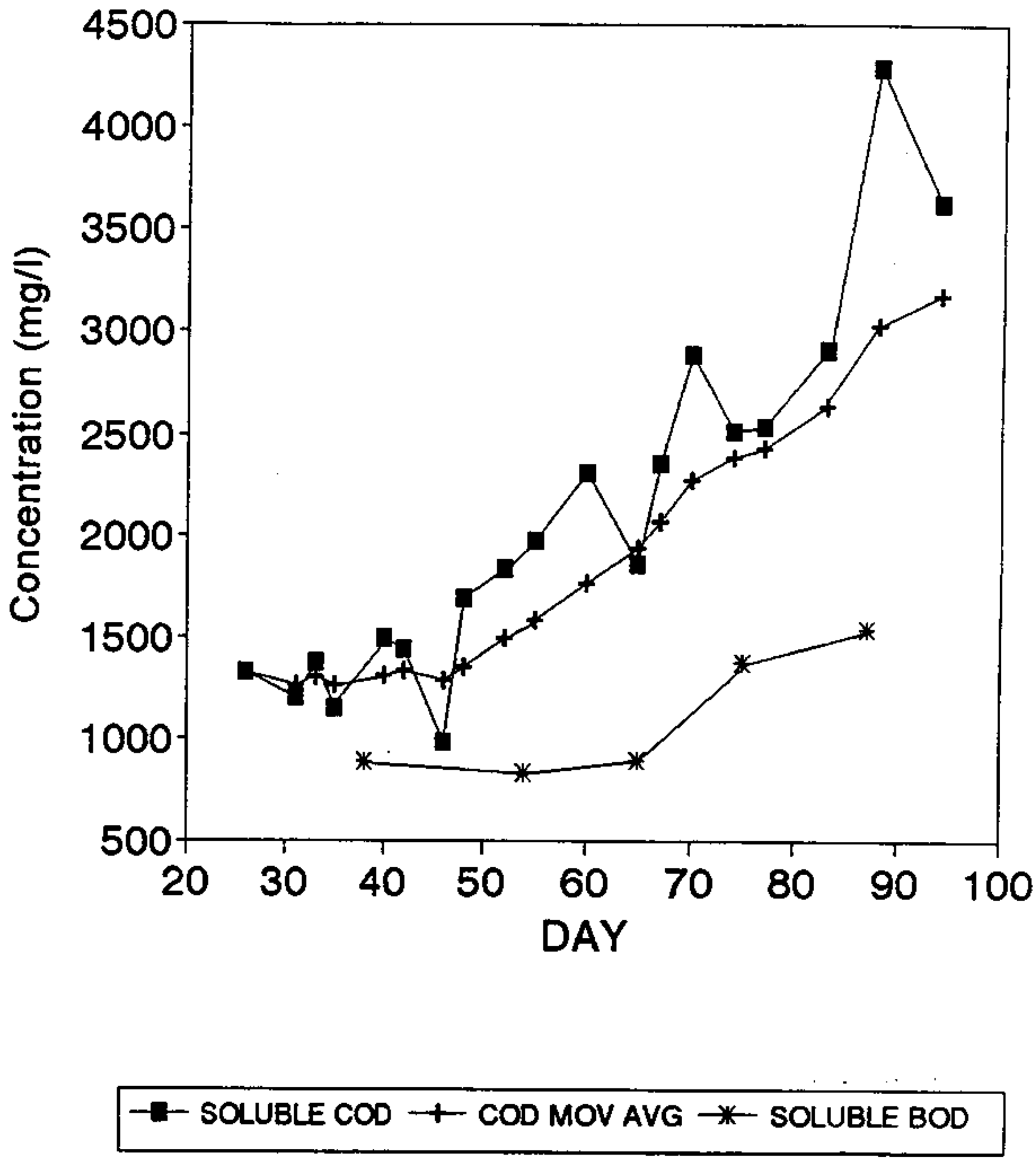


Figure 23. Effluent BOD_5 and COD levels for Reactor D.

Ammonia levels in the reactors, ranging from 748 to 949 mg/L in reactor C and from 592 to 787 mg/L in reactor D, were not seen to be problems for either reactor as they never reached the previously mentioned inhibitory range of 1500 to 3000 mg/L. Ammonia data for reactors C and D are presented in Figures 24 and 25, respectively, together with the TKN data for each.

Cation toxicity, and specifically sodium toxicity, is seen as the primary cause of failure in both reactors. As sodium has been found to be strongly inhibitory at a concentration of 8000 mg/L (Kugelman and McCarty, 1965; McCarty and McKinney, 1961), the fact that the sodium level in each reactor far exceeded this amount clearly suggests sodium toxicity. The question of how these reactors performed as long as they did, albeit at a somewhat lower than hoped-for removal efficiency, given this sodium toxicity, can be answered via the phenomena of cation toxicity antagonism. Potassium concentrations found in both reactor C and D were consistent with those needed to act as an antagonist for sodium toxicity (Kugelman and Chin, 1971). Additionally, the presence of significant concentrations of calcium and magnesium, which are often found to be stimulatory in the presence of another antagonist (McCarty, 1964), suggests that a multiple antagonistic effect (Kugelman and Chin, 1971; Kugelman and McCarty, 1965) may have been responsible for the extended performance exhibited while in a state of sodium toxicity. Concentrations of the four cations, sodium, potassium, magnesium, and calcium are presented in Figure 26 for reactor C and Figure 27 for reactor D.

While it is believed that sodium toxicity was the major cause of the failure of the reactors, the ability of reactors A and C to perform for a greater period of time than their less heavily loaded counterparts, reactors B and D, suggests that there may have been an additional factor contributing to reactor failure. The fact that reactors A and C were operated at lower F/M ratios, 0.15 for reaction A and 0.05 for reactor C, for a period of 55 days prior to this experiment (Harrison et al., 1992) opens the possibility that the sludge was able to become acclimated to the high sodium concentration at these lower loadings.

An additional area which was explored was the possibility of a correlation between the ratio of either the ammonia or sodium concentrations to the biomass level during the failure of the reactors. On inspection of the data obtained from these experiments, no such correlations were found.

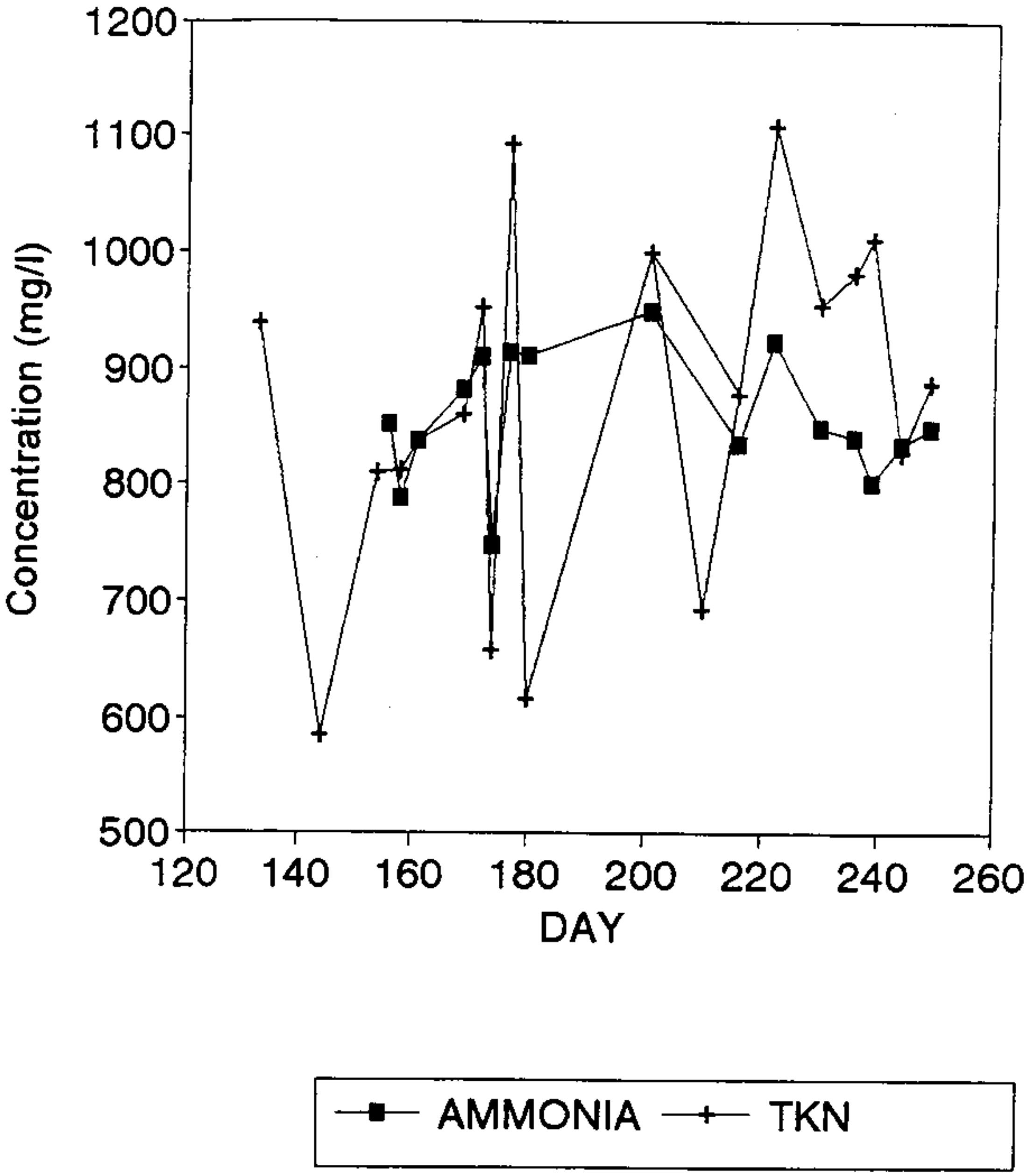


Figure 24. Effluent ammonia and TKN levels for Reactor C.

