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Landfill Leachate Treatment Using Evaporation Technology

by

Aziz-ur Rehman

A thesis submitted under the supervision of Dr. Wayne Parker

Department of Civil and Environmental Engineering

Carleton University

Ottawa, Ontario

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Abstract

A pilot scale evaporator was designed to evaluate the use of evaporation technology for treatment of landfill leachate, where energy from combustion of landfill gases could be used for evaporation. Mass and energy balance calculations were made for all major components of the evaporator, along with a conceptual droplet evaporation model. Performance of the evaporator was evaluated by evaporating tap water and actual leachate from the Trail Road landfill site, Ottawa. During the evaporation process contaminants in the raw leachate were concentrated to a small volume. Exhaust air from the evaporator was used to preheat the leachate and released to the atmosphere. This stream may require some level of treatment prior to release. Mass balances were prepared to compare the fate of contaminant transport in different waste streams. Finally, it was determined that the landfill gases have sufficient energy to evaporate the landfill leachate produced at the Trail Road landfill site.

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Glossary

Abbreviations

ANOVA	= analysis of variance
B	= boron
Ba	= barium
BCL	= Bosik Consultants Limited
BOD	= biochemical oxygen demand
BTEX	= benzene, toluene, ethyl-benzene and xylenes
CH ₄	= methane
Cl	= chloride
CO ₂	= carbon dioxide
COD	= chemical oxygen demand
DFG	= dry flue gases
FID	= flame ionization detector
GC	= gas Chromatograph
H ₂ S	= hydrogen sulfide
HCl	= hydrochloric acid
IC	= ion Chromatograph
ISA	= Ionic strength adjustor
N	= normality
NH ₃	= ammonia
NH ₄	= ammonium
R	= universal gas constant
Re	= Reynolds number
SO ₄	= sulfates
SPME	= solid phase micro extraction
SVP	= saturated vapor pressure
TKN	= total kjeldahl nitrogen
TOC	= total organic carbon
TSS	= total suspended solids
U.S.EPA	= United States Environmental Protection Agency
VOA	= volatile organic acids
VOC	= volatile organic carbon

symbols

<i>A</i>	= Area
<i>B</i>	= mass transfer number
Btu	= British thermal unit
<i>C_d</i>	= drag coefficient
<i>C_g</i>	= gas phase VOC concentration
cm	= centimeter
<i>C_p</i>	= specific heat capacity
<i>C_s</i>	= saturation VOC concentration in liquid
<i>d</i>	= droplet diameter
<i>D_i</i>	= diameter at position <i>i</i>
<i>d_o</i>	= initial droplet diameter
<i>F_d</i>	= drag force
ft	= foot
<i>g</i>	= gas
g	= acceleration due to gravity
H	= height of the evaporation column
<i>H_c</i>	= Henry's law constant
hr	= hour
<i>h_v</i>	= heat of vaporization

in	= inch
k	= thermal conductivity
kg	= kilogram
kJ	= kilo-Joules
L_H	= latent heat of vaporization
lit	= litre
m	= mass
m	= meter
m^3	= cubic meter
M_{air}	= molecular weight of air
mg	= milligram
min	= minutes
mm	= millimeter
M_w	= molecular weight of water
$^{\circ}C$	= degree centigrade
$^{\circ}F$	= degree Fahrenheit
P	= ambient pressure
P_c	= critical pressure
ppm	= parts per million
Pr	= Prandtl number
P_{sat}	= saturation pressure
psi	= pounds per square inch
Q	= heat flow
sec	= seconds
T	= Temperature
T_{∞}	= ambient air temperature
T_c	= critical temperature
t_i	= time required to reach position i
T_w	= surface temperature
U	= heat transfer coefficient
v	= volume
V_i	= velocity at position i
w	= water
w_v	= water vapors
$Y_{f\infty}$	= liquid mass fraction in air
Y_{fr}	= mass fraction of the water inside the droplet surface
Y_{fw}	= mass fraction of the water at the liquid-gas interface
Y_i	= distance from the origin to position i
β	= evaporation coefficient
ΔH_{rxn}	= heat of reaction
Δt_i	= time required to cross the interval i
μg	= microgram
μl	= microlitre
μm	= micrometer
λ_w	= liquid density
λ_{air}	= density of the air
V_{rel}	= relative velocity
μ_{air}	= viscosity of the air
λ_{ev}	= evaporation rate coefficient

Chapter 1: Introduction

1.1 Background

The management of leachate from municipal waste landfills is a major aspect of landfill design and operation. Leachate is formed from water that infiltrates through a landfill. As leachate migrates through the refuse, it accumulates soluble contaminants present in the solid waste and soluble products from anaerobic decomposition reactions. Sanitary landfills are lined with impermeable materials to prevent the movement of leachate into the soil beneath the landfill. A drainage system is installed to convey leachate by gravity to a sump or storage tank for further treatment and disposal (Marks et al., 1994).

The composition of landfill leachate is site specific and depends on various factors. Typical contaminants in leachate include organic compounds (that contribute to BOD or COD), suspended solids, soluble metals, salts and volatile nonionic compounds such as ammonia (Marks et al., 1994). Due to the complex nature of leachate characteristics and increasingly stringent treatment standards it is difficult to treat the leachate in a single traditional biological treatment process.

The Trail Road landfill in the city of Ottawa, on average, generates 7.58 metric tons of landfill leachate per hour (City of Ottawa, 2002). Currently leachate from Trail Road landfill is trucked to the City's wastewater treatment facility. It has been proposed that the leachate be discharged to a sanitary sewer. However, several components of the leachate are present at higher concentrations than are allowed by the existing sewer use

bylaw. The City is considering to adopt some emerging technology for treatment of landfill leachate that can reduce the concentration of these components to the bylaw limits at source.

Evaporation may offer the potential to deal with leachate problem effectively in one principal unit operation at the landfill site. In addition, the two principal by-products of landfill operations, leachate and gas, could be used together beneficially in one process (Birchler et al., 1994). After evaporation the volume of the concentrated residual will be a small fraction of the original leachate and could be recycled back to the landfill. Volatile components in leachate that go in the exhaust air stream could be treated separately, if required.

1.2 Objectives

The objectives of this study were:

- To design and develop a pilot scale evaporation treatment unit for treatment of landfill leachate.
- To evaluate the impact of different operating conditions on the performance of the evaporator.
- To evaluate the impact of these operating conditions, on the fate of target contaminants in the landfill leachate.

Chapter 2: Literature Review

In this section, a brief introduction to landfill leachate, its characteristics, changes in the leachate properties with the age of a landfill site, and the characteristics of Trail Road landfill leachate are provided. Commonly used evaporation techniques, along with previous pilot or full-scale evaporation projects used for treatment of landfill leachate are also discussed briefly.

2.1 Landfill Leachate

Generally, landfill leachate is defined as a contaminated liquid that seeps through a solid waste disposal site. This liquid may enter a landfill from external sources like rainfall, surface drainage and groundwater or may be produced from decomposition of the waste.

2.1.1 Leachate Characteristics

Leachate generation from a landfill and its characteristics vary widely and depend upon many factors like precipitation, ambient temperature, landfill age, waste composition, moisture content, absorbent capacity of the waste, depth of the fill, compaction and permeability. It has been reported that leachate composition and strength vary from landfill to landfill and even within a given landfill (U.S. EPA, 1995).

The volume of leachate generation varies with the amount of precipitation and storm-water run-on and run-off, the volume of ground water entering the waste containing zone, and the moisture content and absorbent capacity of the waste material. Leachate

generation gradually increases for the first 5 to 10 years because the new landfills have good moisture holding capacity, which could retain some of the water that precipitate through landfill. Upon further aging leachate production from the landfill sites starts declining (U.S. EPA, 1995). With the age of the landfill, water is consumed in the formation of the landfill gas during the anaerobic decomposition of the organic constituents.

The composition and concentrations of contaminants in landfill leachates also changes with the age of the landfill. Typical values for new landfills will vary with the metabolic state of the landfill, as readily biodegradable organics undergo anaerobic degradation. Consequently, older landfill leachates are more stabilized and have lower concentrations of organics (U.S. EPA, 1995). Leachate concentrations are frequently reported as ranges, not as discrete values. These ranges are usually quite broad, often spanning several orders of magnitude. The typical composition of leachates from new and mature landfills are given in Table 2.1 (Tchobanoglous et al., 1993).

2.1.2 Trail Road leachate characteristics

The Trail Road landfill site mainly receives commercial and residential waste from the City of Ottawa. Leachate from the landfill is a mixture of leachates from new and old landfill areas. The maximum and minimum concentrations for some of the main contaminants in the Trail Road landfill leachate, analyzed over a period of one year, are provided in Table 2.2.

Table 2.1: Typical characteristics of landfill leachates

Constituent	Value (mg/lit)		
	New landfill (less than 2 years)		Mature landfill (more than 10 yr)
	Range	Typical	
pH	4.5 – 7.5	6	6.6 – 7.5
BOD	2,000 – 30,000	10,000	100 – 200
COD	3,000 – 60,000	18,000	100 – 500
TOC	1,500 – 20,000	6,000	80 – 160
TSS	200 – 2,000	500	100 – 400
Chloride	200 – 3,000	500	100 – 400
Sulfate	50 – 1,000	300	20 – 50
Organic nitrogen	10 – 800	200	80 – 120
Nitrate	5 – 40	25	5 – 10
Total phosphorus	5 – 100	30	5 – 10
Ortho phosphorus	4 – 80	20	4 – 8
Alkalinity	1,000 – 10,000	3,000	200 – 1,000
Total hardness	300 – 10,000	3,500	200 – 500
Calcium	200 – 3,000	1,000	100 – 400
Magnesium	50 – 1,500	250	50 – 200
Potassium	200 – 1,000	300	50 – 400
Sodium	200 – 2,500	500	100 – 200
Total iron	50 – 1,200	60	20 – 200

(Source: Tchobanoglous et al., 1993)

The Trail Road landfill leachate characteristics indicated that most of the components in the Trail Road leachate falls within the range of leachates from new and mature landfills as reported in Table 2.1. The results reported in Table 2.2 show that the Trail Road leachate characteristics varied widely, as would be expected for a landfill cell that is still active. Some of the contaminants are not in compliance with the City of Ottawa's sewer use bylaw and could be divided into two groups. Components in Group-1 (boron, barium, chloride, hydrogen sulfide, toluene and m/p-xylene) are not treatable at the City's wastewater treatment plant and can not be discharged above the permissible discharge limits, whereas, components in Group-2 (BOD, TSS and TKN) are treatable and could be discharged to the sewer above the discharge limits, against a surcharge as a treatment cost. The components, which are not in compliance with the City's Sewer Use Bylaw

represent a range of physical and chemical properties. Treatment of these target components will be studied in the proposed pilot scale evaporation project.

Table 2.2: Characteristics of the Trail Road landfill leachate

Components	Units	Maximum	Minimum	Average	City limits
pH	mg/lit	8.1	7.1	7.6	5.5 - 9.5
BOD*	mg/lit	4200	88	916	300
COD	mg/lit	6400	1060	2194	-
TSS*	mg/lit	700	5	50	350
TKN*	mg/lit	1070	136	664	100
Total phosphorus	mg/lit	6	2	4	10
Chloride*	mg/lit	1750	136	1480	1500
Hydrogen sulfide*	mg/lit	4.0	0.4	2	2
Sulfate	mg/lit	6.9	6.6	6.8	1500
Aluminum	mg/lit	1.2	0.7	0.9	50
Boron*	mg/lit	8.0	7.5	7.8	2
Barium*	mg/lit	1.2	0.9	1	1.2
Calcium	mg/lit	273	156	214	-
Iron	mg/lit	5.5	2.7	4.1	50
Potassium	mg/lit	843	478	660	-
Magnesium	mg/lit	207	140	173	-
Sodium	mg/lit	688	676	682	-
Phosphate	mg/lit	4.7	2.1	3.4	-
Strontium	mg/lit	17.1	8.6	12.9	5
Zinc	mg/lit	0.4	0.2	0.3	3
Benzene	µg/lit	9.3	0.7	5.5	100
Toluene*	µg/lit	120	3.2	50	100
Ethylbenzene	µg/lit	75.8	10.4	43.9	100
m/p-xylene*	µg/lit	195.0	24.5	106.7	100
o-xylene	µg/lit	86.4	14.4	49.3	100

(Source: City of Ottawa, 1999)

Note: * Exceeded City of Ottawa's Sewer Use Bylaw

2.2 Leachate treatment processes

Landfill leachate treatment systems and residual management are becoming more complex with the increasingly stringent effluent discharge quality standards. The type of

treatment facilities employed for landfill leachate treatment depend primarily on the leachate characteristics, size of the landfill and the expected useful life (Tchobanoglous et al., 1993). Leachate characteristics of concern for treatment purposes are suspended or dissolved solids, pH, COD, SO_4 , heavy metals and nonspecific toxic constituents. Leachate containing extremely high TDS concentration may be difficult to treat biologically. High COD values favor anaerobic treatment because aerobic treatment is expensive. High sulfate concentrations may limit the use of anaerobic treatment because of production of odors. Heavy metal toxicity is also a problem with many biological treatment processes (Tchobanoglous et al., 1993).

Due to the complex nature of the leachate characteristics, there are no conventional and relatively few non-conventional technologies, which could remove all of these components in one stage. Conventional treatment of landfill leachate may require several unit operations to remove the various contaminants to acceptable levels. A typical treatment train may include: a) aerobic or anaerobic biological process to stabilize biodegradable organic components and ammonia; b) adsorption, air stripping, or chemical oxidation to remove non-biodegradable organic compounds; and c) chemical precipitation to remove heavy metals (Marks et al., 1994).

2.2.1 Leachate treatment by evaporation

Leachate treatment by evaporation may offer the advantage of dealing with the leachate on the landfill site by employing fewer unit operations as compared to a conventional treatment process composed of several sequential unit operations. Evaporation allows

separation of volatile from non-volatile components. Solids and metals can be concentrated into a small volume after evaporation. In this study, the potential of evaporation technology for treatment of the target contaminants in the Trail Road leachate in one principal unit operation would be explored.

Landfill gas produced by the decomposition of landfill refuse might be used as an energy source for the evaporation of leachate. (Birchler et al., 1994). Thus two by-products from landfill operation, leachate and gas, may be used strategically in an evaporation treatment process.

2.3 Evaporation

Evaporation may be defined as an operation where a substance changes from the liquid state into the vapor state. "The objective of evaporation is to concentrate a solution consisting of a nonvolatile solute and a volatile solvent. In the overwhelming majority of evaporations the solvent is water. Evaporation is conducted by vaporizing a portion of the solvent to produce a concentrated solution or thick liquor" (McCabe et al., 1993).

2.4 Classification of evaporators

A number of different types of evaporator configurations are described in the literature. In selecting the type of evaporator for a particular application, properties of the feed, the product required and the source of heat are the key considerations. Traditional evaporators could be divided into the following four categories (Hewitt et al., 1993).

2.4.1 Film type evaporators

Film type evaporators are designed to vaporize a liquid flowing as a thin film over a heated surface. There are two basic hydraulic arrangements that can be used to produce a thin liquid film, namely, a falling film or a climbing film. In falling film evaporators (see Figure 2.1), the liquid flows downward under gravity over the heated surface while the vapor may flow either downward cocurrently with the liquid or upward in a countercurrent mode. In the climbing film evaporator, the liquid film flows upward cocurrently with the vapor. These evaporators usually employ banks of vertical tubes and come in many shapes and forms (Hewitt et al., 1993).

Falling film evaporators are found in many industrial applications. The inherent advantages of falling thin film flow are short contact times between the process fluid and the heated surface, high heat transfer coefficients, minimal pressure drop, minimal static head, and small process fluid holdup. Short contact time is a key factor when processing heat sensitive fluids such as in food processing or polymer de-volatilization (Alhusseini et al., 1998). Due to the small temperature drop through the film and mixing of the vapor stream with the incondensable gas mixture, the liquid film is not superheated, therefore, fouling associated with nucleate boiling does not usually occur (Hewitt et al., 1993).

The chief problem in these evaporators is that of distributing the liquid uniformly as a film inside the tubes. Normally, this is done by a set of perforated metal plates above a carefully leveled tube sheet, by inserts in the tube ends to cause the liquid to flow evenly into each tube, or by spider distributors with radial arms from which the feed is sprayed

at a steady rate on the inside surface of each tube or by an individual spray nozzle inside each tube (McCabe et al., 1993). The other main problem with these evaporators is fouling of the heat transfer surface.

2.4.2 Boiling evaporators

Many types of evaporators with vapor generation involve nucleate boiling at heated solid surfaces. In boiling processes, the initial stages of vapor generation involve the formation of bubbles at the heated surface. During this stage of nucleate boiling the mechanism of heat transfer differs from that of evaporation from a plain liquid surface. In these processes superheating of the liquid is required to initiate and sustain the bubble growth (Hewitt et al., 1993).

In these evaporators, the complexity of the boiling heat transfer phenomenon is evident, especially in the case of convective boiling. Two complementary mechanisms coexist for heat transfer, heat transfer resulting from nucleate boiling in the liquid film and heat transfer resulting from convection from the tube surface to the interface, where evaporation occurs (Sun and Hewitt, 2001).

Boiling evaporators are generally categorized into two types; evaporators employing boiling inside the tubes or boiling outside the tubes. Common examples of inside boiling evaporators are the short-tube, basket type and long-tube. The most common form of outside boiling evaporator employs banks of horizontal tubes mounted within a cylindrical vessel. Figure 2.2 shows simplified diagrams of the evaporators with boiling

inside and outside the tubes, whereas the real systems employ multiple tube banks (Hewitt et al., 1993).

2.4.3 Flash evaporators

When a liquid is exposed to a sudden pressure drop below its saturation pressure, all the heat cannot be contained in the liquid as sensible heat, and the heat surplus is transformed into latent heat of vaporization. This is the origin of the formation of vapor bubbles inside the liquid bulk and called flash evaporation and flashing phenomenon, resulting in a temperature drop of the liquid. Flash evaporation is used in processes of steam generation such as water desalination (Saury et al., 2001).

In flash evaporators the liquid is pumped through a heating vessel, then through a restrictor and into a flashing chamber where the pressure of the liquid is suddenly reduced to a value below that of its vapor pressure, resulting in rapid generation of vapors. A simple flash evaporator is illustrated in Figure 2.3. Flashing systems have the advantage of avoiding boiling on heated surfaces and the associated problems of fouling and corrosion (Hewitt et al., 1993).

2.4.4 Direct contact evaporators

In a direct contact evaporator, a hot gas is pumped into a vessel containing the liquid, and heat is transferred across the gas/liquid interface, vaporizing part of the liquid. The concentrated liquid leaves at the bottom of the vessel and a mixture of gas and vapor leaves at the top. Direct contact evaporation utilizes the sensible heat of the gas to

provide the required latent heat of evaporation without the need for complex metal heat exchanger surfaces. It is one of the cheapest and simplest methods of evaporation (Hewitt et al., 1993).

The heat transfer efficiency depends strongly on the interfacial surface between the continuous and the dispersed phase. In spite of the same hold up, the interfacial surface available for heat exchange can be very different for different size distributions of the dispersed phase. This method is widely used in many industrial processes, such as water desalination and purification, spray cooling, spray combustion, crystallization, solar, geothermal and ocean thermal power, and thermal energy storage systems (Song et al., 1998).

2.4.4.1 Classification of direct contact evaporators

The relationship between liquid and gas phases in the direct contact evaporators can generally be classified into two categories; one has gas as the continuous phase and the other has liquid as the continuous phase.

a) Continuous gas phase

In this category, mass and heat are transferred between a gas and a liquid in the form of drops, films or sprays. The drops may be formed by allowing liquid to flow through a perforated plate, over a weir or may be produced by a pressurized jet or spray nozzle. Often the liquid falls under gravity and the gas flows upward in a countercurrent direction. The more common forms of direct contact evaporators fall into this category.

They are used extensively in air conditioning, humidification, water-cooling and condensing plants. The simplest form of an evaporation spray column is shown in Figure 2.4 (Hewitt et al., 1993).

b) Liquid as the continuous phase

This form of evaporator employs a continuous liquid phase with gas in the form of bubbles, as illustrated in Figure 2.5. The liquid is contained in a vessel and the gas is introduced at the base through a perforated plate, rising through the liquid in the form of discrete bubbles. Such devices are also used for heating or cooling gas or for condensing vapor (Hewitt et al., 1993).

2.5 Selection of evaporator

The selection of the type of evaporator for a particular application can be influenced by the properties of the liquid (to be evaporated), the product required, the source of heat and the space available. While considering the evaporation of the landfill leachate containing unwanted dissolved material, scaling of heated surfaces is likely to occur and the consideration for cleaning is very important. The use of film or boiling type evaporators, which involve fine metal tubes or plates for heat transfer, was considered inadequate because of potential of scaling due to deposit of contaminants on the metal surface.

Direct contact evaporators and flashing systems have the advantage of avoiding boiling on heated surfaces and the associated problems of fouling and corrosion. The direct

contact technique using an evaporation column (spray column) was selected for this research work because of the following major advantages over the other evaporation processes.

- High mass and heat transfer rates associated with intimate mixing of the hot gas and liquid, which is mainly due to large interfacial surface areas available in dispersed particulate systems.
- Simplified construction and capital cost reduction due to elimination of the relatively complex tubes and metal walls, which impose thermal resistance and are subject to corrosion and deposition.
- The spray column serves to strip organic contaminants from the leachate into the exhaust vapor stream. This stripping action is analogous to that of conventional air strippers.

Direct contact evaporators are not very widely used for treatment of landfill leachate. One serious limitation to their application is the problem of contamination due to mixing of two streams, air and contaminated water. VOCs in the waste stream would be vaporized and create air pollution. Due to complex nature of the contaminants present in the leachate it is difficult in accurately predicting performance of the processes involving mixing of air and leachate streams.

2.6 Contaminant removal in evaporators

The process of evaporation could treat the inorganic compounds and organic pollutants in the leachate in only one process step (Leonhard et al., 1994). The target contaminants

studied for treatment of the Trail Road landfill leachate included TSS, COD, Cl, metals (B, Ba), volatile nitrogen compounds (TKN, NH_3), and volatile nonionic compounds (toluene and m/p-xylene). Dissolved and suspended solids could affect pretreatment and clean out schedules, chloride, heavy metals and pH could create scaling and corrosion problems, and volatile organic compounds could affect the air emissions (Fink and Hart, 2001).