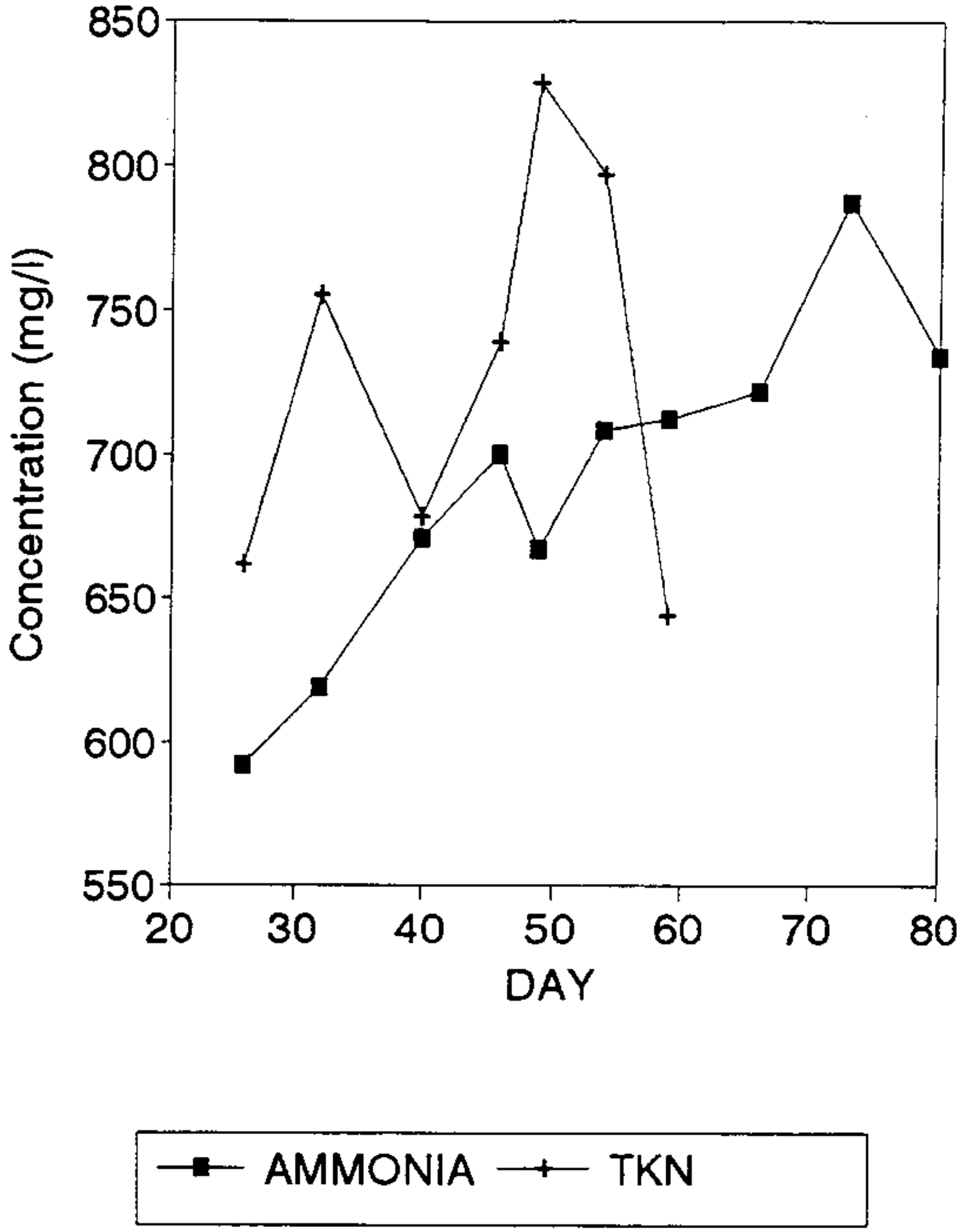


Figure 25. Effluent ammonia and TKN levels for Reactor D.

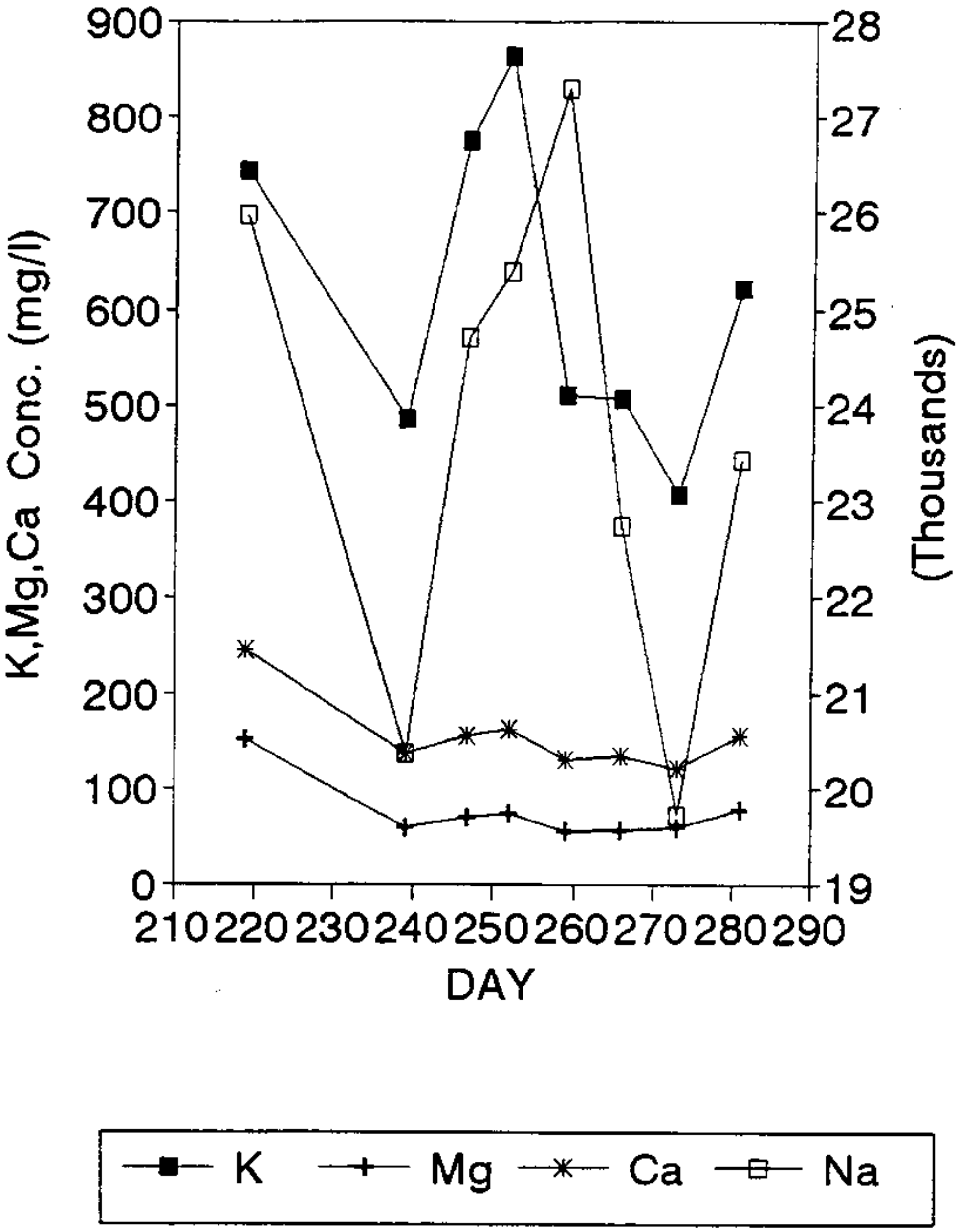


Figure 26. Cation concentrations for Reactor C.

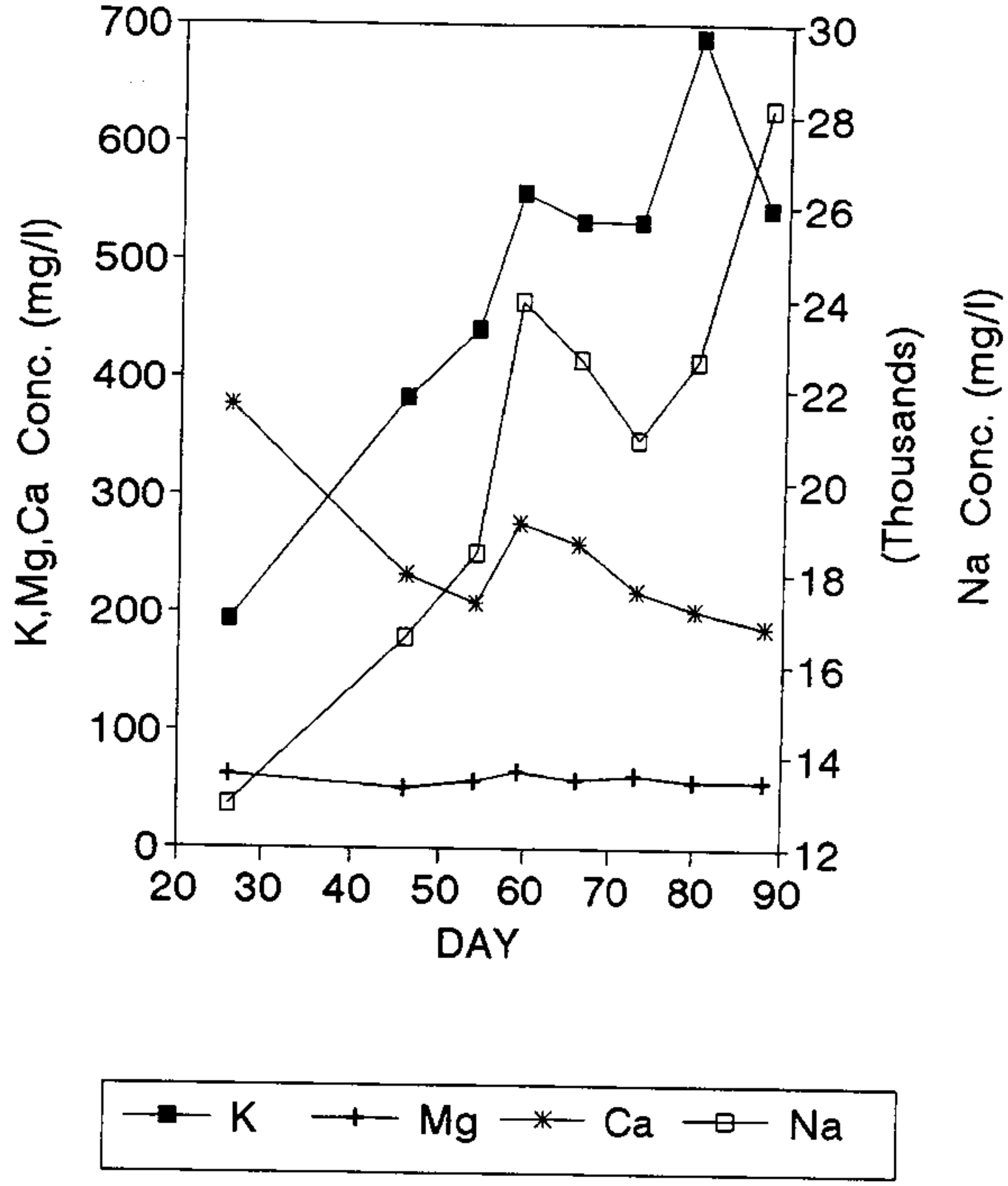


Figure 27. Cation concentrations for Reactor D.

Since three of the four reactors failed at a time when crabs were being taken by the process of bottom dredging, the possibility arises that a change in the composition of the processing wastewater may have contributed to reactor failure. Additionally, the possibility that failure was caused by a deficiency in some trace nutrient (e.g. cobalt, nickel, iron) can not be completely discounted, and any further studies should address these needs.

Cation toxicity problems in any treatment system for either of these wastewaters must be addressed in order for such a system to operate successfully. Such provisions may include an elutriation step where the reactor sludge is periodically "washed" with fresh water which has been purged of dissolved oxygen. After a settling period, this water would then be flushed out of the system in the same manner in which reactor effluent is removed, thereby removing excess sodium. The reactor would then be refilled with fresh, purged water and treatment could continue. Alternatively, a periodic dumping and restarting of the reactor with new biomass could be employed when toxicity effects are observed; however, this method may lead to excessive start-up times. It should be noted that this periodic dumping of the reactor contents would still be subject to the NPDES permit limits and thus would most likely require access to a public sewer.

Based on the results of this experiment, it is suggested that processing facilities which fall under the "existing" classification, which does not have BOD_5 discharge limits, do not employ an anaerobic system for the treatment of their wastewaters unless they are prevented, in some other manner, from discharging directly to the environment. In this case anaerobic treatment could be used as a pretreatment for disposal to the sewers. This recommendation is due to the high effluent ammonia concentrations (592 - 1780 mg/L) resulting from the anaerobic deaminization of proteins. In light of the projected ammonia discharge limits, these levels would make direct discharge of wastewaters very difficult even if the waters were subsequently treated for ammonia removal, and would eliminate the possibility of direct discharge without such ammonia removal treatment.

In the case of facilities which fall under the "new" facilities classification, and are subsequently subject to BOD_5 discharge limits, the data from these experiments suggest that anaerobic treatment would be an effective tool for the treatment of wastewaters prior to discharge to a sewer system. It is further suggested that, if only retort process waters are to be anaerobically treated, this system should be operated at an F/M ratio of 0.35 or less (with improved removal rates to be balanced against larger required reactor volumes at lower F/M ratios). It is also recommended that the anaerobic system include a settling chamber to increase the solids retention time of the reactor, and thereby improve the reactor's performance. A reactor of the type stated could be expected to provide approximately 88% removal of soluble BOD_5, and result in an effluent soluble BOD_5 concentration of approximately 1400 mg/L.

If it is decided that the total plant wastewater stream, such as the mixed wastewater studied here, is to be anaerobically treated, the above system should then be operated at an F/M ratio between 0.07 and 0.10. In this treatment scheme, soluble BOD_5 removal rates of between 79% and 83% can be expected, resulting in effluent concentrations between 870 and 1180 mg/L.

As stated previously, any anaerobic treatment system used to treat these processing waters must be monitored for cation and ammonia toxicity problems. Additionally, provisions, such as the elutriation step mentioned previously, should be made to counteract any such toxicity problems that arise.

In cases where a processing facility's discharges are subject to BOD_5 limits, and where the facility does not have access to a public sewer, it is suggested that the anaerobic digestion process studied here be followed by an aerobic polishing step, and that this be followed by provisions for additional ammonia removal. Due to the low capitalization generally found among crab processors, this three part system may prove to be too costly for much of the industry. This fact may mean that some processors which are currently operating in areas without public sewer access could be forced from business. In any case, the new regulations will mean increased costs throughout the industry.

4.2.5 Air Stripping Experiment

The main thrust of the air stripping experiment was to determine the applicability of this technology as an effective compact alternative for the removal of ammonia from the retort process wastewater. Another area of concern of this experiment was an examination of the cooling benefits derived by passing the retort process effluent through the air stripping tower. The cooling of the retort effluent is important due to the possible inclusion of temperature limits in NPDES permits.

Although the retort processing and sample collection methods remained the same throughout the testing period, a great variability was found in the ammonia levels of the retort effluent; i.e., the tower influent. With the exception of one test, which had an ammonia level of about 50 mg/L, the range of tower influent ammonia concentrations was approximately 80 to 160 mg/L. These levels are consistent with those found in previous research involving this waste (Harrison et al., 1992). As a result of this variability, percent reduction in the ammonia levels was the main criteria used in the comparison of tower performance at the various operating parameter combinations.

In the testing of the ammonia concentrations of the retort process effluent, it was found that there was no significant difference in the levels found in the pH adjusted tower influent as compared to those found in the unadjusted retort process effluent. For this reason, it is believed that the ammonia removal achieved in these tests can be attributed to the air stripping process alone.

Figure 28 shows the impact of varying the influent liquid flow rate, and thereby the air to water ratio, for the tower on the treatment efficiency of the air stripping process. The graph shows that at pH 11, decreasing the liquid flow-rate from 2.2 GPM to 1.2 GPM led to an approximate doubling in removal efficiency, but a further decrease in flow-rate to 0.6 GPM did not provide any increase in ammonia removal. In tests performed at a pH of 10, however, a 7% increase in treatment was observed when the flow-rate was decreased from 1.2 GPM to 0.6 GPM. All tests were performed at average influent temperatures of approximately 56°C to 58°C. As a result of these findings, it is suggested that while the lower flow-rate of 0.6 GPM may not improve treatment efficiency at higher pH levels (11 and above), the higher air-to-water ratio may offset, to some extent, the deleterious effects of the lower pH levels (10 and below). This may allow a choice of treatment schemes which would permit chemical costs to be balanced against the problems associated with lower processing rates.

Figure 29 displays the effect of temperature variation on ammonia removal effectiveness at pH 11.0 and pH 12.2. From this data it can be seen that an increase in influent temperature can be expected to bring about some increase in treatment efficiency. This would suggest that any system of this type be operated in such a manner that the retort effluent be processed through the tower with the minimum possible cooling time allowed.