During the evaporation process non-volatile substances should be concentrated and volatile organic compounds would be volatilized. The rate of volatilization from a water-air interface depends on the concentration of volatile organics in water, volatility of the individual compounds, surface area of the water exposed to gas phase and the surrounding gas phase conditions (Tansel and Eyma, 1999). As part of this study literature review was conducted to study the removal of the target contaminants in the landfill leachate, during the evaporation processes.

2.6.1 Dissolved and suspended solids

Dissolved and suspended solids in water could be organic or inorganic in nature due to the presence of volatile matter, metals, salts and minerals. Both dissolved and suspended solids may create problems in evaporation. Dissolved salts can raise the boiling temperature while higher salt concentrations require more heat and energy to evaporate the waste stream. Suspended solids tend to settle and could develop scaling on the walls of the evaporator, creating a barrier that insulates the heat source. Clean out schedules

could be dictated by the volume of dissolved and suspended solid in the waste stream (Fink and Hart, 2001).

An experimental study on the evaporation rate of a small water droplet containing dissolved salts was performed (King et al., 1997). Solutions of water containing 30% (w/w) and 60% (w/w) of potassium acetate or sodium iodide were evaporated on a heated, polished stainless steel surface during the experiments. Surface temperatures used in the experiments ranged from 50 °C to 100 °C. The average evaporation rates for the potassium acetate and sodium iodide solution were found to be lower than that of pure water. At a given surface temperature, the average mass evaporation rate decreased with the increase in concentration of the dissolved salt. King et al (1997) also developed an evaporation model that agreed qualitatively with the trends of the experimental data. The experimental and predicted average evaporation rates are shown in Figure 2.6. These results depict the effect of potassium acetate concentrations on the evaporation rates, at various surface temperature conditions.

2.6.2 Ammonia and COD

Leachate from sanitary landfills contains substantial amounts of organic and inorganic matter. Nitrogen is one of the important pollutants in leachate. Ammonia is a significant part of the total nitrogen and its concentration increases with the age of the leachate (Kabdasli et al., 2000). Removal of ammonia and COD primarily depends on the initial pH value of the leachate and the extent to which the COD is composed of volatile organic acids (Birchler et al., 1994). Birchler et al (1994) conducted laboratory scale distillation

experiments to study the effect of various pH conditions on the quality of the evaporate. He concluded that two-step acid-base evaporation could effectively remove ammonia and volatile organic acids from leachate, if present together in relatively high concentrations.

Marks et al (1994) prepared a flash vaporization model and evaluated the effects of pH variation on ammonia and COD removals from landfill leachate. In this model a single stage flash evaporation process was simulated using yearly average leachate characterization data from a landfill, which contained 578 mg/l NH_3 and 8800 mg/l COD. The leachate was preheated to 97 °C and flashed with 90% vaporization at 80 °C (Marks et al., 1994).

A one-step, acidic evaporation process simulation showed separation of 99% of ammonia from evaporate. By initially adjusting the pH level to 4, the ammonia in the leachate was transformed to ammonium ion that did not vaporize. However, 75% of the original acetic acid was predicted to carry over to the evaporate. Thus one step evaporation with acidic pH adjustment may be suitable for producing an evaporate having low concentrations of both ammonia and volatile organic acids (VOAs) only if the leachate contains initially low concentrations of VOAs.

A one-step basic evaporation process simulation showed separation of VOAs from the evaporate. With initial pH at 10, acetic acid was ionic and 99% remained in the concentrate, while the majority of the ammonia was transformed to a volatile species and 3% remained in the concentrate. Thus, a one step basic evaporation with basic pH

adjustment may be suitable for producing an evaporate free of both VOAs and ammonia only if low amounts of ammonia are present initially in the leachate.

A two-step acid /base evaporation process was simulated by adjusting pH to 4 in the first stage. In the second stage the evaporate from the first stage was condensed and its pH was adjusted to 10 prior to evaporation. The quality of the evaporate showed that over 99% of both the ammonia and the acetic acid were retained in the concentrate. Thus an acid /base, two-step evaporation process could be needed to obtain satisfactory evaporate quality for a concentrated leachate having high levels of both ammonia and VOAs.

2.6.3 Nonionic volatile organic compound

Evaporation of toluene was also reviewed by Marks et al (1994). Toluene is a non-ionic, volatile organic compound. The concentration of the toluene in the evaporate was between 74-76% of the total toluene for a single stage flash process simulations with pH adjustment of 4, 6.2 and 10. The concentrations of toluene in evaporate did not vary as a result of pH adjustment. Thus if a significant amount of non-ionic volatile organic compounds such as toluene are present in the leachate, additional measures may be necessary to obtain adequate evaporate quality.

2.7 Previous leachate evaporation projects

While evaporation is widely used in many industrial processes, application of evaporation technology for the treatment of landfill leachate has been limited. However, some examples of the use of this technology, in Europe and the United States for treatment of

the landfill leachate are available in literature. A literature review revealed that the following pilot or full-scale studies have been conducted to examine the treatment of leachate, using this technology.

2.7.1 Two-stage flash evaporation process

Marks et al (1994) reported the results of two-step evaporation and distillation tests that were performed on mixed waste landfill leachate of different strengths. In this pilot scale project ammonia was first removed by a basic stream stripping process and the residual liquid was then evaporated in two stage flash evaporators. Preheated leachate entered into the first evaporation chamber, evaporate from the first step was condensed and used as a feed into the next evaporator. The pH of the leachate was adjusted prior to either of the flash steps. The concentrated sludge from the evaporator was dewatered, bagged, and deposited back in the landfill.

The treatment capacity of the plant was about 3.6 m³ leachate/hr. Weekly average results of the full-scale facility showed that the treatment system achieved a 92% removal of COD and 94% removal of ammonia. The removal of phenol, the only individual trace volatile organic compound (VOC) reported, was 50%. The operational problems reported at this facility included foaming and scaling. Carbonate scaling problems were managed by flushing with acid and high-pressure rinse water. Materials such as graphite and Hastelloy C were used to control corrosion.

2.7.2 Semi-continuous four-stage flash evaporator with re-circulation

Birchler et al (1994) described a full scale, semi-continuous multi-stage evaporation process with recirculation of the leachate. Leachate was pretreated by adding HCl to keep NH_3 in solution, air stripping to remove CO_2 and volatile organic compounds, and vacuum degassing to remove dissolved air. The preheated leachate was treated in a four stage flash evaporator, where in each stage a part of the liquid was evaporated, passed through a demister, condensed and collected separately as shown in Figure 2.7. A fraction of the concentrate was continuously recycled and the concentrated sludge was disposed back to the landfill. The evaporation process was performed under vacuum conditions, with the heat from vapor condensation being used to preheat the feed leachate.

This treatment facility had a capacity of $20 \text{ m}^3/\text{day}$. The energy for evaporation was indirectly supplied by the combustion of the landfill gas, which was mainly used for electricity generation. The amount of electrical energy required for the facility was estimated about 13.5 kW.h/m^3 leachate. The results of the distillate quality showed removals of COD, ammonia and conductivity that were greater than 98.5% but the removal efficiency of the VOC's were not reported. No operational problems had been reported at this facility. Carbonates did not cause scale formation because of low pH and CO_2 removal. Sulfates did not precipitate as HCl was used to lower the pH. Corrosion of the material was avoided by coating working surfaces with elastomer or thermoplast.

2.7.3 Falling film evaporation on a polymeric heat transfer surface

A full-scale treatment plant with a capacity of 130m³/d was built in 1996 at the Lahti landfill in Finland (Ettala, 1998). The evaporator applied the falling film principle. The innovation of the technology was a new heat transfer surface concept, where water evaporates on a thin plastic film. The length of the plant was 10.5 m, width 5.0 m and height 5.6 m. The plant was equipped with 1500 m² of plastic heat transfer surface.

A low operating pressure (0.12 – 0.2 bar) and temperature (50 – 60 °C), allowed the use of plastic film as a heat transfer surface. Leachate fell on one side of the surface and the evaporate was collected under vacuum. The concentrate that left at the bottom was either recycled or sent back to the landfill. The condensate on the other side was collected separately. A simple diagram of the technology is presented in Figure 2.8. The energy required in this process was estimated to be about 13 kW.h/m³ for purified leachate. The influent COD (228 mg/l) and ammonia (101 mg/l) were successfully treated to below 30 mg/l and 2 mg/l, respectively, in the effluent. The volume of the concentrate was 18% of the influent.

The pretreatment requirements in this evaporation process included sand filtration. In addition, pH adjustment below 4 was used to achieve a good ammonium reduction. CO₂ was also released in a tower before evaporation. The pH of the condensate and concentrate streams also required adjustment before final disposal. Because of high calcium concentration in the leachate, gypsum precipitation limited the evaporation rate

and doubled the designed concentrate volume. Small amounts of antifoaming agents and cleaning and anti-scaling chemicals were also used in this process.

2.7.4 Direct contact leachate evaporation system

A leachate evaporation system (LES) was conceived to be a simple, reliable and effective tool for on-site leachate management at the Brookhaven Municipal landfill. The LES integrated landfill gas destruction and leachate control in a single process. A LES demonstration project was conducted in 1997 (Emcon, 1998).

In the LES, hot combustion gas produced in an internal air/fuel mix landfill gas burner was dispersed beneath a pool of landfill leachate. A proprietary gas sparge system that produces small bubbles that were uniformly dispersed in the leachate was employed. A centrifugal pump was used to feed leachate to the evaporator. This pump was also used to continuously mix leachate within the feed tank. All the wetted parts of the LES were constructed of 316 stainless steel. The evaporator vessel was connected to the enclosed landfill gas flare. Evaporator exhaust was transferred through sealed piping to the enclosed flare.

The internal evaporator temperature and pressure in the LES did not exceed 88 °C and 76 cm water column. Leachate volume was reduced greater than 98% in this evaporation process. Approximately 0.65 m³ of landfill gas were required to evaporate 3.78 litre of leachate in the LES. Analytical results of the feed and effluent streams demonstrated a 35% increase in concentration of the total solids.

2.8 Conclusions

From the review of the case studies, for treatment of landfill leachate using evaporation technology, it is clear that the use of this technology is recent and further demonstration of this technology is required before it gains acceptance by the engineering and regulatory communities. This technique has not been applied to landfill leachates in Canada. This literature review did not reveal the fate of all the target contaminants, present in the landfill leachates, during the evaporation processes. Much less has been presented about the transformation of VOCs. A recurring theme in the literature is that evaporation of landfill leachates should be evaluated on a case-by-case basis. The requirements for pretreatment of the leachate are an example of such a unique situation.

An evaporation column (spray column) technique was selected for this research work because it has significant advantages over the other evaporation processes like high mass and heat transfer rates, it can strip organic contaminants from the leachate into the exhaust vapor stream and it has a simplified construction due to elimination of the relatively complex tubes and metal walls, that impose thermal resistance and are subject to corrosion and deposition.

Figure 2.1: Falling film evaporator

(Source: Smith, 1986)

Figure 2.2: Nucleate boiling evaporators (a) inside and (b) outside tubes

(Source: Hewitt et al., 1993)

Figure 2.3: Flash evaporator

(Source: Smith, 1986)

Figure 2.4: Direct contact spray column with gas as continuous phase

(Source: Hewitt et al., 1993)

Figure 2.5: Direct contact bubble column with liquid as continuous phase

(Source: Hewitt et al., 1993)

Figure 2.6: Evaporation rates of potassium acetate at various temperatures

(Source: King et al., 1997)

Figure 2.7: Semi continuous four stage flash evaporator.

(Source: Birchler et al., 1994)

Figure 2.8: Falling film evaporation on a polymeric heat transfer surface

(Source: Ettala, 1998)

Chapter 3: Evaporation modeling

The transfer of heat and mass are two basic fundamental processes involved in evaporation. A basic conceptual model of the evaporator for this research along with the mass and energy balance concepts of the system are discussed. In addition, individual liquid droplet evaporation modeling in an evaporation column is provided in detail.

3.1 Conceptual model

The principal components in the conceptual model of the evaporation process for this research project included a combustor to produce hot combustion gases, a heat exchanger, for preheating of the leachate, an evaporation column and a pumping system to pump the leachate from the feed bucket to spray it in the evaporation column. In the conceptual design of the evaporation column, leachate was pumped from the feed bucket and sprayed in the top of the column through a spray nozzle. Hot combustion gases from the burner entered at the bottom of the evaporation column in a counter current direction. Heat and mass transfer occurs across the gas/liquid interface, vaporizing part of the liquid. The mixture of gases and vapors from the evaporation column passed through the heat exchanger to preheat the leachate and the remaining concentrate was left at the bottom of the column. Due to a temperature drop in the heat exchanger some of the liquid vapors condensed and left at the bottom of the heat exchanger. The process flow diagram of the evaporator is shown in Figure 3.1.

3.2 Mass and energy balances

Mass and energy balances were prepared for all major components of the pilot plant (combustor, evaporation column and heat exchanger). These mass and energy balances were preliminary and only intended to assist in process design. As discussed in the conceptual model of the pilot scale evaporator unit, hot combustion gases were used to evaporate the landfill leachate. Therefore the objective of the mass and energy calculations was to estimate the requirement of combustion gases to evaporate the target amount of the leachate.

To develop mass and energy balances for the complete unit, system boundaries were drawn for each component and input and output flow streams were identified, as depicted in Figure 3.1. Initially, mass and energy calculations were made across the evaporation column. Calculations were made to determine the mass and energy of the combustion gases required to evaporate about 10 litres of leachate per hour. In the second stage, mass and energy balance calculations were conducted for the combustor. This provided information about the mass of the fuel and excess air required in the combustion process to achieve the target mass and temperature of the combustion gases. These calculations gave background information that helped in the design and selection of the combustor. Heat and mass balance calculations for the heat exchanger allowed estimation of the condensate generation rates due to temperature changes in the heat exchanger.

It was assumed that the leachate consisted of clean water and the air supplied to the combustor comprised nitrogen and oxygen only. It was assumed that the fuel supplied in

the combustor was pure methane and complete combustion of the fuel took place. The specific heat capacity (C_p) of the air varies with the temperature and was calculated as

$$C_p = (a + bT + cT^2 + dT^3 + eT^4) \quad (\text{eq. 3-11})$$

Where T was the absolute temperature of the ambient air and a , b , c , d and e were constants having values of 3.953, -1.337E-3, 3.294E-6, -1.913E-9 and 2.736E-13, respectively (Jones and Hawkins, 1986).

The pressure inside the system was considered equal to atmospheric pressure and it was assumed that there was no change in the pressure of the evaporation column with the change in the temperature. It was assumed that there was no change in the density and viscosity of the leachate due to evaporation, and it was considered to be the same as that of water. It was assumed that the operating temperature inside the evaporation column was uniform and all the streams leaving the evaporator had the same temperature. It was also assumed that all the streams leaving the heat exchanger were leaving at the same temperature, which was considered to be equal to the internal temperature of the heat exchanger. The details of mass and energy balance calculations for each component are discussed in this section.

3.2.1 Combustor

Mass and energy balances were prepared for the combustor to predict the amount of fuel and air required, and to determine the mass and temperature of the combustion gases entering the evaporator. As shown in Figure 3.1, there were two feed streams entering the combustor. One was fuel (m_{CH_4} at temperature T_{CH_4} , and a specific heat capacity of $C_{p_{CH_4}}$),

which would be biogas at the landfill site and the other was fresh air (m_{air} at temperature T_{air} , and a specific heat capacity of Cp_{air}) that was required to provide oxygen for combustion. This fresh air also contained some water vapors (m_{wv} at temperature T_{wv} , and a specific heat capacity of Cp_{wv}). Combustion gases leaving the combustor consisted of dry flue gases (DFG) and water vapors.

It was assumed that the air supplied to the combustor was mainly composed of nitrogen (79%) and oxygen (21%) only, and the properties of air were used for calculation purposes. The fuel (methane) was assumed to be completely utilized in the combustion reaction. The radiation losses from the combustor were also assumed to be negligible.

A spreadsheet was prepared to compute the mass and energy balances for the combustor. A mass balance for complete combustion of one mole of methane, with 20% excess air is shown in Table 3.1, as an example. The mass of the DFG (m_{2g}) and mass of the water vapors (m_{2wv}) in the combustion gases were calculated on the basis of these calculations. An energy balance was also conducted for the combustor to estimate the temperature of the combustion gases. Of particular interest was the quantity of fuel that was required to achieve a desired temperature in the flue gases. The energy balance across the burner was calculated as

$$m_{air}Cp_{air}T_{air} + m_{CH_4}Cp_{CH_4}T_{CH_4} + m_{wv}Cp_{wv}T_{wv} + \Delta H_{rxn} = m_{2g}Cp_gT_2 + m_{2wv}Cp_{wv}T_2 \quad (\text{eq 3-1})$$

Where T_2 was the temperature of the combustion gases leaving the burner and ΔH_{rxn} was the heat produced in the combustion reaction. To calculate the total heat of reaction of the

fuel the numbers of moles of fuel used in the combustion reaction were multiplied with its heating value. For methane a lower heating value of 802.32 kJ per mole was used in these calculations.

Table 3.1: Mass balance of methane combustion reaction.

Components	Input (methane + air)		Output (Combustion Gases)	
	(k. mole)	(kg)	(k. mole)	(kg)
CH ₄	1.0	16.0	0.0	0.0
O ₂	2.4	76.8	0.4	12.8
N ₂	9.0	252.8	9.0	252.8
CO ₂	0.0	0.0	1.0	44.0
M dry gases	12.4	345.6	10.4	309.6
M H ₂ O vapors	0.0	0.0	2.0	36.0
Total =	12.4	345.6	12.4	345.6

3.2.2 Evaporator

As shown in Figure 3.1 there were two streams entering and two streams leaving the evaporation column. Preheated leachate (m_l at temperature T_l) entered at the top and a part of this was evaporated, and the water vapors (m_{5wv}) left the system in the evaporate stream. The remaining concentrate ($m_4 = m_l - m_{5wv}$) left the system at the bottom. Hot combustion gases, comprising DFG (m_{2g}) and water vapor (m_{2wv}) entered at the bottom. The gas stream leaving the evaporator at the top was a combination of the combustion gases and the components that were evaporated from the leachate in the system.

To calculate the amount of water evaporated (m_{5wv}) in the evaporation column it was assumed that the evaporate stream was completely saturated. The total amount of water vapor present in the evaporate stream was calculated from the vapor pressure of the

stream at the operating temperature. The relationship to estimate the saturated vapor pressure for water was used as

$$\text{Log}_{10} PVS = \frac{a}{T^3} + b \quad (\text{eq 3-2})$$

Where PVS was the saturated vapor pressure of water at operating temperature (T_3), and a and b were constants having values of -2238 and 8.896, respectively (Haug, 1991).

The net amount of the water evaporated in the evaporation column was the difference between the mass flow of vapors exiting and entering the column. A mass balance on water across the evaporator was developed as

$$m_1 = m_4 + m_{5wv} \quad (\text{eq 3-3})$$

The streams entering and leaving the evaporator were at different temperatures. The energy balance for the evaporator was written as

$$\begin{aligned} m_1 C_{p_w} T_1 + m_{2g} C_{p_g} T_2 + m_{2wv} C_{p_{wv}} T_2 = \\ m_4 C_{p_w} T_3 + m_{2g} C_{p_g} T_3 + m_{2wv} C_{p_{wv}} T_3 + m_{5wv} C_{p_{wv}} T_3 + hv_l m_{5wv} \end{aligned} \quad (\text{eq 3-4})$$

Where hv_l was the latent heat of vaporization used to convert m_{5wv} from liquid phase to vapor phase. It was assumed that the temperature inside the evaporator was uniform and all the streams leaving the evaporator were leaving at the same temperature (T_3). The total amount of the water vapor leaving the evaporator was calculated from the saturated vapor pressure of the evaporate stream at temperature T_3 as given in equation 3-2. The amount of leachate evaporated and concentrate produced in the evaporator were calculated using independent water balance equations for the evaporate stream and for the overall evaporator (equation 3-3). The operating temperature (T_3) was determined using solver (from excel) so that both the water balance (equation 3-2) and the energy balance

(3-3) equations were balanced to calculate the actual amount leachate evaporated in the system.

3.2.3 Heat Exchanger

The raw leachate (m_5) entered the heat exchanger at the top at a lower temperature (T_5) and left at the bottom after it was heated. The evaporate stream entered at the bottom of the heat exchanger and left at the top as exhaust gases, at a lower temperature (T_6). Due to the temperature difference some of the water vapors were expected to condense (m_6) and leave the heat exchanger, as a condensate stream.

The mass of the condensate stream was calculated as the difference between total amount of water vapor present in the evaporate stream and total amount of water vapor present in the exhaust gas stream. The partial pressure of water vapor in the evaporate and exhaust gas streams were assumed to be equal to the vapor pressure of the streams at the respective temperature conditions. A mass balance on water around the heat exchanger was written as

$$m_{5_{wv}} = m_6 + m_{7_{wv}} \quad (\text{eq 3-5})$$

The energy balance for the heat exchanger was written as

$$\begin{aligned} m_1 C_{p_w} T_5 + m_{2g} C_{p_g} T_3 + m_{2_{wv}} C_{p_{wv}} T_3 + m_{5_{wv}} C_{p_{wv}} T_3 + hv_1 m_{5_{wv}} = \\ m_6 C_{p_w} T_6 + m_1 C_{p_w} T_1 + m_{2g} C_{p_g} T_6 + m_{2_{wv}} C_{p_{wv}} T_6 + m_{7_{wv}} C_{p_{wv}} T_6 + hv_2 m_{7_{wv}} \end{aligned} \quad (\text{eq 3-6})$$

Where hv_2 is the latent heat of condensation released when m_6 was changed from vapor to liquid. It was assumed that the temperature inside the heat exchanger was uniform and all the streams leaving the heat exchanger had the same temperature (T_6). Total amount of

the water vapors was calculated from the saturated vapor pressure of the exhaust air leaving the heat exchanger at temperature T_6 , as given in equation 3-2. The amount of water vapors leaving and condensed in the heat exchanger were calculated using independent water balance equations for the exhaust air stream and for the overall heat exchanger (equation 3-5). The operating temperature (T_6) was determined using solver (from excel) so that both the water balance (equation 3-5) and the energy balance (3-6) equations were balanced to calculate the actual amount leachate evaporated in the system.