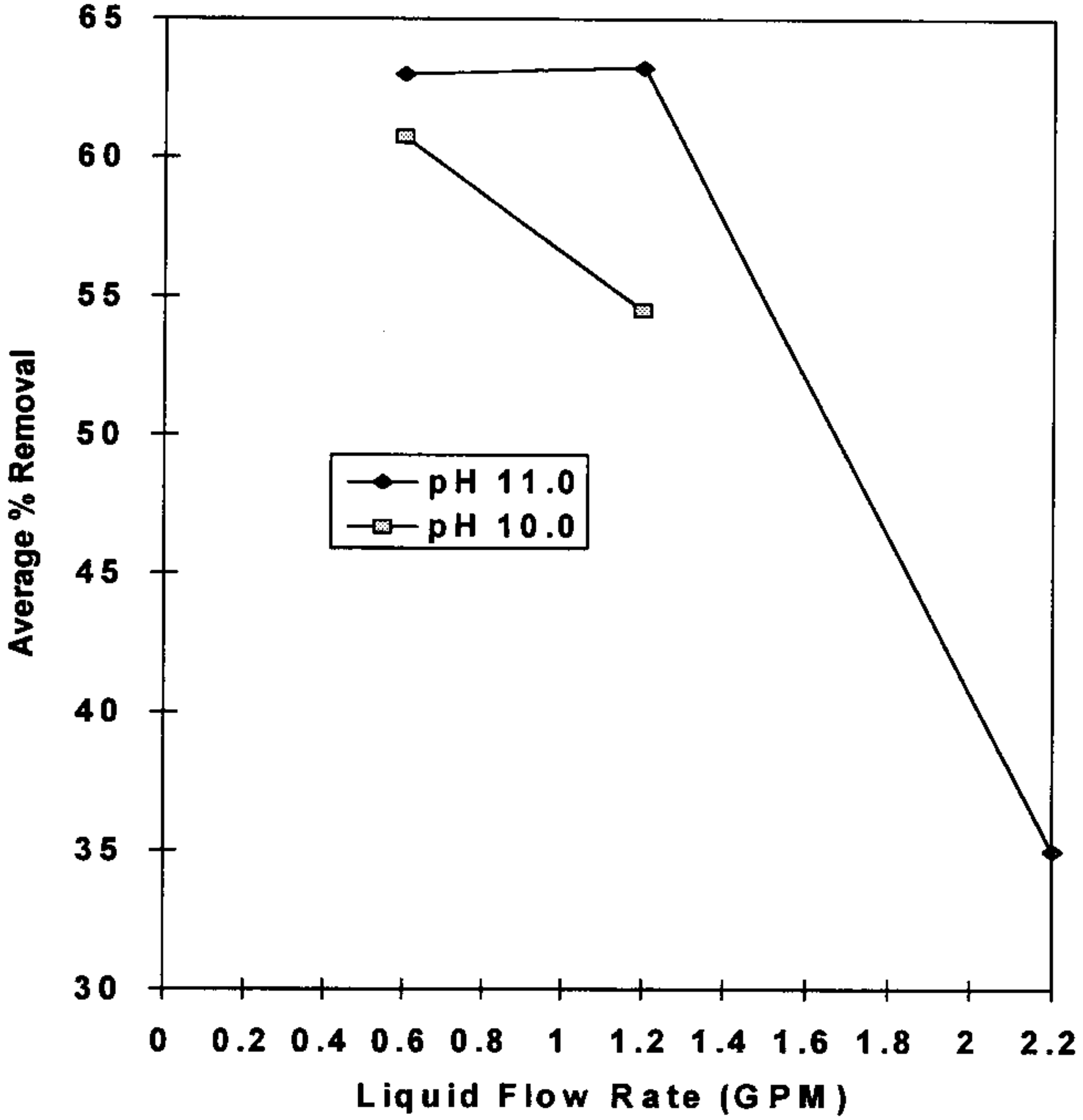
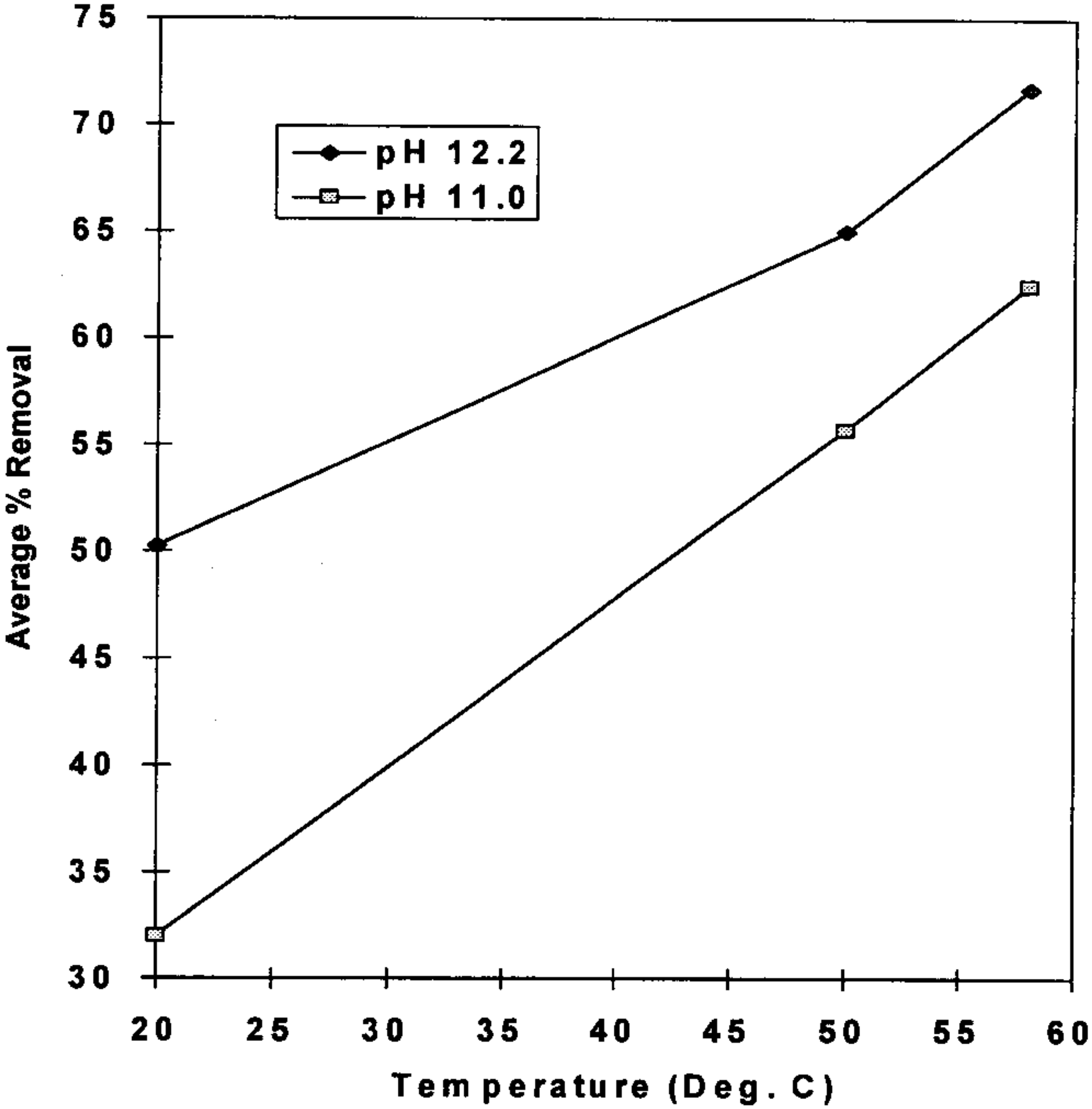


Figure 28. Air stripping ammonia removal rates as a function of liquid flow-rate.



Firgure 29. Air stripping ammonia removal rates as a function of influent temperature.

One cautionary point that must be stated is the fact that the LANPAC® packing material used in the study is not recommended for use at liquid temperatures above 71°C (Joyce, 1993).

Figure 30 compares the treatment achieved by the tower as the pH of the influent stream was varied. It shows a small improvement (from 61% to 63%) in removal efficiency as the influent pH was raised from 10 to 11, while temperature and air-to-water ratio were held constant. A further test conducted at pH 12.2 and the same temperature and air-to-water ratio showed an increase to a 71% removal efficiency.

Finally, tests performed to establish the added treatment achieved by recycling the tower effluent back through the system showed 32% treatment for the tower effluent recycled without mixing and 51% removal of the ammonia in the mix of tower and retort effluent. These figures translate to effluent ammonia concentrations of 23.8 mg/L and 30.4 mg/L, respectively. These levels of treatment were achieved even though the treated liquid entering the tower had an average temperature of 20°C. Ammonia removal efficiencies achieved for all air stripping experiments are provided in Table 16.

An additional benefit of the air stripping process was found in the cooling effect that it imparts upon the wastewater. In the tests conducted, while the influent average temperatures ranged from 50°C to 65°C, the tower effluent was cooled to essentially the ambient air temperature (13°C - 20°C). The majority of this reduction is attributed to the cooling effects of the tower because influent temperature changed very little (0°C-4°C) over the course of each test.

With the exception of the trial performed at an air-to-water ratio of 225 ft^3/gal, the air stripping tower produced an effluent with ammonia concentrations between 19.5 mg/L and 58.3 mg/L. These residuals translate to ammonia removals between 30.1 and 102.0 mg/L for the retort effluent, and between 11.1 and 30.9 mg/L for the recycled tower effluent.

While these tests showed significant reductions in wastewater ammonia levels as a result of the air stripping process, the reductions were not as great as those stated in literature sources (Kuhn, 1956; Slechta and Culp, 1967; Culp and Culp, 1971). While this reduced efficiency could be attributed to the higher ammonia concentrations dealt with in this study, it is more likely due to a shortcoming in one or more aspects of the tower design employed here.

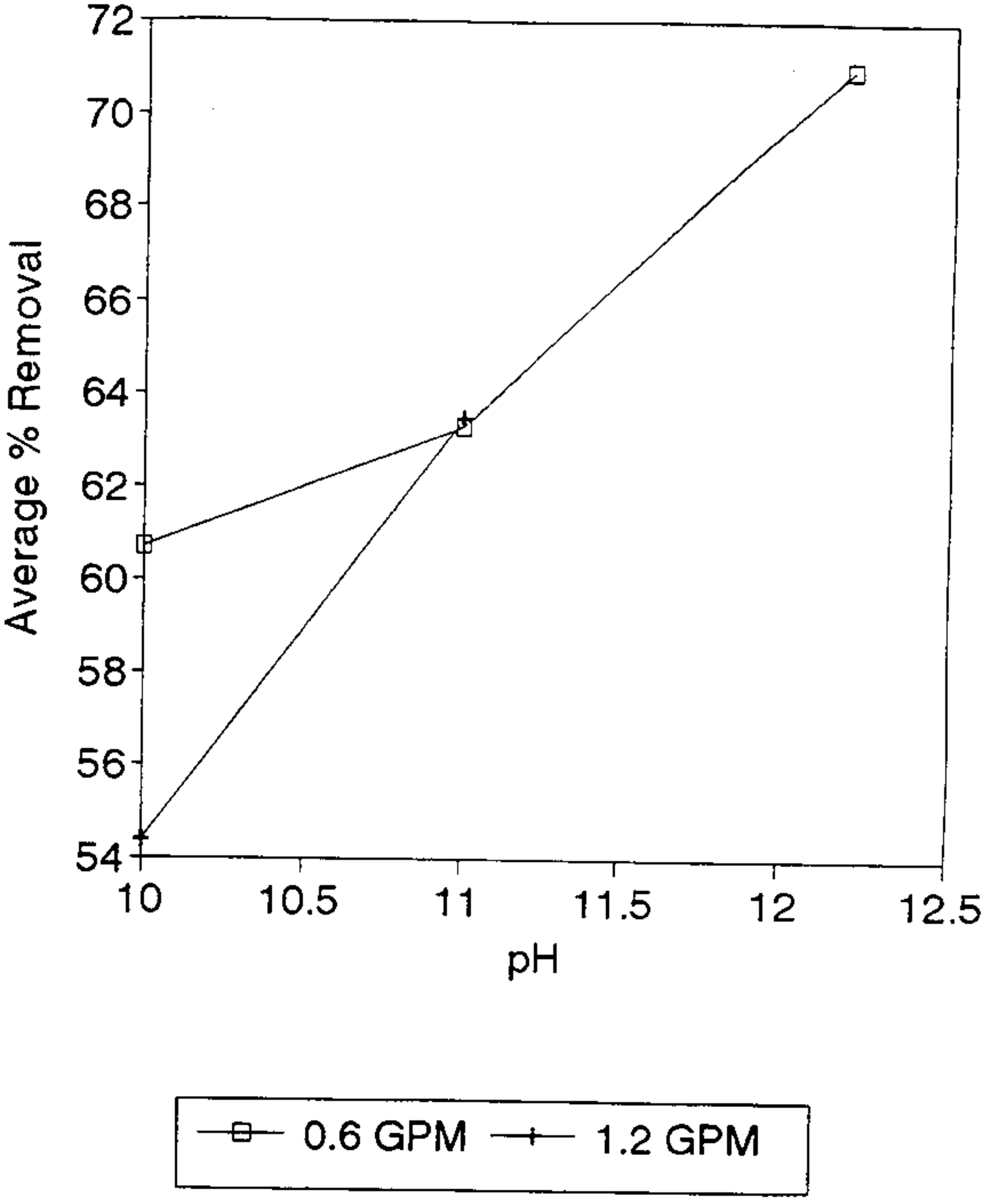


Figure 30. Air stripping ammonia removal rates as a function of influent pH level.

Table 16. Air stripping tower ammonia removal efficiencies.

Order of Trials	pH	Air-To-Water Ratio ft^3/gal	Inf. Temp °C	Eff. Temp °C	Range % Removal[1]	Average % Removal
3	10.0	412	59	16	50 - 59	54
10	10.0	825	58	20	58 - 62	61
1	11.0	225	56	15	25 - 45	33
2	11.0	412	58	15	60 - 67	64
8	11.0	825	58	20	57 - 67	63
4	11.0	412	65	16	64 - 69	67
9	11.0	825	20	20	13 - 45	32
6	12.2	825	20	13	43 - 58	50
7	12.2	825	50	13	60 - 70	64
5	12.2	825	58	15	67 - 73	71

1 = All trials used 5 samples taken at 5 min. intervals with the exception of trials 6 and 9, which used 4 samples.

Additionally, it is believed that a liquid distribution system that is capable of introducing the wastewater into the tower in droplets could improve system performance. Spraying the waste over the top of the tower would increase the surface area of water that would come in contact with the air at a point before the benefits of the high waste temperature have been lost.

Also, while the blower system utilized in these tests was able to maintain a sufficiently high average volumetric air flow, it was observed that the air flow rate varied greatly across the diameter of the tower. This may have caused a certain portion of the waste to channel through "dead zones" where a lower gas to liquid loading ratio may have resulted in less than optimal treatment. It is believed that the use of an in-line blower system, in which a fan, located at the top of the tower, would draw air up from the bottom of the tower, would remedy this problem and lead to higher removal rates.

Another possible design modification which could be employed in order to improve removal efficiencies is an increase in the effective tower packing depth.
This could be accomplished through the use of either one tower of a greater packing depth or two towers operated in series.

A concern which must be addressed before the implementation of any ammonia stripping process is the potential for air pollution problems. The odor threshold for ammonia is approximately 35 mg/m^3 (Culp, 1978). Care should be taken to ensure that levels present in the stripper discharge do not exceed this concentration. At an air-to-water ratio of 500 ft^3/gal (3.7 m^3/L), up to 129 mg/L can be stripped without exceeding this threshold. At the 825 ft^3/gal (5.9 m^3/L) ratio tested here, it would be possible to achieve removals of up to 207 mg/L of ammonia before the odor threshold became a problem. Removals of concentrations greater than these would most likely require the capture and treatment of the tower off-gas. Odor was not a problem in any of the trials performed here.

Concern about the redeposition of the ammonia, after its removal in gaseous state, into the adjacent receiving waters may also result in the need to capture and scrub the air stripper's off-gas. It is believed, however, that these concerns may be unnecessary. While it is true that the total amount of ammonia eventually deposited in the receiving water after air stripping will not be substantially less than the amount deposited if the wastewater were not treated prior to its discharge, the concentrations in which this ammonia will be introduced will be greatly reduced (i.e. 160 mg/L for the untreated wastewater versus approximately 0.03 mg/L for the treated wastewater).

The projected effluent limits of 1 mg/L and 5 mg/L total ammonia suggest that ammonia would prevent the direct discharge of retort process effluent. Although modifications to the design tested here would be needed, the data obtained in these experiments suggest that a treatment process involving air stripping technology could be used to reduce retort effluent ammonia concentrations to a level where direct discharge would be possible.

This system would most likely involve two towers operated in series. The first of these towers, where the bulk of the ammonia removal would occur, would need to be operated at an air-to-water ratio greater than 600 ft^3/gal in order to ensure sufficient ammonia removal, while minimizing the potential for air pollution problems. The exact design of these towers would have to be performed on a case by case basis, taking into account wastewater volume and projected operational air and water temperatures. In addition, it is believed that a three stage air stripping process may be useful in reducing the ammonia levels found in the effluent waters from biological treatment systems, used for the removal of organics, to a level where these waters may be directly discharged. In the event that direct discharge is not possible, an air stripping process could still be used as a pretreatment to reduce sewer charges.

4.3 Potential Treatment System Costs

Any treatment scheme employed in the disposal of crab processing industry wastewaters will impart additional capital and operational costs upon the industry. In the cases of plants with access to public sewer systems, the costs of discharge to public sewers without pretreatment must be compared to the costs of treating the wastewater on site for subsequent disposal. This disposal may take the form of either direct discharge to the environment, if sufficient treatment is accomplished, or discharge of a pretreated, lower strength waste, to public sewers. In order to gain some insight into the potential impact of the aforementioned alternatives on the industry's profitability, a preliminary investigation into the costs of each was performed.

For a hypothetical plant operating five days a week with fifteen cookings per day and employing a Harris claw processing setup, it is predicted that the disposal of untreated processing wastewaters in public sewers would incur surcharges, due to excessive BOD_5, TSS, phosphorous, and TKN-N levels, of approximately $28,000 per year. This value is based on rates for the Hampton Roads Sanitation District. Through the implementation of a pretreatment scheme employing an anaerobic contact system of the type tested here, it is estimated that these annual surcharges could be reduced to less than $10,000. In the event that the wastewaters could be treated to a level where direct discharge to the environment was possible these sewer surcharges would of course be eliminated. These savings would have to be balanced against the costs of implementing and operating any treatment system.

Due to the relatively small volume dealt with in this situation (approximately 5300 gal/day of total plant flow, or a projected reactor volume of 180,000 gallons), it is difficult to obtain reliable cost estimates for the proposed anaerobic reactor. According to a firm which specializes in the construction of anaerobic reactors for industrial waste treatment, it is rare to see a reactor vessel smaller than approximately 300,000 gallons (a volume which carried a price of approximately $400,000). In the event that only wastewaters generated by the retort process were to be treated, a reactor volume of less than 10,000 gallons (based on the HRTs and daily flows observed here) could be expected, further reducing the capital investment.

If an air stripping process of the type studied were to be employed, improvements in the electrical controls, tower strength and stability, and provisions for scale removal would be necessary in addition to previously mentioned design changes. These changes would result in higher capital costs than those seen in the construction of the tested tower. However, an examination of the costs experienced from this study may provide a baseline from which to work. The construction of the tower itself exclusive of the packing material, which was donated, was less than $1,000. It must be emphasized that the design changes stated will impact this figure significantly for the construction of an operational tower.

The packing material used in the study has a normal selling price of $20.45 per cubic foot and is only sold in 10 cubic foot lots. The amount of packing material needed will vary greatly depending upon the air temperature during operation, and the level of ammonia removals required. For example, under cold weather conditions (10°C), the three stage stripper setup described earlier for the treatment of anaerobically treated retort effluent would need a total of approximately 30 cubic feet of LANPAC to achieve removals necessary for direct discharge. However, a two-stage system used for ammonia removal of untreated retort effluent would use less than 10 cubic feet of packing material, if operated in warm weather conditions (25°C), to produce an effluent which could be discharged directly to the receiving water. In no case is it projected that the cost of the packing material will exceed $1,000.

5.0 CONCLUSIONS

The processing effluents from three blue crab processing plants in Virginia were characterized. It was determined that nine types of wastewater streams contributed the majority of the pollutant loadings from these plants. Most of these effluent streams were highly concentrated. For example, the retort water from the cookers exhibited the following concentrations: BOD_5 = 14,000-29,000 mg/L, TSS = 650-6,200 mg/L, TKN-N = 2,500-4,000 mg/L, NH_3-N = 70-160 mg/L and TP= 100-185 mg/L. The Harris Claw process produced effluent streams with chloride concentrations exceeding 100,000 mg/L.

Analysis of production and effluent characterization data from two mechanized plants revealed that wastewater volumes, loads and concentrations varied greatly between the two plants. Therefore, treatment alternatives for the blue crab plants should be developed on a case by case basis.

Settling and filtration studies were performed on these effluents. The organics and nutrients in the wastes were highly soluble and very little reduction of COD, BOD_5, TKN-N, NH_3-N and TP was achieved by these methods. Acidification of the concentrated effluents caused some coagulation of the contents. Coagulation by pH adjustment was most effective between pH 3.0-4.0. TSS removals of 76-93%, and BOD_5 removals of 6-30% were achieved by pH adjustment, thereby showing promise as a pretreatment method to reduce POTW surcharges.

Anaerobic biological treatment systems appear promising for the treatment of wastewater from blue crab processing plants (effluent BOD_5s ranged from 150-420 mg/L), especially for cooker effluent. However, there is concern that salt from the Harris Claw operation will decrease anaerobic treatment performance and that effluent NH_3 and TSS levels will need to be reduced. Pretreatment of process wastewater by means of pH adjustment did not improve anaerobic treatability.

Any use of this anaerobic biotechnology in the treatment of either retort wastewater to total plant discharge must, however, include provisions to diminish the effect of the sodium toxicity found in this study. Additional thought must also be given to the implications of the high concentration of ammonia found in the effluent from each reactor. Finally, while the anaerobic reactors treating retort water produced an effluent with TSS levels in the 200 to 650 mg/L range, which may be acceptable, the reactors which were fed the mixed wastewater (combination of all plant discharges) produced effluent TSS levels ranging on average from 1391 to 1587 mg/L. However, these levels could be reduced through the extension of the settling period beyond the one hour allowed in these experiments. It should also be noted that these solids contributed a relatively small (1% to 15%) portion of the total BOD_5 and COD found in the reactor effluent.