3.3 Mass and energy balance calculations

The pilot scale evaporation model had three streams entering (leachate, fuel and air) the evaporator system when the system boundaries were drawn around the complete evaporation unit. Based upon the mass and energy equations discussed in section 3.2, spreadsheets were developed to evaluate the effect of different leachate flow, fuel flow and excess air flow rates, entering the evaporation processes.

For a fixed amount of leachate flow, the evaporate and the condensate generation rates and the operating temperatures of the evaporation column and the heat exchanger were calculated as a function of the fuel flow and excess air flow rates. The fuel flow and excess air were two main variables, which control the mass and energy of the combustion gases or maintains temperature of the evaporation column that was required to evaporate the target volume of liquid. The results of the calculations for different fuel flow and airflow rates are discussed in the following sections.

3.3.1 Effect of air flow

Table 3.2 presents the estimated operating temperatures as a function of the excess air supplied to the combustor. Excess air was supplied to the combustor to ensure that complete combustion of the fuel had taken place. However, the temperature of the combustion gases also changed with the change in the supply of excess air. The combustor was supplied with a constant fuel (methane) flow of 1 kg/hr during these calculations. From Table 3.2 it can be seen that an increase in the supply of the excess air from 50 to 400 percent resulted in the predicted temperature of the combustion gases decreasing from 1658 to 539 °C.

The predicted operating temperatures inside the evaporator and heat exchanger were estimated during the mass balance calculations as given in Table 3.2. The results showed that with the increase in the percentage of excess air from 50 to 400 the operating temperatures in the evaporator unit could be decreased from 77 to 60, inside the evaporator, and from 54 to 47, inside heat exchanger, respectively.

The effects of excess air on the estimated evaporate and condensate generation rates are also given in Table 3.2. The results in the table demonstrates that changing the amount of excess air did not have a very significant effect on evaporate and condensate generation rates. For a fixed leachate flow rate (100 kg/hr) through the heat exchanger and evaporator column, the predicted amount of evaporate in the evaporator increased from 8.1 to 9.7 kg/hr, whereas, condensate decreased from 7.9 to 6.3 kg/hr, with a change in the excess air from 50 to 400 percent, respectively. The reason for this was that with the

change in the percentage of excess air there was no change in the heat energy of the combustion gases.

3.3.2 Effect of fuel flow

The temperatures of the combustion gases at different fuel flow conditions were estimated with the mass balance calculations and are given in Table 3.3. The percentage of excess air was also changed with the fuel flow rates to keep the mass of combustion gases at a constant rate of 50 kg/hr. All these mass balances were prepared with a leachate flow rate of 100 kg/hr. The change in the fuel flow rate resulted in a significant change in the temperature of the combustion gases. With a change of fuel flow rate from 0.5 to 2.0 kg/hr the temperature of the combustion gases increased from 474 to 1780 °C, respectively. This change in temperature was due to the high heating value associated with the increased amount of fuel flow.

The change in temperature of the combustion gases, due to different fuel flow rates also changed the operating temperatures inside the evaporator and heat exchanger. The predicted temperature conditions are shown in Table 3.3. The effects of fuel flow on evaporate and condensate generation rates are also shown in Table 3.3. The results demonstrated that the fuel flow rate had a significant effect on evaporate and condensate generation rates. The mass balance results show an overall increase in evaporate generation rate from 2.2 to 28.7 kg/hr and condensate generation rate from 2.6 to 12.9 kg/hr when the fuel flow rate was increased from 0.5 to 2.5 kg/hr.

Table 3.2: Effect of airflow on combustion gases, evaporate and condensate

Percentage of excess air	Mass of combustion gases	Combustion gases temperature	Evaporate	Condensate	Temperature of evaporator	Temperature in heat exchanger
(%)	(kg/hr)	(°C)	(kg/hr)	(kg/hr)	(°C)	(°C)
50	27	1658	8.1	7.9	77	54
100	36	1276	8.5	7.6	73	53
200	53	875	9.1	7.1	67	51
400	88	539	9.7	6.3	60	47

Table 3.3: Effect of fuel flow on combustion gases, evaporate and condensate.

Fuel flow (CH ₄)	Percentage of excess air	Combustion gases temperature	Evaporate	Condensate	Temperature of evaporator	Temperature in heat exchanger
(kg/hr)	(%)	(°C)	(kg/hr)	(kg/hr)	(°C)	(°C)
0.5	472	474	2.2	2.6	48	27
1	183	924	9	7.2	68	51
1.5	87	1359	18.3	10.7	78	68
2	39	1780	28.7	12.9	83	77

3.4 Mass and energy balance significance

During mass and energy balance calculations the theoretical mass and energy transferred across the system boundaries of each principal component of the evaporator unit was estimated. The effect of different airflow and fuel flow rates on the operating temperatures of the evaporator and heat exchanger were estimated. The evaporate and the condensate generation rates were also predicted during these mass and energy calculations.

Design and selection of all the major components of the conceptual model of the pilot scale evaporator were primarily based on the mass and energy balance calculations discussed in section 3.3.1 and 3.3.2. Selection of the combustor and design of the heat exchanger are discussed in this section. The design of the evaporation column and behavior of liquid droplets evaporating in the pilot evaporator unit will be discussed separately in section 3.6.

3.4.1 Selection of combustor

It was initially assumed that the pilot scale processes would operate at evaporator temperatures ranging from 40 to 80 °C. Based on the mass balance calculations in Table 3.3, the estimated range of fuel flow required to achieve these temperatures in the evaporation column was between 0.3 to 1.5 kg/hr. The results of the different excess air supplies (Table 3.2) had shown that about 27 to 88 kg/hr of air was required for combustion of 1 kg/hr fuel supply, when the excess air percentage was increased from 50 to 400. For combustion of 0.3 to 1.5 kg/hr of fuel the estimated amount of air supply was

considered to be between the range of 25 to 100 kg/hr. Therefore, for the design of the combustor, it was considered that the fuel supply in the combustor could be adjusted from 0.3 to 1.5 kg/hr and the air supply could be varied from 25 to 100 kg/hr.

3.4.2 Design of the heat exchanger

The main objective of the heat exchanger was to heat the raw leachate, entering the system at a lower temperature. The hot evaporate stream coming from the evaporation column was used to preheat the leachate in the heat exchanger. The evaporate stream left the heat exchanger at a lower temperature as some of the water vapors in the evaporate stream were condensed. The estimated amount and temperatures of the evaporate stream and raw leachate, entering and leaving the heat exchanger were estimated during the mass balance calculations (section 3.3).

A generalized method introduced by Hewitt et al (1993) for the heat exchanger design, when two fluids flowing in the opposite direction to each other, was used to calculate the area of the heat transfer surface. In this method, the surface area (A) required is given by

$$A = \frac{Q}{U\Delta T} \quad (\text{eq 3-7})$$

Where Q was the heat transferred, U was the overall heat transfer coefficient and ΔT was the logarithmic mean temperature difference across the surface area and was calculated as (Hewitt et al., 1993)

$$\Delta T = \frac{(T_{h,in} - T_{c,out}) - (T_{h,out} - T_{c,in})}{\ln[(T_{h,in} - T_{c,out}) - (T_{h,out} - T_{c,in})]} \quad (\text{eq 3-8})$$

Where $T_{h,in}$ and $T_{h,out}$ were the inlet and outlet temperatures of the evaporate stream and $T_{c,in}$ and $T_{c,out}$ were the leachate temperatures entering and leaving the heat exchanger.

The values of the heat energy transferred from the hot evaporate stream to the cold raw leachate, and the temperatures of all the streams entering and leaving the heat exchanger, calculated during mass balance calculations, were used for design of the heat exchanger. A 6.4 mm diameter stainless steel tubing was selected as a heat transfer surface. The value of heat transfer coefficient (U) used for stainless steel was $105 \text{ W/m}^2\cdot\text{K}$ (Hewitt et al., 1993). The total length of the tubing was estimated to be about 10 m. A coil consisting of eight loops was prepared as the heat transfer surface and placed in the heat exchanger assembly.

3.5 Droplet Evaporation Modeling

The main objective of this section was to develop a model, which would assist in the design of an evaporation column by predicting the evaporation behavior of a single liquid droplet. The rate of vaporization of a droplet depends mainly upon the liquid properties, the size of droplet and the ambient temperature. In order to determine this rate, it is necessary to know the profiles of temperature and composition of both the liquid and the gas phases (Kanury, 1995). Physically, heat from the ambient environment supplies the necessary energy to vaporize the liquid droplet, and the vapor then diffuses from the droplet surface into the ambient gas. The mass loss causes the droplet radius to shrink with time until the droplet is completely evaporated (Turns, 1996).

The model was developed on the basis of evaporation of a single water droplet. It calculates the droplet radius as a function of time and hence the droplet lifetime, when the surroundings were considered to be infinite and when the droplet is falling in a flowing air stream. Different approaches have been used for modeling the evaporation of liquid droplets. Changes in the physical properties of the droplet with the vaporization process were modeled by Mellor (1990). This was a relatively simple model based on spherically symmetrical droplet vaporization in an infinite dry medium.

3.5.1 Evaporation of droplet in an infinite surroundings

The approach taken by Mellor for single droplet evaporation model in an infinite environment contained the following assumptions (Mellor, 1990):

1. Quasi-steady evaporation.
2. No free or forced convection in the drop.
3. Uniform and constant drop temperature.
4. Constant pressure.
5. Local thermodynamic equilibrium at the drop surface.
6. Spherical drop.
7. Single component drop.
8. Radiative heat transfer neglected.
9. Constant specific heat.
10. Lewis number of unity.

The temperature and concentration profile around a droplet (evaporating surface is assumed to be flat) exposed to hot air is shown in Figure 3.2. In the Figure, T_∞ represents the ambient air temperature, T_w is the surface temperature of the droplet and $Y_{f\infty}$ is the gas phase liquid concentration. $Y_{f\infty}$ was considered to be equal to 0 because it was assumed that droplet is in the infinite dry medium. The mass fraction of the water inside the droplet surface (Y_{fr}) was equal to 1 since it was assumed that no dissolved air was in the droplet. Y_{fw} is the mass fraction of the water at the liquid-gas interface.

The mass evaporation rate and the droplet radius were determined by writing liquid and gas phase mass conservation equations and a gas phase energy equation. The gas phase energy equation gave the temperature distribution in the gas phase, which, in turn, allowed the evaluation of the conductive heat transfer into the droplet at the surface. To determine the vaporization rate the value of a mass transfer number (B) was calculated, which was defined from energy considerations as (Mellor, 1990)

$$B_T = \frac{C_{p,g}(T_\infty - T_w)}{L_H} \quad (\text{eq. 3-9})$$

and from species considerations as

$$B_D = \frac{Y_{f\infty} - Y_{fw}}{Y_{fw} - Y_{fr}} \quad (\text{eq. 3-10})$$

Where $C_{p,g}$ was the specific heat capacity of the air and L_H was latent heat of vaporization of the droplet. L_H was calculated as a function of T_w as

$$L_H = (597 - 0.581 * T_w) * 4.187 \quad (\text{eq. 3-12})$$

Y_{fw} was given as

$$Y_{fw} = \frac{1}{\left[1 + \frac{M_{air}}{M_w} \left(\frac{P}{P_{sat}} - 1 \right) \right]} \quad (\text{eq. 3-13})$$

M_{air} was the molecular weight of air, M_w was the molecular weight of water, P was ambient pressure and P_{sat} was saturation pressure, which was calculated as (Reid et al., 1987)

$$P_{sat} = P_c * e^{\left[\frac{(VPA * X + VPB * X^{1.5} + VPC * X^3 + VPD * X^6)}{(1-X)} \right]} \quad (\text{eq. 3-14})$$

Where $X = 1 - \frac{T_w}{T_c}$ (eq. 3-15)

and P_c and T_c were the critical pressure and temperature of the liquid, VPA , VPB , VPC and VPD were constants having values of -7.76451, 1.45838, -2.77580 and -1.23303 (Reid et al, 1987).

Physically, the mass transfer number is a measure of the driving force for evaporation. In terms of the energy equation, B_T was the ratio of the energy available for evaporation divided by the energy required. From the species equation B_D was the difference in vapor mass fraction between the environment and the surface divided by the difference between the surface and the liquid. When the Lewis number is unity, the energy and the species equations become one and the same, $B_T = B_D$ (Mellor, 1990).

$$B = \frac{C_{p,g} (T_{\infty} - T_w)}{L_H} = \frac{Y_{f\infty} - Y_{fw}}{Y_{fw} - Y_{fr}} \quad (\text{eq. 3-16})$$

Once the values of C_{pg} , L_H and Y_{fw} were determined for a given temperature and pressure conditions, the value of B and T_w were calculated using equation 3-16. To determine the expression for the droplet diameter as a function of time, the well known “d² law” was used (Mellor, 1990) as given below.

$$d^2 = d_o^2 - \beta * t \quad (\text{eq. 3-17})$$

Where d_o was initial droplet diameter and d was droplet diameter after time t . β was the evaporation coefficient and was given as (Mellor, 1990)

$$\beta = \frac{8k}{\lambda_w \cdot C_{pg}} \ln(1 + B) \quad (\text{eq. 3-18})$$

Where λ_w was the liquid density and k was thermal conductivity of air at a given temperature and was calculated as (Gupta et al., 1991).

$$k = 5.9776 * 10^{-6} \frac{T^{\frac{3}{2}}}{(T + 194.4)} \quad (\text{eq. 3-19})$$

Equation 3-17 indicates that the square of the droplet diameter decreases linearly with time. A droplet evaporation time was estimated from this equation by setting the final value of d equal to zero and solving as (Mellor, 1990)

$$t = \frac{d_o^2}{\beta} \quad (\text{eq. 3-20})$$

The key result here was that droplet evaporation time is proportional to the square of the initial diameter. Any change in the spray nozzle or pressure in the system, that changes the droplet size, can strongly influence the evaporation rate.

3.5.2 Evaporation of droplet while falling in air flowing upstream

Based on the assumptions and findings of the previous section, the evaporation model was improved to evaluate the evaporation of a droplet while falling in air that is flowing in the opposite direction. In this case the properties of the liquid including density, viscosity and surface tension also influence the evaporation coefficient, and are typically taken into account with empirical correlations. A widely used correlation due to Ranz and Marshall is given as (Mellor, 1990)

$$\lambda_{ev} = \beta * (1 + 0.3 * Re^{1/2} * Pr^{1/3}) \quad (\text{eq. 3-21})$$

Where λ_{ev} was the evaporation rate coefficient, Re was the Reynolds number and Pr was the Prandtl number, and were calculated as

$$\text{Reynolds number} \quad Re = \frac{\lambda_{air} * V_{rel} * d}{\mu_{air}} \quad (\text{eq. 3-22})$$

Where λ_{air} was density, V_{rel} was the relative velocity and μ_{air} was the viscosity of the air and were calculated as (Gupta et al., 91)

$$\lambda_{air} = \frac{P}{R * T} \quad (\text{eq. 3-23})$$

$$\mu_{air} = 1.4584 * 10^{-5} * \frac{T^{3/2}}{(T + 110.33)} \quad (\text{eq. 3-24})$$

Where R was the universal gas constant.

$$\text{Prandtl number} \quad Pr = \frac{Cp_{air} * \mu_{air}}{k_{air}} \quad (\text{eq. 3-25})$$

The evaporation rate coefficient (λ_{ev}) replaced the evaporation coefficient β in the d^2 law to calculate the droplet evaporation time while falling in air flowing upstream

$$t = \frac{d_o^2}{\lambda \cdot ev} \quad (\text{eq 3-26})$$

To further develop this model, the evaporation column with height H was divided into a number of segments (n) as shown in Figure 3.3. ΔH was the depth of each segment, i was the segment position, D_i and V_i were the diameter and velocity of the droplet at position i , Y_i was the distance from the origin (positive downward) and t_i was the time required to reach position i . D_o , V_o and Y_o were the diameter, velocity and distance of the droplet at $i = 0$. Δt_i was the time required to cross the interval i .

Calculations were performed while

1. the droplet diameter was greater than zero
2. the droplet velocity was greater than or equal to zero, and
3. the droplet had traveled past the last interval.

For each interval the following calculations were performed.

Time to cross segment i was calculated as

$$\Delta t_i = \frac{\Delta H}{V_{i-1}} \quad \text{if } V_{i-1} \neq 0 \quad (\text{eq 3-27})$$

otherwise

$$\Delta t_i = \sqrt{\frac{2\Delta H}{g}} \quad (\text{eq 3-28})$$

Where V_{i-1} was the droplet velocity at interval $i-1$ and g was the acceleration due to gravity. Time for droplet to travel to a position i was

$$t_i = t_{i-1} + \Delta t_i \quad (\text{eq 3-29})$$

The droplet diameter when it reached position i was calculated using the “ d^2 law” as

$$D_i = \sqrt{D_{i-1}^2 - \lambda_{ev} * \Delta t_i} \quad (\text{eq 3-30})$$

Where D_{i-1} was the droplet diameter at interval $i-1$ and λ_{ev} was calculated according to equation 3-18. Droplet position at time t_i was calculated as

$$Y_i = Y_{i-1} + V_{i-1} * \Delta t_i \quad (\text{eq 3-31})$$

Newton’s second law stated the relationship between the droplet falling velocity and its driving force as

$$\sum Fy = m \frac{dV}{dT} = m \frac{(V_i - V_{i-1})}{\Delta t_i} \quad (\text{eq 3-32})$$

In an evaporation column two forces act on a droplet falling, the force of gravity and the drag force. The net force was calculated as

$$\sum Fy = mg - Fd \quad (\text{eq 3-33})$$

From equation 3-22 and 3-23

$$V_i = V_{i-1} + \Delta t_i * \left[g - \frac{Fd}{m} \right] \quad (\text{eq 3-34})$$

Where m was mass of the liquid droplet and Fd was drag force. The mass of the liquid droplet was calculated as

$$m = \lambda_w * v \quad (\text{eq 3-35})$$

Where λ_w was the density of liquid and v was the volume. The drag force acting on the droplet was calculated as

$$Fd = \frac{1}{2} * Cd * \lambda_{air} * V_{rel}^2 * A \quad (\text{eq 3-36})$$

Where Cd was the drag coefficient, λ_{air} was the density of air, V_{rel} was the relative velocity of the falling droplet, A was area of the droplet, and were calculated as.

$$A = \frac{\pi}{4} D^2 \quad (\text{eq 3-37})$$

The relative velocity was calculated as

$$V_{rel} = V_i - V_{air} \quad (\text{eq 3-38})$$

Cd was calculated on the basis of the Re value under the following different conditions (Bardon and Goodger, 1986).

$$Cd = \frac{24}{Re} \text{ if } 0 < Re < 0.2 \quad (\text{eq 3-39})$$

$$Cd = \left(\frac{24}{Re} \right) \left[\left(1 + 3 * \frac{Re}{16} \right)^{0.5} \right] \text{ if } 0.2 \leq Re < 100$$

$$Cd = \frac{18.5}{Re^{0.6}} \text{ if } 100 \leq Re < 500$$

$$Cd = 0.44 \text{ if } 500 \leq Re < 10^5$$

The droplet diameter and velocity varied with the time of evaporation, as the droplet fell towards the bottom of the evaporation column. These values were calculated for a number of intervals of 10 cm segment depths throughout the evaporation column to design the depth of the column and to predict the evaporation rates in that column by changing different variables. The fraction of the droplet mass that had evaporated at a certain interval was calculated as

$$\Omega = \frac{m_{vapor}}{m_o} = \frac{m_o - m_{liquid}}{m_o} = 1 - \left(\frac{D}{D_o} \right)^3 \quad (\text{eq 3-40})$$

3.6 Model predictions

Based on the previously described model a spreadsheet was developed to study the evaporation rate of the water droplet over a range of process conditions and column depths. The model was employed to assist in the design of the pilot evaporator. In addition, the evaporation behavior of the droplet was analyzed to assess the impact of different control variables. The impact of initial droplet size, ambient temperature, droplet velocity and air velocity were studied on the fraction of droplet evaporated in a 1-meter deep evaporation column.

3.6.1 Depth of evaporation column

Initially the evaporation model was used to estimate the depth of the evaporation column. Simulations were conducted to calculate the total falling depth of the droplet so that the droplet could be evaporated completely. Different droplet sizes were used to check the estimated depth of the evaporation column keeping the ambient temperature, initial droplet falling velocity and airflow conditions constant at 90 °C, 10 m/sec and 10 cm/sec, respectively. The results of the column depths required to evaporate droplets of 200 to 1000 μm are given in Table 3.4. These calculations show that a droplet of 200 μm was evaporated completely in a 1 m deep evaporation column.

3.6.2 Initial droplet size

In a spray column the initial droplet size is a function of the spray nozzle and pressure in the water supply system. In the simulations, droplet diameters ranging from 200 to 1000 μm were used. The resulting percentage of droplet evaporation for the given 1 meter

evaporation column were studied at a constant ambient temperature (70°C), initial droplet velocity of 10 m/sec and air velocity of 10 cm/sec. The results of the fraction of the droplets evaporated in this evaporation column are shown in Figure 3.4. It can be seen that droplet size is an important parameter in the evaporation process. Small initial drops were evaporated quickly. The results showed that about 24 percent of a 200 μm droplet size was evaporated in a 1 m deep evaporation column. These initial calculations helped in selection of the spray nozzles to achieve the target droplet sizes.

Table 3.4: Depth of evaporation column to evaporate different droplet sizes

Droplet Size (μm)	Depth of the evaporation column (cm)
100	70
200	100
300	150
500	340
800	960

3.6.3 Ambient temperature

The impact of ambient temperature on droplet evaporation was calculated with an initial 200 μm droplet, an initial falling velocity of 10 m/sec and an initial air velocity of 10 cm/sec. The predicted fraction of the droplet evaporated at different evaporator temperatures are illustrated in Figure 3.5. This figure shows that with the increase in ambient temperature conditions the fraction of droplet evaporated also increased. The ambient temperature of the evaporation column was considered very significant because it provided the energy required for the droplet to vaporize. The mass and energy balance

calculations across the combustor helped in determining different temperature conditions inside the evaporation column by adjusting fuel and airflow rates.

3.6.4 Air velocity

The impact of air velocity on the evaporation of a 200 μm droplet at 70 $^{\circ}\text{C}$ temperature and 10 m/sec initial velocity are depicted in Figure 3.6. The results indicate that with an increase in the air velocity the fraction of the droplet mass evaporated increased. With air flowing in a countercurrent direction to the falling droplet, a higher air velocity will increase the heat transfer rate, and produce more drag force and hence will reduce the falling velocity of the droplets and thereby increase the residence time for evaporation. The mass of air supply to the combustor was used to determine the air velocity in the evaporation column.

3.6.5 Initial droplet velocity

The initial droplet velocity also contributes towards the evaporation of a droplet. The impact of different droplet initial velocities is shown in Figure 3.7, for a droplet size of 200 μm , at an ambient temperature of 70 $^{\circ}\text{C}$ and air velocity 10 cm/sec. The results indicated that initial droplet velocity significantly affect the evaporation rate of the droplet mass. The droplet mass fraction evaporated decreased with an increase in the initial droplet velocity. With a decrease in the initial droplet falling velocity the evaporation rate would increase because the slower droplet would be suspended in the ambient environment for a longer duration and hence increase its chances of evaporation. The changes in initial droplet velocities were studied to help in determining the behavior

of droplet evaporation in the pilot scale evaporation column at different leachate flow rates.

3.6.6 Conclusion

The droplet evaporation modeling was based on many assumptions to help in the initial design of the evaporation column for example it was assumed that the droplet was evaporating in a infinite dry medium, which may not coincide with the practical scenarios. The results of the droplet modeling helped in determining the depth of the evaporation column. The results of the model predicted the behavior of the droplet evaporating in the evaporation column at different operating conditions. The performance of the evaporation column was estimated on the basis of fraction of droplet evaporated at different evaporator settings. These results ranked the importance of the process variables as droplet size, droplet falling velocity, evaporator temperature and air velocity (with droplet size being the most important variable and air velocity being the least). These results helped in selection of the process variables for the pilot scale evaporator unit and identified the possible impacts of these variables.

Figure 3.1: Process flow diagram

Figure 3.2: Temperature and mass fraction distribution at droplet phase interface

Figure 3.3: Droplet interval position.

Figure 3.4: Effect of droplet size on droplet evaporation

Figure 3.5: Effect of temperature on droplet evaporation

Figure 3.6: Effect of air velocity on droplet evaporation

Figure 3.7: Effect of droplet falling velocity on droplet evaporation

Chapter 4: Material and Methods

In this section, the major components of the evaporator pilot plant are discussed along with their control operations. The experimental plan that was employed to assess the impact of process variables on the performance of the evaporator with clean water and with actual landfill leachate is described. Sampling procedures and analytical methods that were used to analyze the different process streams when leachate was employed are also presented.

4.1 Evaporator configuration

The pilot evaporation process is shown in Figure 4.1 and the detailed design of all the individual components is attached as Appendix A. A brief description of the components is provided in the following sections.

4.1.1 Combustor

The combustor was one of the major components of the evaporator and was designed specifically for this research work. The initial mass and energy calculations for the combustor were completed to calculate the fuel and airflow rates required, while developing the model for the system. The final combustor system was designed and integrated by Bosik Consultants Limited (BCL). A mechanical sketch of the combustor is provided in Figure A-4, Appendix A. The following are major components of the combustor.

4.1.1.1 Burner

The combustor incorporated a gas burner that was capable of continuous operation to achieve the different heat rates. A gas ignition control system was used to ignite and monitor the flame. This had a built-in safety device, where the gas inlet valve shuts off if the flame was no longer detected. The combustor had an insulated cylindrical design, fabricated from stainless steel. A thermocouple was installed above the burner, which provided data on the combustion gas temperature. It was integrated into the safety system to shut off the power to the burner should the maximum temperature be exceeded. In the event of a flameout, or any blockage of the combustion air supply, the burner would “lock out” in a safety condition.

4.1.1.2 Air supply

The combustion air was drawn by the burner blower and was controlled by a damper gate valve located at the air intake louver. The airflow was measured manually by inserting a hot wire anemometer probe (EXTECH, model 407123) into the air stream to determine air velocity. Given the air velocity, the volume of airflow was calculated as per the calculations presented in Appendix H.

4.1.1.3 Fuel gas supply

The fuel gas supply to the burner had a manual shut off valve, followed by a pressure-regulating valve that was set to adjust the flow of fuel and hence the temperature of the evaporator. The burner was designed for use of methane at a landfill site but the same burner was employed for propane, in the lab. This was accomplished by switching the gas

flow meter, from natural gas to propane. The gas flow was adjusted manually and the flow was measured with a glass tube rotameter (model 7400), from King Instrument Company. A correlation between energy generation and fuel flow rates for propane and natural gas was developed and is presented in Appendix H.

4.1.2 Feed tank

A leachate feed tank with a capacity of 32 litres and constructed of stainless steel was fabricated. The top 30 cm of the feed tank had dimensions of 30 cm x 30 cm. This broad cross section area had a conical bottom surface. The center of the bottom surface was connected to a square shaped (7.6 cm x 7.6 cm) 20 cm deep stainless steel vessel having a capacity of 1 litre. It was constructed in a way that in case of any precipitation of heavy particulates, in the feed tank, all the sediments could slide towards the center of the bottom surface and ultimately to the small vessel. During all the batch test runs the foot valve on the suction line was placed in this small vessel, so that small volumes of the leachate in the feed tank could be pumped through the system. It had a flat platform equivalent to the size of a load cell so that it could be placed properly on the load cell for uniform load distribution.

4.1.3 Feed reservoir

A 25.4 cm x 25.4 cm square shaped, 20 cm deep, stainless steel feed reservoir was used during the extended runs only. It had a capacity of about 12 litres. The purpose of this feed reservoir was to provide a continuous flow of leachate to the feed tank and to maintain it at the required level. The drainpipe from the feed reservoir had a float valve,

attached with a stainless steel float ball, placed in the feed tank to automatically control the flow of leachate from the feed reservoir to the feed tank. As the leachate was drained from the tank, the control valve allowed the supply of leachate from the reservoir to maintain its level, in the feed tank to the required level. The main objective of this arrangement was to provide continuous supply of the leachate to the system, during the extended runs to achieve a higher degree of concentrate in the feed tank.

4.1.4 Load cell

The load cell was used in all the leachate batch test runs to determine the leachate evaporation rates during the pilot scale evaporation processes. The load monitoring apparatus consisted of a load cell and a digital data displayer. A TEDEA-HUNTLEIGH high capacity off-center load cell model 1250 was selected with a capacity of 75 kg, an accuracy of 0.02% of full scale and a receiving platform of 60 cm x 60 cm. For data display a Tracker 240 series digital panel load indicator was used. The leachate feed tank was placed on the load cell to monitor the load of the liquid in the feed bucket. With evaporation of leachate, the weight lost in the feed tank was continuously monitored to calculate the evaporation rate.

4.1.5 Pump

A Grundfos vertical in-line multistage pump, model CRI 1s-8 was selected for the system. It was of stainless steel construction with 2.54 cm suction and discharge lines. The pump was mounted on the support structure to pump the leachate from the feed tank to the evaporation column. A foot valve was employed in the suction line, and located in

the bottom portion of the feed tank to maintain the prime of the pump. The flow to the evaporator column was controlled by a manual control valve that diverted a portion of the pump discharge back to the feed tank.

4.1.6 Condensate bucket

A one-litre graduated flask was placed on a platform attached to the support structure to collect the condensate leaving the heat exchanger. Once filled, it was emptied into a separate storage tank. There was a shut off valve on the pipe coming from the heat exchanger to stop the condensate flow.

4.1.7 Evaporation column

The evaporator was 1 m deep and had a diameter of 30 cm. Both ends of the column had 5 cm wide collars to attach the cone shaped caps and to fasten the evaporator to the support structure. The spray nozzle was located at the top of the evaporator, and was suspended from the heat exchanger. A 20 cm diameter circular deflector plate was placed above the opening from the burner to prevent any leachate droplets from falling directly into the burner and for uniform distribution of hot combustion gases in the evaporation column. All the material used in construction of the evaporation column was stainless steel.

The evaporation column was connected to the combustor through a block made of two inverted cones, as shown in figure 4.1. The cones had a 5 cm pipe in the center. The hot combustion gases flowed through this pipe before entering into the evaporation column.

The bottom cone under the evaporation column had a 2.54 cm diameter drainpipe to discharge liquid that was not evaporated, into the leachate feed tank for re-circulation. A thermocouple was placed in the top cone of the evaporation column to monitor the inside temperature. A 6.4 mm diameter sampling port was also provided in the top cone to collect gas phase samples of the evaporate stream.

4.1.8 Heat exchanger

The heat exchanger was 30 cm in diameter and 60 cm deep with 5 cm wide collars on each end and was placed above the evaporator column. The core consisted of a coil, comprising eight loops of 6.4 mm diameter stainless steel tubing. The length of each loop was approximately 1.4 m. These loops were tied with steel plates, which were anchored to the heat exchanger walls, to keep them in place. The leachate passed through the inside of the core prior to entering the evaporation column through the spray nozzle. The hot gas that exited the evaporator column passed on the outside of the core.

The bottom of the heat exchanger contained the same cone type structure, as described in the evaporator column section. The evaporate stream coming from the evaporation column passed through this section before entering to the heat exchanger. A deflector plate was placed above the entrance to evenly distribute the hot air stream entering the heat exchanger. A 2.54 cm diameter drainpipe was provided in the bottom cone to drain condensates to the condensate bucket.

A thermocouple was placed in the top cap to monitor the inside temperature of the heat exchanger. Two thermocouples were attached to the steel tubing, at the points where the liquid entered and left the heat exchanger, to monitor the temperature changes of the liquid through the heat exchanger.

4.1.9 Insulation

The evaporator column and heat exchanger surfaces were insulated with 7.6 cm thick Multiflex, mineral fiber insulation material. It had a maximum service temperature of 850°F and thermal conductivity of 0.31 Btu.in/(hr.ft².°F). It was wrapped around the entire surface of the evaporation column and heat exchanger with an aluminum foil-backed duct tape.

4.1.10 Gaskets

Ceramic paper gaskets with a thickness of 3.2 mm were placed between all metal-to-metal surface connections to avoid any heat loss and liquid leakage from the evaporation column.

4.1.11 Support structure

A three-legged mild steel (3.8 cm x 3.8 cm) structure was fabricated to maintain the pilot plant integrity. The tripod was braced at four locations. The collars of the heat exchanger and the evaporator column were attached to these bracings, in a manner to facilitate the changing of the spray nozzles. A console box containing all the power switches, thermocouples and load cell, was also attached to the support structure. All the equipment

including this support structure was placed on a mobile platform to facilitate the transportation of the unit.

4.1.12 Spray nozzles

Two types of spray nozzles were selected from two different suppliers. All the nozzles were stainless steel and were capable of providing different droplet size distributions at different pressure and flow conditions. Two nozzles (GSS10 and GSS6.5) were from the Spraying System Co. These two nozzles were full jet spray nozzles and had a 6.4 mm ($\frac{1}{4}$ inch) inlet connection port. The flow through these nozzles varies from 1.8 lit per minute to 13.6 lit per minute (0.47 gallons per minute to 3.6 gallons per minute) with a change in pressure from 5 to 150 psi. Spray nozzle GSS6.5 had finer drop sizes as compared to the GSS10. The spray angle for GSS6.5 had a range of 45° to 50° , whereas for GSS10 it was reported as 58° to 67° at different pressure conditions.

The other two nozzles L66 and L54 were obtained from BETE Nozzles. Both these nozzles had a hollow cone spray pattern, were made of 316-stainless steel and had a 6.4 mm ($\frac{1}{4}$ inch) connection port. The flow rate through these nozzles ranges from 0.5 to 14.5 lit per minute (0.14 to 3.84 gallons per minute) and they had a spray angle of 90° . Generally, nozzle L54 had finer droplets as compared to the L66 nozzles.

To determine the average droplet size for each nozzle, the pressure in the system was determined from a table of flow rates and pressures provided by the supplier. The droplet

sizes for each spray nozzle were also plotted against different pressure ranges, to find the drop size for a specific pressure or flow condition.

4.2 Experimental methods

The conceptual model of the evaporator indicated that droplet size; its initial falling velocity, the temperature and countercurrent airflow were the important variables that impact the droplet evaporation rate. In this study the use of different nozzles was employed to change the droplet size. It was difficult to measure the actual velocity of the falling droplets and hence this was not attempted in this research work. However, the liquid flow rate through the nozzle was varied, which would have an impact on the droplet velocity and its suspension time in the evaporation column. By adjusting fuel and excess air supply to the combustor the temperature and air velocity in the evaporation were controlled.

Different runs were performed to evaluate the performance of the evaporator, for all these variables. These runs were performed, outside the Civil Engineering laboratory, in order to avoid any contamination of the laboratory with the exhaust gases. The equipment was moved out before each run and placed back in the lab on completion of the run.

4.2.1 Pilot plant operations

The first step to initiate a test run was to prime the pump, which was carried out according to the standard priming procedure, prescribed in the pump manual. A 2.54 cm inverted pipe was placed on the suction line of the pump to facilitate the priming process.

The load cell monitor was connected to display the amount of liquid in the feed tank. The load cell was set on tare to display only the mass of liquid. The liquid was filled in the feed tank and feed bucket, as required, before the start of each run.

For each nozzle, the liquid flow through the system was measured, before ignition of the combustor. For this the drainpipe from the evaporator was diverted to an 18 litre graduated glass bottle and its fill time was recorded using a stopwatch. The flow through the system was adjusted by the control valve, which was graduated in quarter revolutions.

The second step for each day of operation was ignition of the combustor. For this the propane gas cylinder, which was secured in a safe area, was connected to the fuel supply line of the combustor. Initially, the air and fuel supply valves were completely opened. It took approximately 30 seconds of purge period to flame the burner, once the ignition switch was turned on. The temperature of the combustion flue gases was controlled manually by adjusting the fuel flow valve or the air supply valve. The temperature control gauge for the combustor was set for the target temperature. The temperature of the combustor gases was directly noted from this temperature gauge. The air temperatures in the evaporation column and heat exchanger, as well as the liquid temperatures before and after the heat exchanger were checked manually from each thermocouple.

After the completion of these initial steps, the system was operated until the required temperature of the combustor was achieved and the temperature of the evaporation

column and heat exchanger were steady. Generally it took about 15 to 30 minutes to achieve these stable conditions. All the temperature readings were continuously monitored at constant intervals, normally after 30 minutes. The run was continued until steady state evaporation rates were determined for the particular condition of the test run. Test runs were conducted for water and leachate as discussed below.

4.3 Experimental plan

The evaporator was initially operated with tap water to evaluate the impact of the control variables on evaporation rates. Based on an evaluation of the water test runs important variables were selected to conduct runs for the leachate from Trail Road landfill site.

4.3.1 Water runs

All the water test runs were batch runs, in which a measured quantity of water was added in the feed tank. After each 30-minute interval, water was added to the feed bucket to return its volume to the initial value. The value of the water that was added was recorded and this allowed the rate of evaporation to be estimated. Thirteen water test runs were conducted during November and December 2002, for different conditions. One variable was changed in each run, keeping the other values constant. The following section describes the four variables that were evaluated in the water test runs.

4.3.1.1 Spray nozzles

Four test runs were conducted for the following four types of nozzles.

- a) G-SS-10 (Spraying System Co.)

- b) G-SS-6.5 (Spraying System Co.)
- c) L66-316SS (BETE)
- d) L54-316SS (BETE)

The flow through these nozzles was calculated when the control valve on the return line was completely closed, and hence all the flow was passing through the spray nozzle. For all these runs the combustor temperature was maintained around 800°C and the air supply valve was completely open. The evaporation rate and condensate generation rates were recorded for each run along with temperatures of combustor, evaporator, heat exchanger and liquid, before and after heat exchanger.

4.3.1.2 Water flow rates

A second set of runs were performed with the four types of nozzles but the water flow through the system was reduced, by opening the control valve on the return line to half. The flow through the system was measured for each nozzle. Evaporation data was collected under the same conditions of combustor temperature and airflow as discussed in section 4.3.1.1.

4.3.1.3 Airflow

The pilot plant was operated with the gate valve on the air at three different settings (completely open, 2/3 open and 1/3 open). The combustor temperature was set at 800°C and water flow was at full-flow through the system. An L66 type of nozzle was used for all these test runs.

4.3.1.4 Fuel flow

By adjusting the fuel flow to the combustor the evaporator was run for four combustor temperature conditions with target values of 400 °C, 600 °C, 800 °C and 1000 °C. All these runs were conducted with the L66 nozzle, and the air and water flow rates were kept at full flow conditions.

4.3.2 Leachate runs

The water evaporation data indicated that the type of spray nozzle and the temperature were important variables. These parameters were selected to study their impact on the leachate evaporation. In addition, the impact of leachate pH was considered as an important parameter to study, since it would impact the behavior of substances such as ammonia and volatile acids in the evaporator. All these runs were conducted when water and airflow were set at full flow conditions.

For a leachate run, fresh leachate from the Trail Road landfill site was transported on a daily basis. Nalgene HDPE bottles with a volume of 35 litres were used to haul the leachate. The bottles were completely filled with the leachate and capped to avoid loss of any volatiles. Seven batch type runs and one extended run were conducted in January and February 2003.

4.3.2.1 Batch runs

In each batch run, the feed tank was filled with about 30 litres of leachate. The run was continued until about 3 litres of the leachate was left as concentrate, in the feed tank. The

load cell readings were continuously taken at 30-minute intervals, to calculate the evaporation rate. The temperature of the combustor, evaporation column, heat exchanger, and leachate before and after the heat exchanger were also recorded at the same time.

Leachate evaporation rates were calculated using all four-nozzle types, as listed in section 4.3.1.1. To check the performance of the evaporator with these different nozzles the target combustor temperature was set at 800°C.

The effect of combustor temperature on leachate evaporation was evaluated at 600 °C, 800 °C and 1000 °C using the L66 nozzle. Two runs were performed with pH adjustment of the feed leachate to values of 5.5 and 3.5. The leachate pH was adjusted with hydrochloric acid, in the lab. These two runs were also conducted with the L66 nozzle, at 800 °C.

4.3.2.2 Extended run

In the extended run, the feed tank and feed bucket were both filled with leachate. The float valve was adjusted to maintain a 30-litre leachate volume in the feed tank. The level of the leachate in the feed bucket was continuously monitored and a measured quantity of leachate was added to it on regular intervals. The continuous run was conducted with the GSS6.5 nozzle at 800°C combustor temperature.

The objective of the continuous run was to identify any serious problems in the system, like clogging of the nozzle, if the system were run for an extended period of time. This

also permitted an evaluation of the impact of a higher degree of concentration of the concentrate on process performance. The extended run was conducted for about 22 hours over a four day period and approximately 300 litres of leachate was evaporated in this period. Each day the system was started with the same leachate that was left in the feed tank, from the previous evening. Concentrate was not removed from the system in this run. The run was continued until 300 litres of leachate were concentrated to approximately 30 litres.

4.4 Sampling procedure

Liquid and gas phase samples were collected for each leachate run. Liquid samples of the raw leachate, the concentrate stream and condensate stream were collected, whereas gas samples were collected from the evaporate stream. Details of the sampling procedures are provided in the following sections.

4.4.1 Batch runs

For each leachate batch run, three sets of liquid samples were collected. The raw leachate sample was collected from the leachate bottle, before the start of the run, as the fresh leachate arrived at the lab. The concentrate from the feed tank was drained in an empty bucket, at the end of the run, and a composite of the condensate stream was collected in a separate container for the entire run. The gas phase samples were collected during the initial half hour of each run, and then on one-hour intervals.

4.4.2 Extended runs

For the extended run, raw leachate samples were collected from each fresh leachate sample, whereas the concentrate and condensate samples were collected after the evaporation of every 100 litres. The gas phase samples during a continuous run were collected every eight hours.

4.4.3 Sample preparation

Liquid samples for pH, TSS, COD, Cl and NH_4 for all three waste streams were collected in completely filled 1 litre glass bottles, that were covered with Teflon lids and placed in a refrigerator at 4°C until further analysis. These analyses were conducted within 48 hours of sample collection. BTEX samples were collected in 40 ml amber colored glass vials. These vials were filled completely, sealed with Teflon septa and capped, so that no air bubbles remained in the bottle to avoid any volatilization of BTEX compounds, and placed in the refrigerator at 4°C . Before conducting BTEX analysis in the lab, these samples were filtered through $0.45\ \mu\text{m}$ Durapore membrane filters and filled in 1.5ml glass vials, sealed with Teflon septa and capped.

Samples for TKN, B, Be and H_2S were sent to an external lab (Caduceon Environmental Labs) for analysis. Samples were collected in 200 ml plastic bottles, provided by the lab. All these samples were preserved with 1ml of sulfuric acid solution, whereas about 0.3 ml of zinc acetate solution was added to the H_2S samples. These samples were placed in the refrigerator before they were sent to the lab.

The exhaust gases were analyzed for NH_3 and H_2S on site, during the run, using gas detector tubes. Gas samples from the evaporate stream were collected for BTEX in 20.3 litre, Cole-Parmer, Dual-valve Tedlar bags. These bags have both on/off and septum valves. The tedlar bags were filled from a sampling port provided at the top of the evaporation column.

4.5 Analytical methods

4.5.1 pH

The pH of all the samples was monitored using a Corning pH meter model 340, with a temperature operating range of -5 to 105°C and relative accuracy ± 0.01 . The meter was coupled with an Accumet, pH electrode (model 13-620-108).

4.5.2 Total Suspended Solids (TSS)

Method 2540 D, in the Standard Methods for Examination of Water and Wastewater (APHA, AWWA, WEF, 1992) was used to analyze the total suspended solids of the liquid samples. Whatman filter papers (model 934-AH) with $1.5\ \mu\text{m}$ pore size and 47mm diameter, were used to filter the suspended solids.