The air stripping experiments showed that substantial reductions in the ammonia concentrations found in the retort process wastewater are possible through the use of this technology, and that the high temperature of the retort process wastewater aids in the accomplishment of this reduction. Subsequently, it is suggested that the wastewater be treated as quickly as possible after the retort process in order to minimize any temperature loss. Additionally, while the tests performed showed higher removal rates were obtained as the pH of the wastewater was increased, this benefit must be weighed against the costs of chemicals needed to achieve this increase together with those needed to return the pH of the treated waste to a level that may be discharged directly to the environment. While some design modifications might be in order for the development of an operational air stripper, it is believed that this system holds promise as a possible treatment alternative.

While facilities which have BOD_5 discharge limits will, in all likelihood discharge their process wastewaters to a public sewer where possible, anaerobic biological and air stripping systems show promise as pretreatment steps which would reduce sewer charges. In such an application, anaerobic treatment, for the reduction of BOD_5, could be followed by air stripping in order to reduce ammonia. Alternatively, either system could be used alone to reduce a specific contaminant level.

If BOD_5 removal is not a concern, the air stripping technology tested appears to provide sufficient NH_3 removals to allow, if used in conjunction with settling for solids removal, direct release of the process waters to the environment. The applicability of either system, alone or in series, ultimately depends upon effluent limits currently being developed by the Virginia Department of Environmental Quality.

6.0 REFERENCES

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APPENDIX A
SAMPLE DISCHARGE PERMITS

COMMONWEALTH OF VIRGINIA STATE WATER CONTROL BOARD

PERMITTEE NAME/ADDRESS (INCLUDE FACILITY NAME/LOCATION IF DIFFERENT) SEASONAL

NATIONAL POLLUTANT DISCHARGE ELIMINATION SYSTEM (NPDES) DISCHARGE MONITORING REPORT (DMR)

INDUSTRIAL MINOR 08/08/88

STATE WATER CONTROL BOARD (REGIONAL OFFICE)

CRAB

002, 003,

NAME

ADDRESS

VA

FACILITY

LOCATION

FILE NO.

VA
PERMIT NUMBER

001
DISCHARGE NUMBER

REGIONAL OFFICE

MONITORING PERIOD

	YEAR	MO	DAY		YEAR	MO	DAY
FROM				TO			

804-

NOTE: READ PERMIT AND GENERAL INSTRUCTIONS BEFORE COMPLETING THIS FORM.

PARAMETER		QUANTITY OR LOADING			QUALITY OR CONCENTRATION				NO. EX	FREQUENCY OF ANALYSIS	SAMPLE TYPE
		AVERAGE	MAXIMUM	UNITS	MINIMUM	AVERAGE	MAXIMUM	UNITS			
001 FLOW	REPORTED			MGD	**********	**********	**********				
	PERMIT REQUIREMENT				**********	**********	**********			1/SEA	EST
002 PH	REPORTED	**********	**********			**********					
	PERMIT REQUIREMENT	**********	**********		5.0000	**********	9.0000	SU		1/SEA	GRAB
003 BOD5	REPORTED				**********						
	PERMIT REQUIREMENT			KG/D	**********			KG/K		1/SEA	4HC
004 TOTAL SUS. SOLIDS	REPORTED				**********						
	PERMIT REQUIREMENT	18.9000	56.7000	KG/D	**********	1.8600	5.5800	KG/K		1/SEA	8HC
500 OIL & GREASE	REPORTED				**********						
	PERMIT REQUIREMENT	6.1000	18.2900	KG/D	**********	.6000	1.8000	KG/K		1/SEA	GRAB
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										

ADDITIONAL PERMIT REQUIREMENTS OR COMMENTS PRODUCTION IS THE WEIGHT OF SHUCKED MEAT FOR OYSTERS; FOR OTHER SHELLFISH AND FISH, THE WEIGHT OF PRODUCT PRIOR TO PROCESSING. PRODUCTION WAS KG ON (DATE).

BYPASSES AND OVERFLOWS	TOTAL OCCURRENCES	TOTAL FLOW (M. G.)	TOTAL BOD5 (K G)

I CERTIFY UNDER PENALTY OF LAW THAT I HAVE PERSONALLY EXAMINED AND AM FAMILIAR WITH THE INFORMATION SUBMITTED IN THIS DOCUMENT AND ALL ATTACHMENTS AND THAT, BASED ON MY INQUIRY OF THOSE INDIVIDUALS IMMEDIATELY RESPONSIBLE FOR OBTAINING THE INFORMATION I BELIEVE THE SUBMITTED INFORMATION IS TRUE, ACCURATE AND COMPLETE. I AM AWARE THAT THERE ARE SIGNIFICANT PENALTIES FOR SUBMITTING FALSE INFORMATION, INCLUDING THE POSSIBILITY OF FINE AND IMPRISONMENT. SEE 18 U.S.C. § 1001 AND 33 U.S.C. § 1319. (Penalties under these statutes may include fines up to $10,000 and/or maximum imprisonment of between 6 months and 5 years.)

OPERATOR IN RESPONSIBLE CHARGE			DATE		
TYPED OR PRINTED NAME	SIGNATURE	CERTIFICATE NO.	YEAR	MO	DAY

PRINCIPAL EXECUTIVE OFFICER OR AUTHORIZED AGENT		TELEPHONE		DATE		
TYPED OR PRINTED NAME	SIGNATURE	AREA CODE	NUMBER	YEAR	MO	DAY

COMMONWEALTH OF VIRGINIA STATE WATER CONTROL BOARD

NATIONAL POLLUTANT DISCHARGE ELIMINATION SYSTEM (NPDES)
DISCHARGE MONITORING REPORT (DMR)

INDUSTRIAL MINOR 03/01/90

STATE WATER CONTROL BOARD (REGIONAL OFFICE)

REGIONAL OFFICE

PERMITTEE NAME/ADDRESS (Include Facility Name/Location if different)

NAME

ADDRESS

VA

FACILITY

LOCATION

FILE NO.

VA — PERMIT NUMBER

001 — DISCHARGE NUMBER

MONITORING PERIOD

	YEAR	MO	DAY		YEAR	MO	DAY
FROM				TO			

NOTE: READ PERMIT AND GENERAL INSTRUCTIONS [illegible] THIS FORM.

804-552-1840

PARAMETER		QUANTITY OR LOADING			QUALITY OR CONCENTRATION				NO. EX	FREQUENCY OF ANALYSIS	SAMPLE TYPE
		AVERAGE	MAXIMUM	UNITS	MINIMUM	AVERAGE	MAXIMUM	UNITS			
001 FLOW	REPORTED	************			**********	**********	**********				
	PERMIT REQUIREMENT	************	NL	MGD	**********	**********	**********			1/3M	MEAS
002 PH	REPORTED	**********	**********			**********					
	PERMIT REQUIREMENT	**********	**********		6.0000	**********	9.0000	SU		1/3M	GRAB
003 BOD5	REPORTED				**********						
	PERMIT REQUIREMENT	.0850	.1780	LBS/D	**********	28.0000	57.0000	MG/L		1/3M	GRAB
004 TOTAL SUS. SOLIDS	REPORTED				**********						
	PERMIT REQUIREMENT	.2240	.4420	LBS/D	**********	71.0000	147.0000	MG/L		1/3M	GRAB
500 OIL & GREASE	REPORTED				**********						
	PERMIT REQUIREMENT	.0440	.0780	LBS/D	**********	15.0000	39.0000	MG/L		1/3M	GRAB
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										

ADDITIONAL PERMIT REQUIREMENTS OR COMMENTS

BYPASSES AND OVERFLOWS	TOTAL OCCURRENCES	TOTAL FLOW (M.G.)	TOTAL BOD5 (K G)

OPERATOR IN RESPONSIBLE CHARGE

			DATE		
TYPED OR PRINTED NAME	SIGNATURE	CERTIFICATE NO.	YEAR	MO	DAY

PRINCIPAL EXECUTIVE OFFICER OR AUTHORIZED AGENT

		TELEPHONE		DATE		
TYPED OR PRINTED NAME	SIGNATURE	AREA CODE	NUMBER	YEAR	MO	DAY

I CERTIFY UNDER PENALTY OF LAW THAT I HAVE PERSONALLY EXAMINED AND AM FAMILIAR WITH THE INFORMATION SUBMITTED IN THIS DOCUMENT AND ALL ATTACHMENTS AND THAT, BASED ON MY INQUIRY OF THOSE INDIVIDUALS IMMEDIATELY RESPONSIBLE FOR OBTAINING THE INFORMATION I BELIEVE THE SUBMITTED INFORMATION IS TRUE, ACCURATE AND COMPLETE. I AM AWARE THAT THERE ARE SIGNIFICANT PENALTIES FOR SUBMITTING FALSE INFORMATION, INCLUDING THE POSSIBILITY OF FINE AND IMPRISONMENT. SEE 18 U.S.C. § 1001 AND 33 U.S.C. § 1319. (Penalties under these statutes may include fines up to $10,000 and/or maximum imprisonment of between 6 months and 5 years.)

PAGE OF

PERMITTEE NAME/ADDRESS (INCLUDE FACILITY NAME/LOCATION IF DIFFERENT)

COMMONWEALTH OF VIRGINIA STATE WATER CONTROL BOARD

NATIONAL POLLUTANT DISCHARGE ELIMINATION SYSTEM (NPDES)
DISCHARGE MONITORING REPORT (DMR)

INDUSTRIAL MINOR

03/01/90

STATE WATER CONTROL BOARD (REGIONAL OFFICE)

REGIONAL OFFICE

NAME

ADDRESS

VA

FACILITY

LOCATION

FILE NO.