4.5.3 Chemical Oxygen Demand (COD)

The chemical oxygen demand (COD) of the samples was analyzed according to the procedures described in Standard Methods for Examination of Water and Wastewater (APHA, AWWA, WEF, 1992) using the Closed Reflux, Colorimetric Method (method

number 5220 D). A standard calibration curve was prepared according to the procedures prescribed in the method, for standard COD concentrations of 0, 200, 400, 600, 800 and 1000 mg/L

4.5.4 Total Kjeldahl Nitrogen (TKN)

The liquid samples for analysis of TKN were preserved, stored in the refrigerator and sent to Caduceon Environmental Labs, on a daily basis.

4.5.5 Ammonia (NH₃)

For liquid phase ammonia, a VWR, model 14002-794, Ammonia electrode was used. It has a concentration range of 0.01 to 14000 ppm, a temperature range 0 to 50°C and reproducibility of $\pm 2\%$. The pH of the sample was adjusted to 11, before each analysis by adding Ionic Strength Adjustor (ISA) solution and stirring continuously, as prescribed in the manual. The electrode was attached to an Orion model 420A meter to display the ammonia readings. Ammonia standard solutions were prepared following the procedures in Standard Methods (APHA, AWWA, WEF, 1992), section 4500-NH₃ G. A standard calibration curve was prepared for NH₃ using the standard concentrations of 0.1, 1, 10, 100 and 1000 ppm.

Gas phase ammonia concentrations were measured using Gastec No.3M, Ammonia Middle Range Detector Tubes that were installed in a Gastec precision pump. The detector tubes had a range of 10 to 1000 ppm.

4.5.6 Chlorides (Cl)

A Lachat (model IC5000) Ion Chromatograph (IC), was used for analyzing the chloride concentration in the samples. Quikchem Method 10-510-00-1-A, for determination of inorganic anions in water and solid extracts, was employed using an Anion Analytical Column, QS-A1 from Zellweger Analytics. The following conditions were set for the IC for analysis of the samples.

Eluent	2.2 mM NaHCO ₃ + 2.8 mM Na ₂ CO ₃
Regenerant	0.25 M sulfuric acid
Sample loop	100 µL

A calibration curve for Cl⁻ was prepared from a purchased standard solution with a concentration of 50 mg/l. Different dilution ranges of 10, 20, 30, 40 and 50 mg/l were analyzed to develop the calibration curve for the IC.

4.5.7 Toluene and m/p-xylene

A Varian Chrompack CP 3800 gas chromatograph (GC) was used for toluene and m/p-xylene analysis. A 15 m long, 0.32 mm inner diameter J&W Scientific DB-1 capillary column and a flame ionization detector (FID) were employed. The gas phase toluene and m/p-xylene samples were analyzed manually using a Supelco Solid Phase Micro Extraction (SPME) fiber holder for manual use (model 57330-U), whereas the liquid toluene and m/p-xylene samples were injected by a Varian Chrompack 8200 autosampler, using a 50/30 µm DVB/Carbonex/PDMS Stableflex SPME fiber assembly. The Autosampler and GC operated according to the following conditions;

Adsorb Time	1 minute (with agitation)
Desorb Time	5 minutes
Injector Temperature	250°C
Initial Oven Temperature	30 °C
Initial Hold Time	5 minutes
Rate Increase	10 °C/minutes
Final Temperature	80 °C
Final Hold Time	1 minute
Detector Temperature	300 °C
Column Gases	Helium (carrier) at 2.0 mL/min
Detector Gases	Hydrogen (flame) at 30 mL/min
	Air (flame) at 300 mL/min
	Nitrogen (makeup) at 28 mL/min

To prepare the calibration curve for liquid phase toluene and m/p-xylene, five 40 ml glass vials were partially filled with a solution of 0.1% methanol in water. Different amounts of a 1000 µg/ml standard BTEX solution were added to make up final concentrations of 10, 50, 100, 250, 500 mg BTEX per litre, in these vials. The vials were immediately topped with the methanol solution, sealed with Teflon septa and capped. To prepare the gas phase toluene and m/p-xylene calibration curves, 250 ml glass vials with Teflon septa and caps were used. The same 1000 µg/ml BTEX standard solutions were injected in the bottle to prepare different gas phase concentrations in these vials.

4.5.8 Hydrogen Sulfide (H₂S)

Hydrogen Sulfide Low Range Detector Tubes, Gastec No.4L, were used to measure the gas phase H₂S. A Gastec precision pump was used to collect the air samples from the

exhaust gases. The liquid phase H₂S samples were sent to Caduceon Environmental Labs on a daily basis.

4.5.9 Boron and Barium

Caduceon Environmental Labs conducted analysis for Boron and Barium. The liquid samples were preserved and sent to the lab on a daily basis.

Figure 4.1: Pilot scale evaporation unit

Chapter 5: Experimental Results and Analysis

The experimental results and analysis of the pilot scale evaporation plant are discussed in this section. Initially, results of the water test runs are discussed to evaluate performance of the evaporator unit. Secondly, the results of the leachate evaporation runs are discussed and compared with the water test runs. Analyses for the target contaminants in the leachate are also discussed in detail along with the impacts of different process variables on the fate of these contaminants.

5.1 Water runs

Thirteen water test runs were conducted for four sets of variables, as discussed in section 4.3. The results of the water test runs are provided in Tables B-1 to B-13 of Appendix B. The impact of nozzle type, water flow rate, airflow rate and fuel flow rate on the operating temperature of the evaporation column and heat exchanger were studied to characterize the performance of the pilot scale evaporator unit. The evaporate and condensate generation rates were determined for each test condition. The performance of the heat exchanger was also examined by measuring the temperature of the water, before and after leaving the heat exchanger.

5.1.1 Spray nozzles

The impact of different nozzle types on the performance of the evaporation column was studied. Four different types of spray nozzles were used in different experimental runs with clean water. All these runs were conducted at constant combustor temperature and

airflow conditions. The main objective of changing the spray nozzle was to vary the size of the falling droplets. The effect of nozzle type on water flow rates, evaporate and condensate generation rates were also observed.

Before conducting the water test runs, the spray pattern from all four nozzles was observed. To accomplish this the complete spraying system, which included the pump, the entire 0.64 cm stainless steel tubing and the spray nozzle, was setup outside the evaporation column. Visual observations that were made during the testing of the spray nozzles revealed that the spray pattern from nozzle GSS10 was comparable to a moderate rain, while nozzles GSS6.5 and L66 generated a pattern like a light rain, and nozzle L54 produced misty rain droplets.

Another observation made during the testing of these nozzles was that the nozzles from the two different suppliers had different spray angles, resulting in different spray cone sizes. The L66 and L54 nozzles had wider spray cones than the other nozzles, which would result in more direct contact of water droplets with the walls of the evaporation column. It was estimated that about 30% of the water droplets would have been impinging on the evaporator walls.

5.1.1.1 Effect of spray nozzle on droplet size

For each spray nozzle the water flow rate was measured and the measured values are given in Table 5.1. The manufacturers nomographs were used to estimate the droplet sizes. To accomplish this, the water pressure in the system was estimated from the flow

rate and pressure tables, provided by the supplier. In the second stage the droplet sizes were estimated from the graphs for droplet size versus pressure. The manufacturers graphs are attached as Appendix I. The estimated droplet sizes for the different nozzles are given in Table 5.1.

The droplet sizes reported for two different sets of nozzles from two separate suppliers had a wide range. The droplet sizes for the GSS10 and GSS6.5 nozzle types at flow rates of 3.86 and 3.45 lit/min were 1550 and 1300 μm , respectively. Whereas the droplet sizes for the L66 and L54 type of nozzles at flow rates of 2.84 and 2.45 lit/min were estimated as 110 and 75 μm , respectively. The droplet sizes reported for two different types of nozzles GSS6.5 and L66 by two suppliers had wide difference, whereas the results of visual observations revealed that these two nozzles had no difference in the droplet size, this aspect was not justified in this research.

5.1.1.2 Effect on evaporate and condensate generation

The effect of nozzle type on evaporate and condensate generation rates, resulting mean hourly evaporate and condensate generation rates (and two standard deviations) for each nozzle type are illustrated in Figure 5.1. A comparison of the results of nozzles GSS10 and GSS6.5 indicated that the evaporate and condensate generation rates of the latter was higher than the former. Similarly, the evaporate and condensate generation rates of nozzle L54 were higher than nozzle L66. These results indicated that the nozzles with fine droplet sizes had higher rates of evaporation and condensation. The fine droplets have

Table 5.1: Effect of nozzle on droplet size, flow rate, evaporator efficiency and temperatures in evaporator and heat exchanger

Run	Nozzle Type	Droplet size (μm)	Flow rate (lit/min)	Energy/lit. evaporate (kJ/lit)	Temperature ($^{\circ}\text{C}$)			Saturation of exhaust air (%)
					Evaporator	Heat exchanger	Water before heat exch.	Water after heat exch.
WR-2	GSS10	1550	3.9	2882	69.3	68.3	61.9	63.2
WR-4	GSS6.5	1300	3.5	2593	60.8	59.6	56.6	57.8
WR-8	L66	110	2.8	2602	60.7	59.6	56.8	57.9
WR-6	L54	75	2.5	2493	52.7	50.5	47.5	48.6

Table 5.2: Effect of water flow rate on evaporator efficiency and temperatures in evaporator and heat exchanger

Run	Nozzle Type	Flow control valve	Flow rate (lit/min)	Energy/lit. evaporate (kJ/lit)	Temperature ($^{\circ}\text{C}$)			Saturation of exhaust air (%)
					Evaporator	Heat exchanger	Water before heat exch.	Water after heat exch.
WR-1	GSS10	Open	3.5	3017	69.6	68.9	62.2	63.3
WR-2	GSS10	Closed	3.9	2882	69.3	68.3	61.9	63.2
WR-3	GSS6.5	Open	3.2	2740	60.6	59.6	56.1	56.9
WR-4	GSS6.5	Closed	3.5	2593	60.8	59.6	56.6	57.8
WR-7	L66	Open	2.4	2759	61.2	60.1	57.0	58.0
WR-8	L66	Closed	2.8	2602	60.7	59.6	56.8	57.9
WR-5	L54	Open	2.3	2683	52.9	50.9	48.0	49.0
WR-6	L54	Closed	2.5	2493	52.7	50.5	47.5	48.6

more surface area available for the transfer of heat from the gas phase to the liquid phase, this could be one of the main factor that increased the mass transfer from the surface of the water droplet.

Calculations were made to evaluate the saturation level of the exhaust air stream during the process runs with different spray nozzles. For this saturation vapor pressure of the exhaust air at respective evaporator temperature was calculated to determine the theoretical amount of water in the exhaust air. This water was compared with the actual amount of water evaporated in the evaporation column to determine the percentage saturation of the exhaust air. The results of the percentage saturation of the exhaust air stream are shown in Table 5.1. The results indicated that the exhaust air was saturated more than 100 percent during all the process runs except when the GSS10 nozzle was used. Therefore, it was difficult to compare the results of different types of nozzles from different suppliers on the basis of droplet size only. However, the trends between the two spray nozzles from the same supplier could be compared.

It is believed that some of the water droplets were carried over to the exhaust air without evaporating, due to high air velocity in the evaporation column. This could have resulted in an increase in the percentage of the water in the exhaust air more than 100 percent. The results also show that the fine droplet size nozzle had a high saturation percentage as compared to the large droplet size nozzles. This was likely observed because fine droplets had a higher likely hood of being carried over to the exhaust air. The saturation percentage of the exhaust air was calculated on the basis of the evaporator temperature,

which was measured using a thermocouple only at one point close to the top portion of the evaporation column. This temperature may not truly represent the exact temperature of the complete exhaust air stream. This could also be one of the reasons for more than 100 percentage saturation of the exhaust air.

The energy required to evaporate one litre of water was calculated for each nozzle type and is given in Table 5.1. Only the energy associated with the combustion of fuel was used for these calculations. The energy utilized by the pump, used for pumping the water in the evaporation column was not considered in these calculations. These results indicated that the fine droplet nozzles are more energy efficient. The energy required to evaporate a litre of water was decreased from 2882 to 2593 and from 2602 to 2493 kJ/lit with nozzles GSS10, GSS6.5, L66 and L54, respectively.

5.1.1.3 Effect on evaporator and heat exchanger temperatures

The operating temperatures inside the evaporator and heat exchanger were monitored during the experimental runs with different nozzle types. The average hourly operating temperatures are given in Table 5.1. It was observed that the nozzles with small droplet size had a lower temperature inside the evaporator as compared to the nozzles with larger droplet sizes. The small sized droplets had a high surface area and transferred more heat to the liquid phase resulting in lower ambient temperatures.

The water temperature entering and leaving the heat exchanger were also monitored to evaluate the performance of the heat exchanger. The average hourly water temperatures,

at both stages, are presented in Table 5.1. An overall temperature difference of 1.1 to 1.2 °C was observed in the heat exchanger during the experimental runs using all four nozzles. It was found that the water was already preheated before entering in the heat exchanger. Most of the heat transfer to the liquid took place inside the evaporation column, which was recycled continuously.

5.1.2 Water flow rates

Each nozzle was operated at two different water flow conditions, keeping the combustor temperature and airflow constant. In condition one, the flow control valve was open partially, allowing some of the flow to pass through the return line, whereas in the second condition the control valve was closed completely, so that all the water would flow through the nozzle only, as explained in section 4.3.1.2. The flow rates were measured for all four-spray nozzles during eight different process runs and are given in Table 5.2.

5.1.2.1 Effect on evaporate and condensate generation

Figure 5.2 compares the mean hourly evaporate and condensate generation rates along with two standard deviation values, during the eight process runs, using all four-nozzle types. While comparing the results of each nozzle type for two flow conditions it was clearly indicated that the increase in the flow rate through the system increased the evaporate and condensate generation rates. It was evident that the increased quantity of the water flowing through the system improved the heat and mass transfer across the water surface. The increase in flow would increase pressure in the system and reduce the droplet size and hence increase the evaporation rates.

The saturation level of the evaporate air stream was calculated for the process runs at the different water flow rates for the respective temperatures in the evaporation column. The results are given in Table 5.2. These results indicated that the evaporate air stream was more than 100 percent saturated during all the process runs except when the GSS10 nozzle was used. The fractional saturation of the evaporate air stream, when the GSS10 nozzle was used, increased from 56 to 61 percent when the flow rate was increased from 3.5 to 3.9 lit/min, respectively. This shows that saturation state of the air stream had a significant impact on the evaporation of water. With the increase in the evaporation rate, at high water flow rates, the saturation level of the air stream was also increased.

The energy required to evaporate a litre of water was calculated and is given in Table 5.2. The energy utilized by the water pump was not considered in these calculations. The comparison of nozzle GSS10 for two water flow conditions, 3.49 and 3.86 lit/min, showed that the energy requirement decreased from 3017 to 2882 kJ/lit, respectively. Similar trends were observed for all the other nozzles in that an increase in the flow rate decreased the energy requirement and hence increased the efficiency of the evaporator.

5.1.2.2 Effect on evaporator and heat exchanger temperatures

The operating temperatures of the evaporation column and heat exchanger for all four nozzles for different water flow conditions were monitored and mean hourly temperatures are given in Table 5.2. The results indicated that the increase in water flow rates decreased the temperature inside the evaporator and heat exchanger. The increase in

flow rate increased the heat and mass transfer rates and thus decreased the ambient temperatures inside the evaporator and heat exchanger.

The average hourly temperatures of the water entering and leaving the heat exchanger were recorded and are also given in Table 5.2. The trends observed in the temperature differences of water at different flow rates show that the temperature difference increased with the flow rate. The results suggest that at low water flow rates through the system the temperature of the evaporation column was increased, resulting in an increase in the temperature of the preheated water, which reduced the temperature difference between the water entering and leaving the heat exchanger.

5.1.3 Airflow rates

As discussed in section 4.3.1.3, the pilot scale evaporator was operated for three airflow conditions during the water evaporation runs. The L66 type nozzle was used in all these runs keeping water flow and combustor temperatures constant. During the experimentation it was observed that with the change in the airflow control valve settings, the velocity of air did not change very significantly. A change in the air velocity from 148 to 130 m/min was observed when the control valve settings were changed from full flow mode to the 1/3 open position.

5.1.3.1 Effect on evaporate and condensate generation

The evaporate and condensate generation rates at different air velocities were measured during three different runs. The mean hourly evaporate and condensate generation rates

as a function of air velocity are depicted in Figure 5.3, along with two standard deviation values. It was observed that evaporate and condensate generation rates increased with the air velocity. This may have resulted from the increased drag force on the water droplet that resulted from an increase in the air velocity. Secondly, with the increase in the amount of dry air the driving force for mass transfer from the liquid to the gas phase was increased. Analysis of variance (ANOVA) was used at a 95% confidence interval level to compare the water evaporation rate at different airflows. The results of ANOVA are given in Table D-1 of Appendix D. These results indicated that the water evaporation rates were statistically different at different airflow conditions. This show that even a small change in the air velocity from 148 to 130 m/min, the water evaporation rates were statistically different. This meant that the airflow rate significantly impacted the evaporation of water in the evaporation column.

Calculations were made to evaluate the saturation level of the exhaust air stream during the process runs with different airflows. The saturation vapor pressures of the exhaust air at the respective evaporator temperatures were calculated to determine the theoretical amount of water in the exhaust air. This water was compared with the actual amount of water evaporated in the evaporation column to determine the percentage saturation of the exhaust air. The results of the percentage saturation of the exhaust air stream, calculated as theoretical amount of water on the basis saturation vapor pressure and actual water evaporated in the evaporation column are shown in Table 5.3. The results indicated that the saturation percentage of the exhaust air decreased with an increase in the air velocity.

Hence, increase in airflow is an important variable, which significantly affects the evaporation rates and an increase in airflow would enhance the process efficiency.

The energy required to evaporate a litre of water at different air velocities was calculated and given in Table 5.3. With a change in the air velocity from 148 to 130 m/min the energy requirements were increased from 2602 to 2750 kJ/lit. This indicated that the increase in air velocity increased the evaporator efficiency.

5.1.3.2 Effect on evaporator and heat exchanger temperatures

The mean hourly temperatures of the combustion gases coming out of combustor, inside operating temperatures of the evaporator and heat exchanger are given in Table 5.3. The results show that reducing the airflow to the combustor increased the temperature of the combustion gases, because at low air supply, the fuel to air supply ratio was increased.

The hourly mean temperatures of the water entering and leaving the heat exchanger are shown in Table 5.3. The data indicate that with an increase in the airflow the temperature difference in the water between these two stages was reduced. This was due to the low temperature difference between the evaporator and heat exchanger with an increase in the airflow.

5.1.4 Temperature

The pilot scale evaporator was operated at four different temperature conditions to evaluate the performance of the evaporator using clean water. Four water test runs, WR-

11, WR-12, WR-8 and WR-13 were conducted at combustor temperatures of 400, 600, 800 and 1000 °C, respectively. These were achieved by adjusting the fuel flow to the combustor. All these runs were conducted with the L66 nozzle with full airflow and water flow conditions. The effect of different fuel flow rates on the combustor temperatures, evaporation and condensation rates are described in this section.

5.1.4.1 Effect of fuel flow on combustor temperature

The temperature of the combustion gases coming out of the combustor was recorded for different fuel flow rates. The mean hourly temperatures of the combustion gases along with fuel flow rates are given in Table 5.4. The temperatures of the combustion gases were plotted against the fuel flow rates, as shown in Figure 5.4. It was observed that the combustion gases temperatures increased with the fuel flow rates. With an increase in the fuel flow rate from 0.20 to 0.65 m³/hr the mean hourly combustor temperature increased from 401 to 1005 °C. The same trends were observed as was predicted by the mass balance calculations for the combustor, as discussed in section 3.3.2.

5.1.4.2 Effect on evaporate and condensate generation

The mean hourly evaporate and condensate generation rates (and two standard deviations) were plotted against the evaporation column temperatures as shown in Figure 5.5. The plot shows that the evaporate and condensate generation rates increased with the evaporator temperatures. With the increase in the fuel flow rate the evaporator temperature was increased significantly, which resulted in an increase in the transfer of

Table 5.3: Effect of airflow on evaporator efficiency and temperatures in the evaporator and heat exchanger.

Run	Air velocity m/min	Combustor temp. (°C)	Energy/lit. evaporate (kJ/lit)	Temperature (°C)			Saturation of exhaust air (%)
				Evaporator	Heat exchanger	Water before heat exch.	Water after heat exch.
WR-8	148	803	2602	60.7	59.6	56.8	57.9
WR-9	133	815	2692	62.5	60.9	57.2	58.5
WR-10	130	820	2750	63.1	61.3	57.4	58.7

Table 5.4: Effect of fuel flow on evaporator efficiency and temperatures in combustor, evaporator and heat exchanger.

Run	Fuel flow (m ³ /hr)	Energy/lit. evaporate (kJ/lit)	Combustor	Temperature (°C)			Saturation of exhaust air (%)
			Evaporator	Heat exchanger	Water before heat exch.	Water after heat exch.	
WR-11	0.2	3855	401	31.0	29.8	30.5	163
WR-12	0.31	2836	602	42.0	39.2	40.1	180
WR-8	0.45	2602	803	59.6	56.8	57.9	113
WR-13	0.65	2712	1005	80.0	69.5	71.1	35

heat from the gas phase to the liquid phase and increased the evaporation rates. Analysis of variance was used at a 95% confidence interval level to compare the water evaporation rate at different evaporator temperatures. The results of ANOVA are given in Table D-2 of Appendix D. These results indicated that the water evaporation rates were statistically different at different evaporator temperatures. This shows that the evaporator temperature has a very significant impact on the water evaporation.

The saturation level of the evaporate air stream was calculated for the process runs at different fuel flow rates for the respective temperatures in the evaporation column. These results are given in Table 5.4. These results indicated that the evaporate air stream was more than 100 percent saturated during all the process runs except when the combustor temperature was set to 1000 °C. The fractional saturation of the evaporate air stream at 1000 °C was only 35 percent saturated.

Calculations were made to evaluate the energy utilized to evaporate one litre of water for the above mentioned different fuel flow rates, as given in Table 5.4. These energy calculations showed that at low fuel flow rates or low temperatures more energy was required to evaporate one litre of water. With an increase in the combustor temperatures from 400, 600, 800 to 1000 °C the energy requirements observed in this pilot scale evaporator were estimated to be 3855, 2836, 2602 and 2712 kJ, respectively. The theoretical amount of energy required to evaporate one litre of water at 20 °C was calculated as 2594 kJ.

It was observed that the pilot scale evaporator system gave best performance at a combustor temperature of 800 °C. With a further increase in the temperature from 800 to 1000 °C though the evaporation rate in the evaporation column increased, the energy efficiency of the evaporator was reduced. The reason understood for is that the evaporate air stream was changed from saturated to unsaturated condition at temperatures above 800 °C. This implies that the rate limiting mechanism at 1000 °C was changed. At 1000 °C the air was unsaturated and all the heat energy in the air stream was not utilized to evaporate the water droplet. It is also believed that at high temperature there might be more heat losses in the system, but these investigations were not conducted in detail during this research work.

5.1.4.3 Effect on evaporator and heat exchanger temperatures

The mean hourly temperatures of the evaporator and heat exchanger at different fuel flow conditions are given in Table 5.4. The fuel flow had a significant effect on the operating temperature of the evaporator and heat exchanger. The temperatures inside the evaporator and heat exchanger increased with the combustor temperature. The difference between the temperatures of evaporator and heat exchanger also increased with the combustor temperature. The mean hourly temperatures of the water entering and leaving the heat exchanger are also given in Table 5.4. An overall increase in the temperature of water was observed from 0.7 to 1.6 °C if the combustor temperature was increased from 400 to 1000 °C, respectively.

5.2 Comparison of water evaporation and model results

In this section, the experimental data collected during the water evaporation runs is compared with the evaporation behavior predicted in the evaporation model. Generally all the experimental data showed the same trends as was predicted in the model. However, the intensity of evaporation rates in both data's had different values against different process variables. In the evaporation model, four different process variables; droplet size, evaporator temperature, air velocity and initial droplet velocity, were studied. But during the experimental runs initial droplet velocity was not measured therefore performance of the evaporator was studied for only remaining three variables.

5.2.1 Droplet size

During the experiments four types of nozzles were used to produce different droplet sizes in the evaporation column. Two nozzles were selected from each of two different suppliers. All these nozzles had different flow rates. The two full jet spray nozzles, GSS10 and GSS6.5 that were from one supplier had droplet sizes of 1550 μm and 1350 μm at flow rates of 3.85 and 3.45 lit/min, respectively. The other two nozzles, L66 and L54, were from the other supplier and had droplet sizes of 110 μm and 75 μm at flow rates of 2.84 and 2.45 lit/min, respectively. The experimental runs of these four nozzles, at constant evaporator temperature and airflow conditions, had shown that the fractions of the water that were evaporated during the runs were about 5.8, 7.2, 8.7 and 10.6 percent for the droplet sizes of 1550, 1350, 110 and 75 μm , respectively.

The evaporation model, developed on the basis of evaporation of a single water droplet, was run for the same droplet sizes and at the same evaporator temperature and airflow conditions monitored during the experimental runs. The results of the evaporation modeling showed that the evaporation rate was very sensitive to the water droplet size. The fraction of the droplet that was evaporated ranged from 0.07 to 0.11 only, when droplet sizes of 1550 and 1350 μm were used, while it was predicted that droplets having sizes of 110 and 75 μm , were completely evaporated in the evaporation column.

The experimental results showed that the percentage of the mass of water evaporated increased when fine droplet nozzles were used; as predicted in the model. However, the values of the fraction of the water droplet evaporated in the evaporation column based on the model calculations were significantly different from the experimental results. It is difficult to compare the experimental results with the model on the basis of droplet sizes only. The droplet sizes reported for two different types of nozzles GSS6.5 and L66 were 1350 and 110 μm . As discussed in section 5.1.1, observations made on the basis of visual observations had revealed that the two nozzles had minimal differences in the droplet sizes, whereas the respective droplet sizes reported by two suppliers had a very wide difference.

As discussed in section 3.6.5, the velocity of the falling droplets was also one of the important variables, which strongly influence the evaporation rate of the droplet, but this was not estimated during the experimentation. In this study the different nozzles had different flow rates and spray patterns. GSS nozzles had high flow rates as compared to L

type nozzles, which could increase the evaporation rate for the GSS nozzles. L type nozzles had a wider spray pattern as compared to GSS nozzles. This results in more direct contact of the water droplets with the walls of the evaporation column, as discussed in section 5.1.1 and could decrease the evaporation rate when L type nozzles were used.

5.2.2 Temperature

Water evaporation rates in the pilot scale evaporator unit were monitored during the experimental runs for four different evaporator temperature conditions. These results were compared with the results of the water droplet evaporation model, keeping the evaporator temperature and airflow conditions same as were observed in the experimental runs. The comparison of the experimental results and model calculations are depicted in Figure 5.38. The results show that the trends in water evaporation rates observed during experimentation were similar to those predicted by the model. However the actual values of the evaporation rates observed during the experimentation were somewhat lower than the values calculated in the model. This was likely due to assumptions made in the model calculations, such as the initial falling velocity of the droplet and the saturation state of the air stream in the evaporation column.

5.2.3 Air velocity

The air control valve on the air supply line was used to change the airflow rate to the combustor, manually, which ultimately controlled the air velocity in the evaporation column. During the experimental runs velocity of air inside the evaporation column could not be changed significantly, as discussed in section 5.1.3. With the different settings of

the air control valve, air velocity inside the evaporation column was changed from 2.5, 2.3 to 2.2 cm/sec only. The predictions of the model did not show any significant change in the evaporation of the droplet at these air velocities. Therefore, with this small change in the air velocity no reasonable comparison could be prepared for the experimental and model results.

5.3 Leachate runs

Leachate runs were conducted with the leachate from the Trail road landfill site. Seven batch runs and one extended run were conducted to study the evaporation behavior of the actual landfill leachate, in the pilot scale evaporation unit. The results of the leachate test runs are provided in Table C-1 to C-8 of Appendix C. These results are discussed in the following sections.

5.3.1 Batch runs

The results of the water test runs, as discussed in section 5.1 demonstrated that the nozzle type and evaporator temperature were the main factors affecting the performance of the evaporator. The impacts of these two process variables were studied in the leachate runs. The impact of another important variable, pH of the influent leachate, was also studied during these leachate batch runs. During a batch run raw leachate was concentrated from 30 litres of raw leachate to 3 litres of concentrate. The remaining 27 litres was evaporated as water vapor. Some of these vapors were condensed in the heat exchanger and collected as a condensate.

The effect of nozzle type, temperature and pH on the evaporate and condensate generation rates were monitored. The effect of different process variables on operating temperatures inside the evaporation column and heat exchanger during the batch leachate runs were also studied and discussed in this section.

5.3.1.1 Effect of nozzle type

The evaporate and condensate generation rates were monitored for all four nozzle types. Nozzle GSS10, L66 and L56 were used during 3 leachate batch test runs (LR-7, LR-1 and LR-6, respectively) whereas nozzle GSS 6.5 was installed during the extended run LR-8. Flow rates from these nozzle types were measured during these test runs and found to be the same as monitored during the water test runs. The hourly averages of evaporate and condensate generation rates are given in Table 5.5. All these test runs were operated around 800 °C combustor temperatures and at full water and airflow conditions.

The water and leachate evaporation rates were compared using different types of spray nozzles. The comparison of the mean water and leachate evaporation rates are shown in Figure 5.6. In general, the trends in evaporation rates of the leachate matched with the results of the water runs. The fine droplet nozzles had higher evaporation rates in comparison to the large droplet nozzles. The data sets for water and leachate evaporation rates were statistically tested for differences between their means via a two-sample t-test with a two-tailed hypothesis, at a 95% confidence interval level. The results of the t-test are given in Table D-3 of Appendix D. These results indicated that the evaporation rates of the water and leachate were statistically identical for all nozzle types.

The inside temperatures of the evaporator and the heat exchanger along with the temperature of the water entering and leaving the heat exchanger were monitored during the leachate runs. The overall averages of these temperatures are given in Table 5.5. The results show the impact of different nozzle types on temperature were the same as discussed in the water test runs. Fine droplet sized nozzles resulted in decreased operating temperatures of the evaporator and heat exchanger.

5.3.1.2 Effect of combustor temperature

Leachate evaporation and condensate generation rates were measured for three combustor temperature conditions, which were set around 600, 800 and 1000 °C. All these runs were conducted using the L66 nozzle and with air and water at full flow rates. The resulting average evaporate and condensate generation rates are given in Table 5.6. These results indicate that the leachate evaporation rates increased with the temperature of the combustion gases, which coincided with the results of the water evaporation rates at different temperatures.

The comparison of mean hourly leachate and water evaporation rates for three different evaporator temperature conditions are shown in Figure 5.7. The water and leachate evaporation data sets were tested for differences between their means using a two-sample t-test with a two-tailed hypothesis, at a 95% confidence interval level. The results of the t-test are given in Table D-4 of Appendix D. These results indicated that the evaporation rates of the water and leachate were statistically identical for all the temperature conditions.

Data for operating temperatures of the evaporator and heat exchanger, and water entering and leaving the heat exchanger, during the leachate runs is given in Table 5.6. The overall trends in this data were similar to the results of the water test runs. The temperature difference between the evaporator and heat exchanger decreased with the decrease in the combustor temperature.

5.3.2 Extended run

An extended run was conducted to concentrate the contaminants in the raw leachate to a higher degree and to observe the performance of the pilot scale evaporator if the run was continued for an extended time period. In the leachate run, LR-8, the evaporator was operated for four days using the GSS6.5 nozzle. The fuel flow rate was adjusted to maintain a combustor temperature around 800 °C, whereas leachate and airflow rates were set at full flow conditions. The results of the overall average daily evaporate and condensate generation rates were observed as given in Table 5.7. The operating temperatures inside the evaporator, heat exchanger, leachate entering and leaving the heat exchanger were monitored on regular intervals. The averages of these temperatures on daily basis are also given in Table 5.7. The comparison of the results of average daily data did not show very significant changes for the evaporation rates and operating temperatures inside the evaporator and heat exchanger.

Table 5.5: Effect of nozzle on leachate evaporation rate and temperatures in the evaporator and heat exchanger

Run	Nozzle Type	Evaporate		Condensate (lit/hr)	Combustor	Temperature (°C)		
		(lit/hr)	(lit/hr)			Heat exchanger	Leach. before heat exch.	Leach. after heat exch.
LR-7	GSS10	13.1	13.1	0.9	802	75.3	74.2	68.5
LR-8	GSS6.5	14.3	14.3	1.7	800	60.2	59.0	56.1
LR-1	L66	14.2	14.2	1.6	801	60.5	59.4	56.3
LR-6	L54	14.7	14.7	2.2	800	52.9	50.9	48.0

Table 5.6: Effect of fuel flow on leachate evaporation rate and temperatures in the evaporator and heat exchanger

Run	Fuel flow (m ³ /hr)	Evaporate		Condensate (lit/hr)	Combustor	Temperature (°C)		
		(lit/hr)	(lit/hr)			Heat exchanger	Leach. before heat exch.	Leach. after heat exch.
LR-3	0.31	9.3	9.3	1.3	601	42.2	41.2	38.5
LR-1	0.45	14.2	14.2	1.6	801	60.5	59.4	56.3
LR-2	0.65	20.0	20.0	2.7	999	81.6	80.0	68.6

Table 5.7: Leachate evaporation rate and temperatures in the evaporator and heat exchanger during the extended run

Run	Evaporate (lit/hr)	Condensate		Combustor	Evaporator	Temperature (°C)		
		(lit/hr)	(lit/hr)			Heat exchanger	Leach. before heat exch.	Leach. after heat exch.
LR-8 (day)								
Day-1	14.3	1.7	1.7	800	60.2	59.0	56.0	57.2
Day-2	14.4	1.7	1.7	802	60.3	59.2	56.3	57.5
Day-3	14.4	1.7	1.7	802	60.2	59.1	56.2	57.4
Day-4	14.1	1.6	1.6	798	60.0	58.9	56.0	57.1
Average:	14.3	1.7	1.7	800	60.2	59.0	56.1	57.3

5.3.2.1 Problem observed during extended run

Over the period of the landfill leachate runs, a layer of scaling was observed on the walls of the evaporator. A layer of the concentrated scum was also deposited on the bottom of the evaporation column. Since the inside of the evaporator was not cleaned during these runs, the scale and scum layer thickened with the passage of time. This problem was identified during the extended leachate run when the combustor became clogged at the end of the fourth day of operation.

The combustor and evaporation column were dismantled to investigate the reason for clogging. During the inspection of the bottom cone of the evaporator it was observed that some scaling scraps were mixed with the bottom scum and blocked the opening of the drainpipe coming out of the evaporation column. The leachate had overflowed from the bottom cone of the evaporation column and entered in the combustor chamber. The combustor burner had a built-in safety control system and was stopped automatically.

5.4 Leachate analysis

Liquid and gas phase samples were collected for contaminant analysis during all the leachate experimental runs, as discussed in section 4.4. Liquid samples from the raw leachate, concentrate and condensate streams were analyzed for pH, TSS, COD, TKN, ammonia, chloride, Boron, Barium, H_2S , toluene and m/p-xylene, whereas, ammonia, H_2S , toluene and m/p-xylene were monitored in the air stream, for each leachate run. The target compounds were analyzed in the different waste streams according to the

procedures prescribed in section 4.5. The results of the analysis for each experimental run are found in Table E-1 to E-10 of Appendix E.

5.4.1 Data quality

5.4.1.1 Duplicate sample analysis

Measures were taken to ensure the highest possible data quality. Duplicate samples were analyzed for several parameters. Average values of the duplicates were calculated and used to represent the target contaminant in a particular waste stream. An analysis of variability was performed for the duplicated samples. The results of the duplicate analysis are found in Appendix F Table F-1 to F-6. The following formula was used to assess the variability between duplicate samples of a parameter.

$$\frac{|s_1 - s_2|}{\bar{x}}$$

Where s_1 and s_2 are sample values, $\bar{x}$ is mean of the two samples.

The results of the duplicate analysis indicated that the variability in the TSS results was less than 2 percent. The variability in the COD, NH_3 and chlorides was less than 1 percent. Generally variability of toluene and m/p-xylene was less than 2 percent except for one sample when the variability was 6 and 8 percent for toluene and m/p-xylene, respectively. It was considered that the variability in the duplicate samples was not very significant and mean values of the sample were used for data analysis.

5.4.1.2 Calibration

The concentrations of the COD, ammonia, chloride, toluene and m/p-xylene were estimated from calibration curves. The calibration curves were prepared from their respective standards as discussed in section 4.5. The coefficient of determination (R^2) values for these curves were very high indicating that nearly linear calibration curves had been achieved which would provide acceptable estimation of the desired concentrations. Table 5.8 presents the R^2 values for each parameter.

Table 5.8: Coefficient of determination (R^2) values for the calibration curves

Parameter	R^2 value
COD	0.977
NH ₃	0.999
Cl ⁻	0.999
Toluene (liquid phase)	0.988
Toluene (gas phase)	0.987
m/p-xylene (liquid phase)	0.981
m/p-xylene (gas phase)	0.998

5.4.1.3 Raw leachate characteristics

For each leachate run fresh leachate from the Trail Road landfill site was transported on daily basis to be used in the pilot scale evaporator. All the target components in the raw leachate were analyzed. The results of these analyses are given in Table 5.9. The maximum and minimum values, means, standard deviations and coefficient of variations for all the target contaminants in raw leachate are also given in Table 5.9 to compare the initial concentration of these contaminants in the raw leachate over the period of experimental runs. Generally, the coefficient of variation for all the components varies raw leachate was adjusted to 5.5 and 3.5 during experimental runs 4 and 5, respectively.

Table 5.9: Raw leachate characteristics

Parameter	units	Raw leachate concentrations										Maximum	Minimum	Mean	STDEV	Coefficient of variation
		Run 7								Run 8						
		Run 1	Run 2	Run 3	Run 4	Run 5	Run 6	Run 7	(sample 1 sample 2)							
pH	pH	7.55	7.24	7.5	5.5	3.5	7.54	7.28	7.59	7.6	7.6	3.5	6.8	1.4	0.21	
TSS	mg/l	27.5	36.8	26.7	24.6	36.8	40.8	35.4	29.8	34.3	40.8	24.6	32.5	5.5	0.17	
COD	mg/l	721	1090	1170	662	681	569	543	803	770	1170	543	778.7	217.0	0.28	
NH ₃	mg/l	640	548	551	453	433	544	421	582	589	640.04	420.88	528.9	76.3	0.14	
TKN	mg/l	807	664	682	725	708	604	506	1380	1270	1380	506	816.2	301.3	0.37	
Cl ⁻	mg/l	1896	1905	1819	1916	1966	1480	1550	1754	1660	1966	1480	1771.7	173.5	0.10	
B	mg/l	7.1	7.1	7.3	7.7	7.1	6.4	6.1	18.7	19.1	19.1	6.1	9.6	5.3	0.55	
Ba	mg/l	1.0	0.9	1.0	0.8	0.8	1.1	1.2	1.1	1.2	1.2	0.8	1.0	0.1	0.15	
H ₂ S	mg/l	1.6	6.0	0.2	0.1	0.1	1.2	0.8	0.4	0.9	6.0	0.1	1.2	1.8	1.49	
Toluene	µg/L	102.2	114.0	99.6	57.9	58.7	53.7	50.8	75.1	58.6	114.0	50.8	74.5	24.3	0.33	
m/p-xylene	µg/L	63.2	73.3	66.3	59.1	43.2	42.8	33.4	59.1	41.6	73.3	33.4	53.6	13.6	0.25	

The most significant variation in the raw leachate concentration was observed for the H_2S , the maximum and minimum values vary from 6 mg/lit to 0.05 mg/lit.

The concentrations of these target contaminants in concentrate or exhaust air stream is based on their initial concentration in the raw leachate. A large variation in the concentration of the raw leachate could strongly affect the results of the concentrate and exhaust air streams. However, a reasonable comparison could be made if the results of the concentrate and exhaust air streams are compared as mass fraction of their original concentration in the raw leachate.

5.4.2 Batch runs

The batch type leachate evaporation runs evaluated three process variables, nozzle type, temperature and pH adjustment of the influent leachate. Each process variable ran at three conditions. Three types of spray nozzles; GSS10 (droplet size 1550 μm), L66 (110 μm) and L54 (75 μm) were used during batch process runs; LR-7, LR-1 and LR-6, respectively. The combustor temperature was evaluated at 1000 °C, 800 °C and 600 °C in process runs LR-2, LR-1 and LR-3, respectively. The influent leachate pH was adjusted during LR-4 and LR-5 to values of 5.5 and 3.5, respectively. All the runs were conducted with full air and liquid flow conditions.

During the batch leachate runs, about 90 percent of the raw leachate was evaporated to attain 10 percent concentrate volume. Therefore the concentrations of the conservative

contaminants in the concentrate streams were expected to be concentrated by a factor of 10 times, as compared to the raw leachate.

The condensate was expected to be comparatively clean, as compared to the raw leachate. However, in most of the process runs the condensate also contained a significant amount of contaminants. It is believed that some of the leachate droplets were carried over to the exhaust air stream, without evaporating, due to the high air velocity in the evaporation column. Some of these leachate drops were trapped in the heat exchanger and entered the condensate collection bucket, which elevated the concentration of contaminants in the condensate stream.

Mass balances were prepared for each contaminant for all the batch test runs. The results of the mass balances are produced in Appendix G. The concentrations of a contaminant entering the evaporation unit, as raw leachate were compared with its concentration in the waste streams leaving the system, as concentrate, condensate or exhaust air. The total mass of contaminants in each stream was calculated by multiplying the volume of the stream with the concentration of the contaminant in that stream. The impact of each process variable on the target contaminant was studied. The behavior of all the target compounds for three process conditions were examined and discussed in this section.

5.4.2.1 pH

The pH of the raw leachate, concentrate and condensate streams were monitored during all the batch type leachate runs. The resulting pH for all streams is given in Table 5.10.

The results indicated that pH of the concentrate and condensate increased during all the leachate runs. The final pH of concentrate stream, during the process runs when the initial pH of the raw leachate was not adjusted increased by a value ranging from 1.1 to 1.5, whereas the final pH of condensate increased by 1.1 to 1.6 pH points. During experimental runs, LR-4 and LR-5, when the initial pH of the raw leachate was adjusted to 5.5 and 3.5, increases in the pH values of concentrate and condensate streams were observed by 1.28 to 2.36 points and 2.75 to 4.12 points, respectively.

Table 5.10: pH of different waste streams during batch runs

	pH values during the batch runs		
	Raw leachate	Concentrate	Condensate
Run 1	7.55	8.79	8.89
Run 2	7.24	8.73	8.91
Run 3	7.5	8.72	8.58
Run 4	5.5	6.78	8.25
Run 5	3.5	5.86	7.62
Run 6	7.54	8.66	8.79
Run 7	7.28	8.69	8.81

It is believed that volatilization of the weak organic acids resulted in an increase in the pH of the concentrate and condensate. Due to volatilization of the weak organic acids in the leachate the H^+ concentration in the concentrate could be reduced because more H^+ ions will associate with the ionic components of the acids to compensate the loss of volatile acids to achieve the equilibrium. The pH of the condensate was higher than the concentrate. It is believed that some of the ammonia, stripped off from the leachate during the evaporation process, was trapped in the condensate and increased the pH of the condensate stream.

5.4.2.2 Total suspended solids (TSS)

The TSS of the raw leachate, concentrate and condensate were analyzed for each test run. Mass balances were prepared to compare the TSS entering and leaving the system on the basis of mass fractions of TSS observed in the concentrate and condensate streams. The results of the mass balances on TSS during these process runs are given in Table G-1, Appendix G. The TSS results during different experimental runs and the impact of different process variables on TSS are discussed in the following section.

a) Effect of temperature

The impact of temperature on the TSS for three different evaporator temperatures 42, 61 and 82 °C was studied during experimental runs LR-3, LR-1 and LR-2, respectively. The mass fractions of TSS calculated in the concentrate and condensate streams were plotted against the evaporator operating temperature, as shown in Figure 5.8.