VA
PERMIT NUMBER

002
DISCHARGE NUMBER

MONITORING PERIOD

FROM	YEAR	MO	DAY	TO	YEAR	MO	DAY

NOTE: READ PERMIT AND GENERAL INSTRUCTIONS [illegible] THIS FORM.

804

PARAMETER		QUANTITY OR LOADING			QUALITY OR CONCENTRATION				NO. EX	FREQUENCY OF ANALYSIS	SAMPLE TYPE
		AVERAGE	MAXIMUM	UNITS	MINIMUM	AVERAGE	MAXIMUM	UNITS			
001 FLOW	REPORTED	***********			●●●●●●●●●●	●●●●●●●●●●	●●●●●●●●●●				
	PERMIT REQUIREMENT	***********	NL	MGD	●●●●●●●●●●	●●●●●●●●●●	●●●●●●●●●●			1/3M	MEAS
002 PH	REPORTED	●●●●●●●●●●	●●●●●●●●●●			●●●●●●●●●●					
	PERMIT REQUIREMENT	●●●●●●●●●●	●●●●●●●●●●		6.0000	●●●●●●●●●●	9.0000	SU		1/3M	GRAB
003 BOD5	REPORTED				●●●●●●●●●●						
	PERMIT REQUIREMENT	.0850	.1700	LBS/D	●●●●●●●●●●	28.0000	57.0000	MG/L		1/3M	GRAB
004 TOTAL SUS SOLIDS	REPORTED				●●●●●●●●●●						
	PERMIT REQUIREMENT	.2240	.4480	LBS/D	●●●●●●●●●●	71.0000	147.0000	MG/L		1/3M	GRAB
500 OIL & GREASE	REPORTED				●●●●●●●●●●						
	PERMIT REQUIREMENT	.0440	.0780	LBS/D	●●●●●●●●●●	15.0000	39.0000	MG/L		1/3M	GRAB
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										

ADDITIONAL PERMIT REQUIREMENTS OR COMMENTS

BYPASSES AND OVERFLOWS	TOTAL OCCURRENCES	TOTAL FLOW (M.G.)	TOTAL BOD5 (K G)

OPERATOR IN RESPONSIBLE CHARGE

			DATE		
TYPED OR PRINTED NAME	SIGNATURE	CERTIFICATE NO.	YEAR	MO	DAY

PRINCIPAL EXECUTIVE OFFICER OR AUTHORIZED AGENT

		TELEPHONE		DATE		
TYPED OR PRINTED NAME	SIGNATURE	AREA CODE	NUMBER	YEAR	MO	DAY

I CERTIFY UNDER PENALTY OF LAW THAT I HAVE PERSONALLY EXAMINED AND AM FAMILIAR WITH THE INFORMATION SUBMITTED IN THIS DOCUMENT AND ALL ATTACHMENTS AND THAT, BASED ON MY INQUIRY OF THOSE INDIVIDUALS IMMEDIATELY RESPONSIBLE FOR OBTAINING THE INFORMATION I BELIEVE THE SUBMITTED INFORMATION IS TRUE, ACCURATE AND COMPLETE. I AM AWARE THAT THERE ARE SIGNIFICANT PENALTIES FOR SUBMITTING FALSE INFORMATION, INCLUDING THE POSSIBILITY OF FINE AND IMPRISONMENT. SEE 18 U.S.C. § 1001 AND 33 U.S.C. § 1319. (Penalties under these statutes may include fines up to $10,000 and/or maximum imprisonment of

COMMONWEALTH OF VIRGINIA STATE WATER CONTROL BOARD

PERMITTEE NAME/ADDRESS (INCLUDE FACILITY NAME/LOCATION IF DIFFERENT)

NATIONAL POLLUTANT DISCHARGE ELIMINATION SYSTEM (NPDES)
DISCHARGE MONITORING REPORT (DMR)

INDUSTRIAL MINOR 03/01/90

STATE WATER CONTROL BOARD (REGIONAL OFFICE)

REGIONAL OFFICE

NAME

ADDRESS

VA

FACILITY

LOCATION

FILE NO.

VAL
PERMIT NUMBER

000
DISCHARGE NUMBER

MONITORING PERIOD							
	YEAR	MO	DAY		YEAR	MO	DAY
FROM				TO			

NOTE: READ PERMIT AND GENERAL INSTRUCTIONS BEFORE COMPLETING THIS FORM.
804-552-1844

PARAMETER		QUANTITY OR LOADING			QUALITY OR CONCENTRATION				NO. EX	FREQUENCY OF ANALYSIS	SAMPLE TYPE
		AVERAGE	MAXIMUM	UNITS	MINIMUM	AVERAGE	MAXIMUM	UNITS			
001 FLOW	REPORTED	**********			**********	**********	**********				
	PERMIT REQUIREMENT	**********	NL	MGD	**********	**********	**********			1/3M	MEAS
002 PH	REPORTED	**********	**********			**********					
	PERMIT REQUIREMENT	**********	**********		6.0000	**********	9.0000	SU		1/3M	GRAB
003 BOD5	REPORTED				**********						
	PERMIT REQUIREMENT	.0400	.0800	LBS/D	**********	20.0000	40.0000	MG/L		1/3M	GRAB
004 TOTAL SUS. SOLIDS	REPORTED				**********						
	PERMIT REQUIREMENT	.1200	.2400	LBS/D	**********	60.0000	120.0000	MG/L		1/3M	GRAB
500 OIL & GREASE	REPORTED				**********						
	PERMIT REQUIREMENT	.0170	.0350	LBS/D	**********	8.5000	17.5000	MG/L		1/3M	GRAB
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										

ADDITIONAL PERMIT REQUIREMENTS OR COMMENTS

BYPASSES AND OVERFLOWS	TOTAL OCCURRENCES	TOTAL FLOW (M.G.)	TOTAL BOD5 (K G)

I CERTIFY UNDER PENALTY OF LAW THAT I HAVE PERSONALLY EXAMINED AND AM FAMILIAR WITH THE INFORMATION SUBMITTED IN THIS DOCUMENT AND ALL ATTACHMENTS AND THAT, BASED ON MY INQUIRY OF THOSE INDIVIDUALS IMMEDIATELY RESPONSIBLE FOR OBTAINING THE INFORMATION I BELIEVE THE SUBMITTED INFORMATION IS TRUE, ACCURATE AND COMPLETE. I AM AWARE THAT THERE ARE SIGNIFICANT PENALTIES FOR SUBMITTING FALSE INFORMATION, INCLUDING THE POSSIBILITY OF FINE AND IMPRISONMENT. SEE 18 U.S.C. § 1001 AND 33 U.S.C. § 1319. (Penalties under these statutes may include fines up to $10,000 and/or maximum imprisonment of between 6 months and 5 years.)

OPERATOR IN RESPONSIBLE CHARGE			DATE		
TYPED OR PRINTED NAME	SIGNATURE	CERTIFICATE NO.	YEAR	MO	DAY
PRINCIPAL EXECUTIVE OFFICER OR AUTHORIZED AGENT		TELEPHONE	DATE		
TYPED OR PRINTED NAME	SIGNATURE	AREA CODE / NUMBER	YEAR	MO	DAY

PAGE OF

COMMONWEALTH OF VIRGINIA STATE WATER CONTROL BOARD

PERMITTEE NAME/ADDRESS (INCLUDE FACILITY NAME/LOCATION IF DIFFERENT)

NATIONAL POLLUTANT DISCHARGE ELIMINATION SYSTEM (NPDES) DISCHARGE MONITORING REPORT (DMR)

INDUSTRIAL MINOR 03/01/90

STATE WATER CONTROL BOARD (REGIONAL OFFICE)

REGIONAL OFFICE

NAME

ADDRESS

VA

FACILITY

LOCATION

FILE NO.

VA
PERMIT NUMBER

004
DISCHARGE NUMBER

MONITORING PERIOD

FROM YEAR MO DAY TO YEAR MO DAY

NOTE: READ PERMIT AND GENERAL INSTRUCTIONS STARTING THIS FORM.

8044

PARAMETER		QUANTITY OR LOADING			QUALITY OR CONCENTRATION				NO. EX	FREQUENCY OF ANALYSIS	SAMPLE TYPE
		AVERAGE	MAXIMUM	UNITS	MINIMUM	AVERAGE	MAXIMUM	UNITS			
001 FLOW	REPORTED	***********			**********	**********	**********				
	PERMIT REQUIREMENT	***********	NL	MGD	**********	**********	**********			1/3M	MEAS
002 PH	REPORTED	**********	**********			**********					
	PERMIT REQUIREMENT	**********	**********		6.0000	**********	9.0000	SU		1/3M	GRAB
003 BOD5	REPORTED				**********						
	PERMIT REQUIREMENT	.0400	.0800	LBS/D	**********	20.0000	40.0000	MG/L		1/3M	GRAB
004 TOTAL SUS. SOLIDS	REPORTED				**********						
	PERMIT REQUIREMENT	.1200	.2400	LBS/D	**********	60.0000	120.0000	MG/L		1/3M	GRAB
500 OIL & GREASE	REPORTED				**********						
	PERMIT REQUIREMENT	.0170	.0350	LBS/D	**********	8.5000	17.5000	MG/L		1/3M	GRAB
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										

ADDITIONAL PERMIT REQUIREMENTS OR COMMENTS

BYPASSES AND OVERFLOWS	TOTAL OCCURRENCES	TOTAL FLOW (M. G.)	TOTAL BOD5 (K G)

OPERATOR IN RESPONSIBLE CHARGE

TYPED OR PRINTED NAME	SIGNATURE	CERTIFICATE NO.	DATE YEAR	MO	DAY

I CERTIFY UNDER PENALTY OF LAW THAT I HAVE PERSONALLY EXAMINED AND AM FAMILIAR WITH THE INFORMATION SUBMITTED IN THIS DOCUMENT AND ALL ATTACHMENTS AND THAT, BASED ON MY INQUIRY OF THOSE INDIVIDUALS IMMEDIATELY RESPONSIBLE FOR OBTAINING THE INFORMATION I BELIEVE THE SUBMITTED INFORMATION IS TRUE, ACCURATE AND COMPLETE. I AM AWARE THAT THERE ARE SIGNIFICANT PENALTIES FOR SUBMITTING FALSE INFORMATION, INCLUDING THE POSSIBILITY OF FINE AND IMPRISONMENT. SEE 18 U. S. C. § 1001 AND 33 U. S. C. § 1319. (Penalties under these statutes may include fines up to $10,000 and/or maximum imprisonment of between 6 months and 5 years.)