The results of the mass balances on TSS at different temperature conditions showed that about 47 to 63 percent of the TSS remained in the concentrate stream, whereas 4 to 6 percent were carried over to the condensate stream. The remaining 34 to 47 percent could not be accounted for in the waste streams leaving the system. The reason for the disappearance of the TSS could be: a) some of the TSS that was composed of volatile matter might have been volatilized; b) some of TSS might have been attached to the metal surface inside the evaporator and formed a scale layer or entered in the scum, which had been deposited in the bottom of the evaporator; c) some fine TSS particulates

might have flown away in the evaporate stream along with the other water vapors; and d) some of TSS might have been decomposed or solubilized.

Comparing the results at different temperatures (Figure 5.8) it was observed that the TSS mass fraction in the concentrate increased with a decrease in the temperature. The leachate runs conducted at low temperature had concentrated the TSS more than the runs conducted at high temperatures. This is likely due to the fact that with an increase in the temperature the air velocity inside the evaporation column would be increased and exert more drag force on the leachate droplets. The increased drag force would result in them being carried over to the exhaust air stream or increase their attachment to the wall of the evaporator. The increased carry over of leachate droplets to the air phase, at high temperatures may have reduced the TSS concentration in the concentrate, as compared to the runs at low temperature. The results of the condensate stream also showed that TSS concentration in the condensate increased with the evaporator temperatures.

b) Effect of nozzle

The effect of droplet size on TSS was analyzed, by using different spray nozzles, L54, L66 and GSS10 during three leachate runs LR-6, LR-1 and LR-7, at constant temperature, leachate flow and airflow conditions. The resulting mass fractions of TSS in the concentrate and condensate streams were plotted for different spray nozzles, as shown in Figure 5.9.

The results indicated that the TSS mass fraction in the concentrate and condensate streams changed with the nozzle types. The nozzles with large droplet sizes contained more TSS in the concentrate stream as compared to the nozzles with fine droplet sizes. As discussed above, some of the leachate droplets were moved up in the exhaust air along with evaporate stream due to the high air velocity in the evaporation column. Therefore, chances of fine droplets to go in the exhaust air were higher as compared to large droplets. The TSS results of the condensate stream had also shown that the mass fractions of TSS in the condensate were higher when fine drop size nozzles were used.

As discussed in section 5.2.1, the leachate evaporation rate increased with the temperature of the evaporation column. But the evaporation rate also increased when fine droplet nozzles were used, despite the decrease of temperature in the evaporation column. It is believed that when two runs were conducted at constant combustor temperature and airflow conditions, droplet size is the leading variable which controls the evaporation rate in the evaporator. This could also lead to an increase in the carry over rate of the droplets to the exhaust air stream.

5.4.2.3 COD

Mass balances for COD were prepared to determine the mass fraction of COD remaining in the concentrate and condensate streams, for each test run. Results of the mass balances are given in Table G-2, Appendix G. The impacts of different process variables on COD are discussed in this section.

a) Effect of temperature

Figure 5.10 shows the mass fractions of COD in the concentrate and condensate streams plotted against the respective evaporator temperatures. In general, the mass balances for COD indicated that COD concentration in the concentrate stream increased during the evaporation process. From the COD mass fractions in the concentrate stream, during the three test runs at different temperatures it was observed that about 31 to 24 percent of the total COD in the raw leachate was retained in the concentrate, whereas only 3 to 5 percent was observed in the condensate. The remaining 66 to 70 percent COD could not be accounted for in the concentrate or condensate streams. The disappearance of COD was likely due to the same reasons as discussed in the TSS section (5.3.2.2-a) in that some matter contributing to COD could be decomposed, volatilized, attached to the evaporator surface or carried over with the exhaust air stream.

The results in Figure 5.10 show the same trends as observed for the TSS in that the COD mass fraction in the concentrate increased with a decrease in the evaporator temperature, whereas the mass fraction in the condensate increased with temperature. This may have been due to the fact that the solubility of the volatile organic carbons (VOC) in liquids decreases as the temperature increases. At high temperatures VOCs volatilize quickly and thereby decrease the COD of the concentrate.

Henry's law gives the relationship between the partial pressure of the gas in the atmosphere above the liquid and the concentration of the gas in the liquid as provided in equation 5-1 (Tchobanoglous et al., 1993)

$$\frac{C_g}{C_s} = H_c \quad (\text{Eq 5-1})$$

Where C_g = concentration of VOC in the gas phase ($\mu\text{g}/\text{m}^3$)

C_s = saturation concentration of VOC in liquid ($\mu\text{g}/\text{m}^3$)

H_c = Henry's law constant.

Henry's law constant increases with the temperature (Tchobanoglous et al., 1993). Therefore with the increase in the temperature volatilization of the VOCs contributing to COD concentration was increased, resulting in a decreased COD in the concentrate stream and increasing its concentration in the condensate stream.

b) Effect of pH

The effect of pH on the COD concentrations in the different waste streams was studied for three different process runs (LR-5, LR-4 and LR-1), conducted with an influent raw leachate pH at 3.5, 5.5 and 7.5, respectively. All these runs were conducted with one nozzle and at a constant temperature, leachate flow and airflow. The final pH of the concentrate and condensate was increased during all these runs, as given in Table 5.10. The results for COD mass fraction in the concentrate and condensate streams are shown in Figure 5.11.

The results indicated that the COD mass fraction in the concentrate stream decreased with an increase in the pH. The COD concentration in the different waste streams depends on the initial pH of the influent leachate and the extent to which the COD is

composed of volatile organic acids (VOAs). The ions of VOAs are not volatile. At low pH values VOAs are changed into their non-ionic forms and increase the volatilization. Ionization of acetic acid is shown in equation 5-2.

The pKa of the acetic acid (CH_3COOH) at 25 °C is reported as 4.7 (Snoeyink, 1980). Therefore, at low pH (with the increase in H^+ ion concentration) the acetate ion will be converted into acetic acid, which has high volatility. Therefore, during low pH runs the COD mass fraction decreased in the concentrate but increased in the condensate.

c) Effect of nozzle type

The COD mass fraction in the concentrate and condensate streams, during the process runs was plotted against the respective nozzle types, as shown in Figure 5.12. All these runs were conducted at a constant combustor temperature. The results indicated that with the fine droplet nozzles the COD mass fraction in the concentrate decreased but increased in the condensate. The fine droplets have large surface area as compared to the big droplets, which increase the chances of volatilization of the VOCs from the droplet surface and hence reduce the COD mass fraction in the concentrate stream. Secondly, as discussed in section 5.3.2.2-b, the likelihood of carry over of the fine droplets to the condensate and exhaust air stream was increased when fine droplet nozzles were used.

5.4.2.4 TKN

The TKN analysis measures ammonia and organic nitrogen. Liquid samples from the raw leachate, concentrate and condensate streams were collected to analyze TKN and ammonia. The exhaust gases leaving the system were analyzed for ammonia only. The results of the analysis are available in Table E-1 to E-10 Appendix E. The results of the raw leachate samples indicated that the TKN values ranged from 506 to 807 mg/lit whereas the results for ammonia ranged from 421 to 640 mg/lit. This indicated that about 79 to 83 percent of the TKN was composed of ammonia. Mass balances for TKN and ammonia were prepared, and are discussed separately. The effects of different process variables on TKN are described in this section.

a) Effect of temperature

The TKN results, during three leachate runs, LR-3, LR-1 and LR-2 at evaporator temperatures of 42, 61 and 82 °C, respectively, were analyzed. The total mass of TKN entering the evaporator was calculated and compared with the total mass leaving the system. The mass balances on TKN are given in Table G-3, Appendix G, along with the mass fraction examined in different waste streams.

The results of the mass balances indicated that only a very small fraction of the TKN was found in the concentrate and condensate streams. The TKN mass fractions remaining in the concentrate and condensate were plotted against the respective temperatures in the evaporation column, as shown in Figure 5.13. These results indicated that with the increase in temperature the mass fraction of the TKN in the concentrate decreased, but

increased in the condensate. It is believed that some of the organic components contributing to TKN were decomposed at higher temperatures and most of the ammonia has been volatilized. As discussed in section 5.3.2.3-a, the Henry's law constant increases with temperature. Therefore volatilization of the ammonia would have been increased with the increase in temperature.

b) Effect of pH

The pH of the raw leachate was adjusted to 5.5 and 3.5 during two leachate runs, LR-4 and LR-5. These results were compared with LR-1, when the initial pH of the raw leachate was not changed. Mass balances on TKN were prepared on all the liquid waste streams entering and leaving the system. The mass fraction of TKN remaining in the different waste streams during these process runs were plotted against respective pH values, as shown in Figure 5.14.

The results of the mass balances indicated that with a decrease in pH of the raw leachate from 7.5 to 3.5 the TKN mass fraction remaining in the concentrate increased from 2 to 43 percent. The final pH of the concentrate and condensate streams increased during all the experimental runs. The final pH of concentrate during these three runs was increased to 8.8, 6.8 to 5.9, respectively. This means that at lower pH values high concentrations of TKN would be achieved in the concentrate stream. The reason for this is that at low pH, more H^+ ions are available, which could convert the NH_3 component of the total nitrogen into NH_4^+ ions, as shown in equation 5-3. The ammonium ions are more stable in the liquid phase than nitrogen as NH_3 .

5.4.2.5 Ammonia (NH_3 / NH_4^+)

The concentrations of ammonia in liquid and gas phases were analyzed for each leachate run. Mass balances for ammonia compared the total mass of ammonia entering the system as raw leachate with the total mass of ammonia leaving the system as concentrate, condensate or evaporate. The results of mass balances for ammonia during different leachate runs are presented in Table G-4 Appendix G. The effects of different process variables on ammonia transformation are discussed in this section.

a) Effect of temperature

The ammonia mass fractions in the concentrate, condensate and air streams at the three evaporator temperatures of 42, 61 and 82 °C, were plotted, as shown in Figure 5.15. All these runs were conducted without any pH adjustment of the raw leachate and at constant leachate and airflow rates.

The results of the mass balances indicated that only a very small fraction of the ammonia was found in the concentrate and condensate streams. The results showed that the liquid phase ammonia remaining in the concentrate was less than 1 percent, whereas, about 2 percent was trapped in the condensate stream. It is believed that the remaining ammonia was transformed into the gas phase and flown away in the evaporate stream. The results of the mass fraction of ammonia in concentrate and condensate showed the same trends as discussed for TKN, in section 5.3.2.4-a. With the increase in the temperature the mass

fraction of ammonia in the concentrate decreased but increased in the condensate. However, the values of these variations are not very significant.

The results of the gas phase ammonia analysis showed that during the high temperature runs the liquid ammonia was quickly dissipated into the gas phase as compared to the runs conducted at lower temperatures. During the low temperature run (at 42 °C) the gas phase ammonia concentration initially increased with the time and then finally decreased, whereas during the high temperature runs most of the ammonia was stripped of during the first 30 minutes of the run. Since the gas phase ammonia concentration was monitored as a grab sample at 30 minutes interval, it was difficult to calculate the actual gas phase ammonia for the entire run, especially when ammonia was observed in the exhaust air for a longer duration. The best mass balance closures were observed during the run when most of the ammonia was observed during the first 30 minutes of the run, which was during the high temperature run at 82 °C.

b) Effect of pH

The ammonia mass fraction in the concentrate, condensate and air streams, during three leachate runs (LR-4, LR-5 and LR-7) were plotted versus pH and are shown in Figure 5.16. The results of the mass balances indicated that with a decrease in pH from 7.5 to 3.5 the mass fraction of ammonia in the concentrate increased from 0 to 80 percent. This means that pH of the influent leachate had a very significant effect on the volatilization of ammonia. At low pH, more free H^+ ions are available, which changed un-ionized ammonia (NH_3) to the ionized ammonium ion (NH_4^+). Therefore at lower pH more NH_3

would be converted into NH_4^+ ions, which were more stable in the liquid phase. The resulting mass fraction of ammonia observed in the gas phase also showed that during the low pH runs, ammonia concentration in the exhaust air was decreased. The mass balance results showed recovery of ammonia little more than 100 percent. As discussed above, it was difficult to calculate the actual mass of ammonia in the gas phase. The gas phase ammonia was measured on the basis of grab air samples taken at every 30 minutes intervals, using Gastec tubes, which measured ammonia to an approximate value.

The results of the final pH of the concentrate and condensate streams (Table 5.10) indicated that the pH of concentrate and condensate increased during all the process runs. The results of the ammonia concentration in the exhaust air stream with time are given in Table E-9, Appendix E. During the process run, LR-1, when initial pH of the raw leachate was not adjusted the gas phase ammonia concentration was decrease with the time, because most of the ammonia was dissipated in the first hour of the run. During the process runs, RL-4 and LR-5, when initial pH of the raw leachate was adjusted to 5.5 and 3.5, respectively, most of the ammonia was remained in the concentrate. During these runs the ammonia concentration in the gas phase increased with the time, because pH of the leachate in concentrate also increased.

c) Effect of nozzle type

The effect of nozzle type on ammonia was monitored during three leachate runs, LR-6, LR-1 and LR-7, using nozzles, L54, L66 and GSS10, respectively. The impact of the

different nozzles on the mass fraction of ammonia remaining in the concentrate, condensate and air streams are shown in Figure 5.17.

The results of the mass balances on ammonia, while comparing different nozzles, had shown that less than 2 percent remained in the concentrate and condensate streams. It is believed that the remaining ammonia in the liquid phase was stripped off into the gas phase. The results of the mass fraction of ammonia in the concentrate showed that the ammonia concentration in the concentrate was reduced with the nozzles having fine drop sizes but these values are not very significant. This may have occurred because the fine droplets had more surface area as compared to large droplets, thereby increasing the chances of ammonia to escape from the liquid to the gas phase. The results of the gas phase ammonia (Table E-9, Appendix E) also showed that with the fine droplet nozzle L54 (LR-6) ammonia was quickly dissipated into the gas phase as compared to the large droplet nozzle GSS10 (LR-7).

5.4.2.6 Chloride (Cl)

The effect of different process variables on chloride behavior was studied for all batch type evaporation runs. Chloride mass balances during different leachate process runs are given in Table G-5 Appendix G. The results of these mass balances are discussed in this section.

a) Effect of temperature

The results of the mass balances for chloride in the different waste streams, during runs at different temperatures, indicated that about 60 to 73 percent of the chlorides were concentrated in the concentrate stream, and about 2 percent were trapped in the condensate stream. The remaining 25 to 38 percent could not be detected in these waste streams. It was assumed that some of the missing chloride was either lost in the exhaust gases in the form of leachate drops or may be attached to the wall of the evaporator or heat exchanger as a scale layer or may be contained in the scum, which was deposited at the bottom of the evaporation column.

To observe the impact of the temperature on the different waste streams the chloride mass fraction in the concentrate and condensate were plotted against the evaporator temperatures as shown in Figure 5.18. These results indicated that with an increase in the temperature, the mass fraction of the chlorides in the concentrate decreased. With an increase in the temperature, the air velocity inside the evaporation column would be increased and exert more drag force on the leachate droplets causing them to be carried over to the exhaust air stream or increasing their chances of being attached to the wall of the evaporator. The mass fraction in the condensate increased with the increase in temperature.

b) Effect of pH

The effect of pH on chloride behavior was determined for three process runs conducted at initial pH values from 7.5, 5.5 and 3.5. All of these runs were conducted with one nozzle

and at constant temperature, leachate and airflow rates. Mass balance calculations were made on the basis of total chlorides present in the raw leachate and chlorides added in the form of HCl, considering its normality as 5N. The results of the chloride mass fraction in the concentrate and condensate streams were plotted against respective pH values, as shown in Figure 5.19. The results of the mass balances show that with a decrease in the pH from 7.5 to 3.5 the percentage of the chloride mass retained in the concentrate decreased from 67 to 36 percent. It is believed that this low concentration of chloride mass in the concentrate, at low pH, was because of a poor confidence in the actual normality of the HCl, since an old HCl sample existing in the lab was used to adjust pH of the leachate. The results of the mass balances, considering the normality of HCl as 2N, showed that about 67 to 64 percent of the chlorides were contained in the concentrate during the runs when the initial pH of the raw leachate was adjusted to 5.5 and 3.5, respectively. The results of mass balances with a 2N HCl are comparable with the chloride concentrations in the concentrate that were observed in the other process runs. It is believed that the remaining 33 to 36 percent chlorides were lost as leachate droplets, which were either carried over to the exhaust air or deposited as scale or scum.

c) Effect of nozzle type

The mass fractions of chloride in the concentrate and condensate streams were plotted for the three nozzles (L54, L66 and GSS10), used in different leachate process runs (LR-6, LR-1 and LR-7), as shown in Figure 5.20. All these runs were conducted at constant pH and temperature.

The mass balances indicated that about 64 to 78 percent of the chloride mass remained in the concentrate stream, whereas about 5 to 1 percent appeared in the condensate stream. The remaining 31 to 21 percent could not be detected in these waste streams. The results from the Figure 5.20 indicated that with the use of fine droplet nozzles the chloride concentration in the concentrate decreased but increased in the condensate. It is believed that the escape of fine drops to the exhaust air, due to high air velocity in the evaporation column, was greater than that of large drops.

5.4.2.7 Boron (B)

Boron in water is present mainly in the form of boric acid, which is a very weak acid. At a pH lower than 7, boric acid is present mainly as undissociated $B(OH)_3$. At a pH higher than 11.5, boron occurs mainly as dissociated borate $B(OH)_4^-$, as shown in equation 5-4 (Maciel et al., 1978).

Boron samples from different waste streams, raw leachate, concentrate and condensate, were collected, as described in section 4.4, and analyzed by Caduceon Laboratory. The results of the mass balances on boron for different leachate process runs are presented in Table G-6, Appendix G. The effects of different process variables on boron concentration during the leachate batch runs are discussed in this section.

a) Effect of temperature

The results of the mass balances on boron for three process runs (LR-1, LR-2 and LR-3) conducted at 42, 61 and 82 °C indicated that about 57 to 61 percent boron remained in the

concentrate stream. About 2 to 3 percent of the total boron appeared in the condensate stream. The remaining 36 to 41 percent could not be detected in these streams. It is believed that either some of the boron left in the exhaust air stream in the form of leachate drops or some may be attached to the walls of the evaporator or heat exchanger as scale, or may be deposited in the bottom of the evaporation column as scum.

The mass fraction of boron in the concentrate and condensate was plotted against the respective evaporator temperatures and are shown in Figure 5.21. The overall trends have shown that with the increase in temperature of the evaporator the boron mass in the concentrate decreased but increased in the condensate. These trends indicated the same tendency as discussed for the chloride in section 5.3.2.6-a.

b) Effect of pH

To determine the effect of pH on boron during three process runs conducted at the initial pH values from 7.5, 5.5 and 3.5, boron mass fractions in the concentrate and condensate streams were calculated and plotted against the respective pH values, as shown in Figure 5.22. The temperature, leachate and airflow were constant during these process runs. The results indicated that only about 52 to 59 percent boron could be recovered in the concentrate stream. It is believed that the remaining boron was lost as leachate droplets, which either left the system with exhaust air or deposited as scale or scum in the evaporation column. On the basis of only three data points for these process runs no clear trend could be identified to determine the effect of pH on boron.

c) Effect of nozzle

The mass fractions of boron in the concentrate and condensate were calculated and plotted against the respective nozzle type, as shown in Figure 5.23. These results showed that the boron mass in the condensate stream increased when fine droplet nozzles were used. However, no clear trend could be identified in the concentrate stream on the basis of three data points only.

5.4.2.8 Barium (Ba)

Barium falls in the group of magnesium, calcium and strontium, in the periodic table. These metals react with anions like CO_3 and SO_4 and have very low solubility in water and form scale on metal surfaces. During the combustion of propane CO_2 was produced. When dissolved this CO_2 would react with the barium to form BaCO_3 as shown in equation 5-6. BaCO_3 has a very low solubility in water and could form scaling inside the evaporation column.

Barium samples from the raw leachate, concentrate and condensate stream were collected as discussed in section 4.4 and analyzed by the Caduceon Laboratory. Mass balances on Ba^{+2} in the different waste streams were prepared and results are given in Table G-7 Appendix G. The effect of the process variables on Ba^{+2} during batch type leachate process runs are discussed in this section.

a) Effect of temperature

The results of the Ba^{+2} mass fractions in the concentrate and condensate at different temperature conditions are shown in Figure 5.24. The mass balances during the process runs, LR-3, LR-1 and LR-2, conducted at the evaporator temperatures of 42, 61 and 82, showed that only 3 to 5 percent of Ba^{+2} was present in the concentrate stream. The amount of Ba^{+2} mass fraction measured in the condensate was less than 1 percent. The remaining 97 to 95 percent of Ba^{+2} disappeared and could not be identified in these waste streams.

To trace this missing Ba^{+2} , samples of the scaling and scum were collected from the evaporator. The results showed that the Ba^{+2} concentration in the scale was 473 mg/lit and in scum it was 238 mg/lit. This scale and scum were deposited in the evaporation column over the entire period of leachate process runs. The total quantity of scale and scum were not calculated. Therefore it was difficult to make a complete mass balance on the Ba^{+2} for each run. However, it was assumed that the missing Ba^{+2} was either deposited in the evaporation column either as scale or scum or may have left in the exhaust air in the form of leachate droplets.

b) Effect of pH

The mass fraction of Ba^{+2} in the concentrate and condensate streams was shown in Figure 5.25 against the respective pH conditions, during different process runs. The mass balance results indicated that the pH of the raw leachate had a very significant effect on the Ba^{+2} concentration in the different waste streams. With a change in pH of the raw

leachate from 7.5, 5.5 to 3.5 the mass of Ba^{+2} observed in the concentrate increased from 3, 101 to 109 percent, respectively. The slight increase in the Ba^{+2} mass more than 100 percent, at low pH, indicated that some of the Ba^{+2} , which had already been deposited on the evaporator walls as scaling during the previous runs, could have been dissolved in the liquid and appeared in the concentrate stream.