PRINCIPAL EXECUTIVE OFFICER OR AUTHORIZED AGENT

TYPED OR PRINTED NAME	SIGNATURE	TELEPHONE AREA CODE	NUMBER	DATE YEAR	MO	DAY

PAGE OF

COMMONWEALTH OF VIRGINIA STATE WATER CONTROL BOARD

NATIONAL POLLUTANT DISCHARGE ELIMINATION SYSTEM (NPDES)
DISCHARGE MONITORING REPORT (DMR)

INDUSTRIAL MINOR 02/22/90

STATE WATER CONTROL BOARD
(REGIONAL OFFICE)

REGIONAL OFFICE

PERMITTEE NAME/ADDRESS (INCLUDE FACILITY NAME/LOCATION IF DIFFERENT)

NAME

ADDRESS

VA

FACILITY

LOCATION

FILE NO.

PERMIT NUMBER: VA

DISCHARGE NUMBER: 005

MONITORING PERIOD: FROM YEAR MO DAY TO YEAR MO DAY

NOTE: READ PERMIT AND GENERAL INSTRUCTIONS [illegible] COMPLETING THIS FORM.

804

PARAMETER		QUANTITY OR LOADING AVERAGE	QUANTITY OR LOADING MAXIMUM	UNITS	QUALITY OR CONCENTRATION MINIMUM	QUALITY OR CONCENTRATION AVERAGE	QUALITY OR CONCENTRATION MAXIMUM	UNITS	NO. EX	FREQUENCY OF ANALYSIS	SAMPLE TYPE
001 FLOW	REPORTED	***********		MGD	**********	**********	**********				
	PERMIT REQUIREMENT	***********	NL		**********	**********	**********			1/3M	MEAS
002 PH	REPORTED	**********	**********			**********					
	PERMIT REQUIREMENT	**********	**********		6.0000	**********	9.0000	SU		1/3M	GRAB
003 BOD5	REPORTED			LBS/D	**********						
	PERMIT REQUIREMENT	NL	NL		**********	NL	NL	MG/L		1/3M	GRAB
004 TOTAL SUS. SOLIDS	REPORTED			LBS/D	**********						
	PERMIT REQUIREMENT	NL	NL		**********	NL	NL	MG/L		1/3M	GRAB
080 TEMP	REPORTED	**********	**********		**********	**********					
	PERMIT REQUIREMENT	**********	**********		**********	**********	32.0000	C		1/3M	GRAB
500 OIL & GREASE	REPORTED			LBS/D	**********						
	PERMIT REQUIREMENT	NL	NL		**********	NL	NL	MG/L		1/3M	GRAB
	REPORTED										
	PERMIT REQUIREMENT										
	REPORTED										
	PERMIT REQUIREMENT										

ADDITIONAL PERMIT REQUIREMENTS OR COMMENTS

BYPASSES AND OVERFLOWS	TOTAL OCCURRENCES	TOTAL FLOW (M. G.)	TOTAL BOD5 (K G)

OPERATOR IN RESPONSIBLE CHARGE

TYPED OR PRINTED NAME	SIGNATURE	CERTIFICATE NO.	DATE YEAR	MO	DAY

I CERTIFY UNDER PENALTY OF LAW THAT I HAVE PERSONALLY EXAMINED AND AM FAMILIAR WITH THE INFORMATION SUBMITTED IN THIS DOCUMENT AND ALL ATTACHMENTS AND THAT, BASED ON MY INQUIRY OF THOSE INDIVIDUALS IMMEDIATELY RESPONSIBLE FOR OBTAINING THE INFORMATION I BELIEVE THE SUBMITTED INFORMATION IS TRUE, ACCURATE AND COMPLETE. I AM AWARE THAT THERE ARE SIGNIFICANT PENALTIES FOR SUBMITTING FALSE INFORMATION, INCLUDING THE POSSIBILITY OF FINE AND IMPRISONMENT. SEE 18 U.S.C. § 1001 AND 33 U.S.C. § 1319. (Penalties under these statutes may include fines up to $10,000 and/or maximum imprisonment of between 6 months and 5 years.)

PRINCIPAL EXECUTIVE OFFICER OR AUTHORIZED AGENT

TYPED OR PRINTED NAME	SIGNATURE	TELEPHONE AREA CODE	NUMBER	DATE YEAR	MO	DAY

PAGE OF

APPENDIX B
AIR STRIPPING PACKING DEPTH CALCULATIONS

This appendix explains how the values for the COD loads, flow rates, reactor volumes and HRT's in Table 15 were determined. The COD loads and effluent volumes were calculated using a combination of the process wastewater characterization data measured during the study and the last 2-4 years of production data at plants #1 and #2. The process wastewater volumes and concentrations shown in Table 4 are based pounds or gallons per 1,000 pounds of product for each process effluent. These values can be multiplied directly by the amount of product processed to determine the resulting loading. For example, at plant #1 it was determined that 11.3 pounds of COD were created for every 1,000 pounds of live crabs cooked. If a blue crab processor cooked 20,000 pounds of live crab on a given day the resulting COD loading could be calculated as follows:

$$\frac{11.3 \text{ lb COD}}{1{,}000 \text{ lb live crab}} \quad \text{X} \quad 20{,}000 \text{ lb live crab} = 226 \text{ lb COD.}$$

The other process effluent loadings are calculated by the same method. Finally, all loadings from each process are added together to give the total effluent loading from the plant each day based on the amount of product processed. The effluent volumes were calculated with the data in Table 12 in gallons per 1,000 pounds of product. Please note that the retort effluent is presented as pounds or gallons per 1,000 pounds of live crab and all other effluents are presented as pounds per 1,000 pounds of final product.

The maximum, average, and minimum daily COD loads and effluent volumes on a monthly average were calculated using production data provided by the processors (Figures 5-10). The reactor volumes were determined using these COD loading values and the F/M ratios obtained from the anaerobic treatment studies. The following equation was used to calculate reactor volume:

$$\text{reactor volume} = \frac{(\text{COD load})}{(\text{F/M})(\text{MLVSS concentration})}$$

where:

reactor volume = gallons

COD load = lb/day

F/M = day^-1

MLVSS concentration = 0.03331 lb/gal (4,000 mg/L).

Derivation of the reactor volume for plant #1 in case #1 (Table 15) is shown below as an example. The maximum COD load in this case was 254 lb/day. The F/M was 0.05/day because combined effluent was treated. These values are plugged into the above equation as follows:

$$\frac{254\ \text{lb/day}}{(0.05/\text{day})(0.03331\ \text{lb/gal})} = 152{,}500\ \text{gallons}\ .$$

Finally, the hydraulic retention time (HRT) is determined for peak, average and low flow conditions by simply dividing the calculated reactor volume by the peak, average and low flow rate values. As an example, the peak flow HRT for plant #1, scenario #1 would be calculated as follows:

$$\frac{152{,}500\ \text{gallons}}{11{,}960\ \text{gallons/day}} = 13\ \text{days}\ .$$

The same method is used to calculated average and low flow HRTs.

APPENDIX C

CALCULATIONS FOR POTENTIAL ANAEROBIC TREATMENT SCENARIOS

PACKING DEPTH CALCULATIONS
FOR TESTED TOWER

Parameters:

Removal Rate = 95 %
Influent Concentration = 160 mg/l
Liquid Flow-Rate = 1.2 GPM
Liquid Loading Rate = 339.4 lbs/hr/ft^2
Air Flow-Rate = 500 cfm
Water Temperature = 15 deg. C
Henry's Constant (at 15 deg. C) = .57 atm

$$HTU = A * (L/M_L)^{0.33} * (M_L/\rho_L D_L)^{0.5} * (T/286)^{-4.255} \qquad (5)$$

Where:

HTU = Height of Transfer Unit in ft
A = Constant, 0.0042 for Ammonia
L = Liquid Loading Rate in lbs/hr/ft^2
M_L = Viscosity of Water in lbs/hr-ft

$$M_L = 4.3231 * (T/273)^{-7} \qquad (6)$$

M_L (at 15 deg. C) = 2.973 lbs/hr-ft

ρ_L = Density of Water, 62.4 lbs/ft^3
T = Temperature of Water in ○K
D_L = Diffusivity of Ammonia in Water in ft^2/hr

$$D_L = 6.3635 * 10^{-4} * (T/273)^{8} * (1/V_c^{0.6288}) \qquad (7)$$

D_L (at 15 deg. C) = $6.6032 * 10^{-5}$ ft^2/hr

V_c = Critical Molar Volume of Ammonia, 72.5 cm^3/g-mole

$$\text{HTU at 15 deg.C} = 0.0042 * (339.4/2.973)^{0.33} * [2.973/(62.4 * 6.6032*10^{-5})]^{0.5} * (288/286)^{-4.255}$$

HTU = .523 ft

$$NTU = \ln \frac{(C_1 - (1331.2/(H_c * AW)) * (C_1 - C_2))/C_2}{(1 - (1331.2/(H_c * AW))} \qquad (8)$$

Where:

NTU = Number of Transfer Units
C_1 = Influent Ammonia Concentration = 160 mg/l
C_2 = Effluent Ammonia Concentration desired = 8 mg/l

AW = Air to Water Ratio = 3121 cfm/cfm at stated air and liquid rates

$$\text{NTU} = \ln \frac{((160 - (1331.2/(0.57 * 3121)) * (160 - 8)/8}{(1 - (1331.2/(0.57 * 3121)))}$$

NTU = ln (22.97) = 3.13

Packing Height = (3 * HTU) * NTU * 1.1 (4)

Packing Height = (3 * .523 ft) * 3.13 * 1.1 = 5.41 ft