It is believed that due to addition of HCl the Cl^- ions concentration in the liquid was increased. Ba^{+2} may have reacted with Cl^- in the acidic environment to form BaCl_2 due to lack of carbonates at low pH. The solubility of BaCl_2 is more than the BaCO_3 . The hot water solubility of BaCl_2 is 0.0192 gm/100ml as compared to 0.006 gm/100ml for BaCO_3 (Weast, 1970). Therefore during the low pH runs Ba^{+2} remained in the liquid phase and was measured in the concentrate stream.

c) Effect of nozzle type

The mass fraction of Ba^{+2} in the concentrate and condensate streams was plotted against the respective nozzle types and shown in Figure 5.26. These results indicated that the Ba^{+2} mass in the concentrate increased when the large drop size nozzle was used. The mass fraction of the Ba^{+2} observed in the concentrate was about 9 percent when GSS10 nozzle was used as compared to about 3 percent when nozzles L54 and L66 were used. The reasons for this could be that the nozzles L54 and L66 had wider spray cones as compared to GSS10 nozzle, which increased the chances of direct contact of leachate with the walls of the evaporation column and hence increased the potential of scaling when L54 and L66 nozzles were used. The results presented in Figure 5.26 also indicated

that the Ba^{+2} mass in the condensate stream increased when the fine droplet nozzles were used.

5.4.2.9 Hydrogen sulfide (H_2S)

The liquid phase H_2S analysis for the raw leachate, concentrate and condensate stream were performed by the Caduceon Laboratory. The method employed by the laboratory reported total S^{-2} in the liquid and further calculations were performed to find H_2S , based on the pH of the liquid. H_2S in the exhaust air stream was analyzed using Gastec detection tubes. H_2S concentrations in the air phase could not be determined with a detection limit of 0.1 mg/lit. The results of the mass balances on H_2S during different process runs are given in Table G-8, Appendix G.

Dissociation of the H_2S is described by the following two chemical equations (Eq 5-6 and 5-7). These equations show that with an increase in H^+ ion concentration in the liquid, transformation of the S^{-2} or HS^- to produce H_2S would be increased. With limited H^+ ions the concentration of the HS^- or S^{-2} in the liquid will be increased.

a) Effect of temperature

The results of mass balances for H_2S indicated that with a change in the evaporator temperature from 42 to 82 °C, the remaining H_2S observed in the concentrate was changed from 4 to 1 percent. The mass of H_2S that appeared in the condensate stream

was less than 1 percent. It is believed that all the remaining mass H_2S might have been converted into the gas phase. The fractions of the mass observed in the concentrate and condensate streams were plotted against the temperatures and are shown in Figure 5.27. These results show that at high temperatures in the evaporator, the H_2S concentration in the concentrate decreased as compared to the runs conducted at low temperatures. It is believed that the reason for this was that with the increase in the temperature the value of the Henry's law coefficient was increased, which resulted in an increased concentration of H_2S in the gas phase. With an increase in temperature from 40 to 60 °C the value of Henry's law coefficient for H_2S increases from 0.074 to 0.103 (Tchobanoglous et al., 1993), respectively.

b) Effect of pH

During process runs LR-4 and LR-5 the pH of the raw leachate was adjusted to 5.5 and 3.5, respectively. H_2S in the raw leachate was monitored after the initial adjustment of pH. The resulting H_2S concentrations in the raw leachate, concentrate and condensate are given in Tables E-4 and E-5, Appendix E. The results show very low concentrations of H_2S in the raw leachate, during the processes when initial pH was adjusted. This indicated that at low pH conditions, more free H^+ ions were available, that converted S^{2-} or HS^- ions present in the liquid to produce H_2S , which volatilized and transformed into the gas phase.

H_2S was a very unstable contaminant in the leachate and its concentration in the liquid phase changed very rapidly. During the evaporation process the pH of the concentrate

increased, as clear from the final pH values given in Table 5.10. With the increase in pH, at low H^+ ion concentration, H_2S would start producing more stable forms of S^{2-} ions, which resulted in an increased value of H_2S in the concentrate stream.

The results of mass fraction of H_2S in the concentrate and condensate streams, at different raw leachate pH values are plotted as shown in Figure 5.28. These results did not indicate any clear trend to determine the behavior of H_2S in the concentrate and condensate streams on the basis of these three data points only.

c) Effect of nozzle

The mass fraction of H_2S observed in the concentrate and condensate streams, when different types of spray nozzles were used are shown in Figure 5.29. These results indicate that the H_2S mass in the condensate was increased when fine droplet nozzles were used. No clear trend could be discerned from the values of the concentrate stream on the basis of these three data points.

5.4.2.10 Toluene and m/p-xylenes

The concentrations of toluene and m/p-xylene were analyzed in both liquid and gas phases, as discussed in section 4.5.7. Mass balances were prepared to compare the mass of toluene and m/p-xylene entering the system as raw leachate and leaving the system as concentrate, condensate and exhaust air streams. The results of these mass balances are given in Table G-9 and Table G-10 Appendix G. The results showed that no toluene and m/p-xylene was observed in the concentrate stream, which indicated that the entire

toluene and m/p-xylene present in the raw leachate were evaporated and transformed into the gas phase. The condensate stream had small amounts of toluene and m/p-xylene. The total mass fraction of the toluene and m/p-xylene in the condensate was less than 1 percent in all the process runs.

Mass balances for toluene and m/p-xylene were performed for the entire leachate process runs to observe the impact of different process variables. Each process variable (temperature, pH and nozzle type) was examined for three conditions to observe the changes in toluene and m/p-xylene concentration in different waste streams, entering and leaving the system. The results of the exhaust gases indicated that most of the toluene and m/p-xylene mass was stripped off in the air phase. But on the basis of current data no specific trends in the behavior of toluene and m/p-xylene could be determined in the evaporation process.

5.4.3 Calculations for leachate droplets carried over to the gas phase

As discussed in section 5.2.2, some of the leachate droplets appeared to have been carried over to the exhaust gas stream, without evaporating, due to the high air velocity in the evaporation column. Calculations were made to estimate the amount of these leachate droplets going in the exhaust gases. For this, the mass balance data for chlorides during the leachate runs was used.

During the process run 30 litres of raw leachate was evaporated to produce 3 litres of concentrate. The amount of condensate generated in this process run was assumed to be comprised of two components: a) produced due to condensation of the water vapors in

the heat exchanger; and b) produced due to accumulation of the leachate droplets present in the exhaust gases which were trapped in the heat exchanger. Similarly it was assumed that the exhaust gases leaving the heat exchanger were also comprised of the same two components: a) water vapors; and b) leachate droplets. The 'a' component of both the waste streams was assumed to be clean water without any contamination. The 'b' component of the waste streams carried the contaminants from the feed tank and polluted the condensate stream and air stream. The remaining part of the chlorides that could not be detected in the concentrate or condensate streams was considered to be carried over as leachate droplets. It is believed that these leachate droplets either left the system along with the exhaust gases or attached to the walls of the evaporator or deposited as scum in the bottom of the evaporation column, as discussed in section 5.3.2.5.

To calculate the volume of the leachate droplets carried over to the exhaust gas stream four equations were developed. Two water balance equations were: 1) water balance across the complete system (equation 5-8); and 2) water balance for the condensate stream only (equation 5-9). Similarly, two mass balance equations developed were: 3) mass balance across the complete system (equation 5-10); and 4) mass balance for the condensate stream (equation 5-11).

$$V_0 = V_1 + V_2 + V_{3a} + V_{3b} \quad (\text{Eq 5-8})$$

$$V_2 = V_{2a} + V_{2b} \quad (\text{Eq 5-9})$$

$$V_0 S_0 = V_1 S_1 + V_2 S_2 + V_{3b} S_{3b} \quad (\text{Eq 5-10})$$

$$V_2 S_2 = V_{2b} S_{2b} \quad (\text{Eq 5-11})$$

In these equations: V_0 is the volume of raw leachate; V_1 is volume of concentrate; V_2 is volume of condensate; V_{2a} is volume of water vapors in condensate; V_{2b} is the volume of leachate droplets carried over to condensate; V_{3a} is volume of water vapors and V_{3b} is volume of leachate droplets carried over to the exhaust air stream. S_0 is the chloride concentration in the raw leachate; S_1 is chloride concentration in the concentrate; S_2 is chloride concentration in the condensate; S_{2b} is the chloride concentration in the leachate droplets carried over to condensate; S_{3b} is chloride concentration in the leachate droplets carried over to the exhaust air stream.

The above-mentioned equations (Eq 5-8 to Eq 5-11) were solved simultaneously to determine the volume of the leachate droplets carried over to the exhaust gases. It was assumed that the chloride concentration in leachate droplets in both the streams (condensate and exhaust gas) were the same. The chloride concentration in the leachate droplet was assumed to be equal to the chloride concentration of the concentrate, to simplify the calculations. It was difficult to predict the actual amount of the water vapors and leachate droplets leaving the system, however, these calculations were conducted to have rough estimation of the leachate droplets that carried over to the gas phase with out evaporation.

On the basis of the chloride data the volume of the leachate droplets carried over to the exhaust air stream was calculated for three leachate runs, LR-7, LR-1 and LR-6, conducted with three different nozzles, GSS10, L66 and L54, respectively. The results indicated that the fraction of leachate droplets carried over to the exhaust air during these

runs increased from 2.9 to 6.5 percent, respectively. This clearly shows that a higher percentage of the leachate droplets were going in the exhaust air when the fine droplet nozzle (L54) was used as compared to the large droplet nozzle (GSS10). Similarly, the fraction of the leachate droplets carried over were calculated for three process runs conducted at evaporator temperatures of 42, 60 and 82 °C, during process runs, LR-3, LR-1 and LR-2, respectively. The results for these process runs indicated that the fraction of leachate carry over increased from 3.4 to 6.3 percent, respectively. This shows that with the increase in the evaporator temperature more leachate droplets were carried over to the exhaust air.

5.4.4 Extended run analysis results

During the extended leachate run, LR-8, 300 litres of the raw leachate were evaporated for about 23 hours to reduce its volume to 30 litres and to concentrate the contaminants in the raw leachate to a higher degree. Nozzle GSS6.5 was used during the extended run and the combustor temperature was set around 800 °C. Samples of the raw leachate, concentrate, condensate and exhaust air streams were collected as explained in section 4.4. Three concentrate and two condensate samples were analyzed, to monitor the changes in the concentration of the target contaminant in the liquid phase. The results of the liquid and gas phase samples, collected at different intervals, to observe the behavior of the target contaminants in different waste streams and discussed in this section.

The pH of the concentrate and condensate increased during the extended run. The pH increased from a raw leachate pH of 7.6 to 8.66, 8.72 and 8.98 in the concentrate stream

for the 8, 16 and 23 hour intervals, respectively. The pH of the condensate was observed to be 8.86 and 8.99 for the 8 and 16 hours intervals. This elevated pH of the concentrate is believed to substantially have impacted on the fate of leachate contaminants.

The TSS concentrations measured in the concentrate stream, collected at different intervals, during the extended run are shown in Figure 5.30, along with TSS concentrations of the raw leachate. These results clearly demonstrated that the TSS concentration in the concentrate increased with the time. The slope of the plot increased with the run duration. Similar trend was observed in the TSS concentration of the condensate stream, where TSS values changed from 18 to 32 mg/lit in condensate samples 1 and 2, respectively.

The COD concentration in the concentrate stream was plotted against the run duration, as depicted in Figure 5.31. The results showed that the COD concentration in the concentrate stream increased with the run duration, as explained in section 5.3.2.2. The COD of the concentrate at the end of the run was increased to about 4 times higher as compared to COD of the raw leachate. An increase in COD of the condensate from 383 to 706 mg/lit was observed at 8 and 16 hours intervals.

The TKN concentrations in the concentrate stream were plotted versus time and are shown in Figure 5.32. The plot of TKN indicated that the TKN concentration of the raw leachate (1380 mg/lit) decreased rapidly to 533 mg/lit during the initial 8 hours, which is most likely due to the volatilization of NH_3 component of the total nitrogen. On further

evaporation the TKN concentration started to increase to 571 and 818 mg/lit, at the end of 16 and 23 hours duration, respectively. It was anticipated that during this time very little amount of ammonia was left in the concentrate stream, because most of the ammonia component of TKN was volatilized during the initial period of the run, therefore, the other forms of TKN were concentrated to increase the TKN concentration in the concentrate stream.

The ammonia concentrations in the concentrate stream were also plotted against the process duration, as shown in Figure 5.33. The figure shows that the ammonia concentration in the concentrate decreased with the process time. It is believed that with the increase in the pH that increased with the process duration, more ammonium ions (NH_4^+) were changed to produce ammonia (NH_3), which disappeared in the exhaust air quickly due to its high volatility. The results of ammonia during the initial 8 hours duration showed a rapid decrease in the ammonia concentration in the concentrate, which was comparable to the results of TKN, as clear from the TKN plot, Figure 5.32. During the extended evaporation duration, the ammonia concentration in the concentrate continued decreasing with time, because, as the process continued very little amount of ammonia remained in the liquid phase.

The results of the gas phase ammonia are given in Table E-10, Appendix E. The results showed that the exhaust air had a very high concentration of ammonia at time zero (which was monitored during the first 30 minutes of the run). As the run continued for an extended period the concentration of ammonia in the exhaust air was almost at a steady

rate. It is believed that at time zero the feed bucket was filled with the fresh raw leachate which had a high concentration of ammonia. Therefore a high concentration of ammonia was observed in the gas phase during the initial phase of the evaporation process. However as time progressed the mass flow of ammonia in the gas phase would be limited by the leachate feed rate.

The chloride concentrations in the concentrate stream are plotted in Figure 5.34, for the extended run (LR-8). The results showed that the chloride concentrations in the concentrate stream continuously increased with the run duration. The same trend was observed in the condensate stream, where the chloride concentration increased from 506 mg/lit to 1206 mg/lit in condensate samples 1 and 2, after 8 and 16 hours, respectively.

Figure 5.35, showed the boron results versus time, during the extended run. These results showed that the boron concentration in the concentrate stream increased with the process duration. The raw leachate boron concentration of 18.7 mg/lit increased to 120 mg/lit, in the final concentrate. The condensate stream also showed an increase in boron concentration from 11.3 to 19.2 mg/lit, after 8 to 16 hours duration, respectively.

The barium concentrations in the concentrate stream during the extended run were plotted against the process duration, as shown in Figure 5.36. These results indicated that the barium in the concentrate decreased with the duration. It is believed that the barium could have reacted with the CO_2 to form BaCO_3 and produced scaling inside the evaporation column, as discussed in section 5.3.2.8. Due to continuous recycling of the leachate

during the extended run chances of scale forming were increased and hence decreased the amount of barium in the concentrate stream.

The H_2S concentrations in the concentrate stream were plotted against the evaporation duration, during the extended run, as shown in Figure 5.37. The results of the plot showed that the H_2S concentration in the concentrate increased with the time. Over the period of the run the pH of the concentrate stream increased from 7.6 to 8.9, respectively. With the increase in pH of leachate the concentration of H^+ ions in the concentrate were reduced and the non-volatile S^{2-} in the leachate started accumulating in the other forms, which ultimately resulted in an increased concentration of H_2S , when these samples were analyzed.

During the extended run, no significant concentrations of toluene and m/p-xylene could be observed in the concentrate. The results of the condensate and air streams during the extended run are given in Table E-8 and E-10 of Appendix E. A very small amount of toluene and m/p-xylene was observed in the condensate, which was further reduced with time. Similarly, very little benzene and m/p-xylene were observed in the exhaust air stream. No toluene and m/p-xylene could be observed in the exhaust air stream after the initial air samples taken at the start of the run.

5.5 Calculations of energy in landfill gases to evaporate the leachate

It was projected that landfill gases could be used as an energy source to evaporate the landfill leachate. To make these calculations actual landfill leachate and gas generation

data, for year 2001 and 2002, from the Trail Road landfill site was analyzed. On an average about 7577 litre per hour (City of Ottawa, 2003) of leachate is produced from the landfill site. As discussed in the section 5.1.4 on an average about 2600 kJ energy is required to evaporate one litre of the leachate. Therefore the energy required to evaporate total landfill leachate could be estimated to be about 19.7×10^6 kJ per hour.

The Trail Road landfill site has an estimated gas flow range of $1458 \text{ m}^3/\text{hr}$ to $2429 \text{ m}^3/\text{hr}$, with an average of $1721 \text{ m}^3/\text{hr}$ (City of Ottawa, 2003). Methane is one of the major components in the landfill gases, which is associated with substantial amount of energy. Trail road landfill gases has a methane concentration of 28% to 40%, with an average being 33 to 34 percent. Using a lower heating value of methane as 802 kJ per mole, it is estimated that landfill gas can produce total energy of about 36.13×10^6 kJ per hour. On the basis of these calculations it is evident that the landfill gases have enough energy available to treat the landfill leachate using evaporation technology.

Figure 5.1: Effect of nozzle type on evaporate and condensate generation

Figure 5.2: Effect of water flow rate on evaporate and condensate generation

Figure 5.3: Effect of airflow rate on evaporate and condensate generation

Figure 5.4: Combustor temperatures at different fuel flow rates

Figure 5.5: Effect of temperature on evaporate & condensate generation

Figure 5.6: Comparison of leachate & water evaporation with different nozzles

Figure 5.7: Comparison of leachate & water evaporation rates at different temperatures

Figure 5.8: Effect of temperature on TSS behavior

Figure 5.9: Effect of nozzle type on TSS

Figure 5.10: Effect of temperature on COD

Figure 5.11: Effect of pH on COD

Figure 5.12: Effect of nozzle type on COD

Figure 5.13: Effect of temperature on TKN

Figure 5.14: Effect of pH on TKN

Figure 5.15: Effect of temperature on ammonia

Figure 5.16: Effect of pH on ammonia

Figure 5.17: Effect of nozzle type on ammonia

Figure 5.18: Effect of temperature on chloride

Figure 5.19: Effect of pH on chloride

Figure 5.20: Effect of nozzle on chloride

Figure 5.21: Effect of temperature on boron

Figure 5.22: Effect of pH on boron

Figure 5.23: Effect of nozzle type on boron

Figure 5.24: Effect of temperature on barium

Figure 5.25: Effect of pH on barium

Figure 5.26: Effect of nozzle on barium

Figure 5.27: Effect of temperature on H_2S Figure 5.28: Effect of pH on H_2S

Figure 5.29: Effect of nozzle on H₂S

Figure 5.30: TSS concentrations during the extended run

Figure 5.31: COD concentrations during the extended run

Figure 5.32: TKN concentrations during the extended run

Figure 5.33: Ammonia concentrations during the extended run

Figure 5.34: Chloride concentrations during the extended run

Figure 5.35: Boron concentrations during the extended run

Figure 5.36: Barium concentrations during the extended run

Figure 5.37: H₂S concentrations during the extended run

Figure 5.38: Comparison of experimental and model results at different temperatures

Chapter 6: Conclusions and recommendations

6.1 Conclusions

The pilot scale evaporator was designed to evaluate the use of evaporation technology for treatment of landfill leachate. This is an on-site treatment option, where energy from combustion of the landfill gases could be used to evaporate the landfill leachate. The performance of the evaporator was evaluated by evaporating tap water and actual landfill leachate from the Trail Road site.

The results indicated that the evaporator temperature was the most important variable, which had a significant effect on the evaporation rate. Airflow rate and saturation of the air stream was also a limiting factor controlling the evaporation rate, during these runs. The evaporator provided best efficiency with a fine droplet nozzle type at a high evaporator temperature and airflow rate.

The results of evaporation showed that about 2600 kJ energy was required to evaporate 1-litre of water, which was comparable to the theoretical amount of energy required to evaporate 1-litre water at 20 °C. Calculations based on the actual landfill data showed that the landfill gases generated at the Trail Road have adequate energy to evaporate the leachate generated at the site.

During the leachate evaporation runs leachate was continuously recycled and 90% of the raw leachate volume was evaporated. Contaminants in the leachate were concentrated in

the remaining 10% volume. Therefore, with the use of the evaporation technology contaminants in the leachate could be concentrated into a small fraction of the original leachate volume.

About 60 to 70 percent of the total mass of nonvolatile compounds such as boron and chloride were observed in the concentrate stream. Their percentage in the concentrate was decreased with increase in the temperature of the evaporation column and when fine droplet nozzles were used.

The COD mass fraction in the concentrate decreased with increase in the temperature of the evaporation column and when fine droplet nozzles were used. With an increase in the initial pH the COD mass fraction in the concentrate increased. The results of the concentrate streams showed complete volatilization of toluene and m/p-xylene from the liquid phase.

Results of the raw leachate for TKN and ammonia indicated that about 80 percent of TKN was composed of ammonia. Mass balances for ammonia indicated that only less than 1 percent of ammonia was remained in the concentrate, during the evaporation runs conducted without pH adjustment. Rate of ammonia volatilization from the liquid phase increased with the temperature of the evaporation column and with the use of fine droplet nozzles.

The pH of the raw leachate had a very significant affect on the mass of the TKN and ammonia in different waste streams. Ammonia mass retained in the concentrate stream was about 80 percent during the run conducted at a pH of 3.5.

The mass of barium in the concentrate decreased during the process when the pH of raw leachate was not adjusted, because most of the barium was deposited as scale and scum inside the evaporation column. During the process runs when the pH was reduced the total barium in the raw leachate remained in the concentrate stream.

The condensate stream contained a noticeable amount of contaminants because about 3 to 6 percent of the leachate droplets were carried over to the exhaust air stream and some of which appeared in the condensate.

6.2 Recommendations

During the evaporation process all the contaminants in the raw leachate were either concentrated in the concentrate stream or lost in the exhaust air. The air stream was analyzed for ammonia, H_2S , toluene and m/p-xylene on the basis of grab air samples from the evaporator. Detailed investigations are recommended to evaluate the exhaust air stream, whether it comply with the air emission regulations or if any level of treatment is required for this stream.

Due to the high air velocity in the evaporation column some leachate droplets left in the air stream, which increased the concentration of the contaminants in the condensate

stream. Some modification should be made to avoid the transfer of the leachate drops to the air stream. An increase in the evaporator diameter could reduce the air velocity.

Over the period of the experimental runs a layer of scale and scum was deposited on the walls and bottom of the evaporation column, which ultimately clogged the drainpipe in the evaporation column. Some measures should be taken to avoid this fouling or periodic cleaning of the column is advised. Some spray nozzles had wider spray angles and some of the water droplets directly impinge the walls of the evaporator. An increase in the diameter of the evaporation column would minimize this effect.

Saturation of air is a limiting factor affecting the evaporation rate. An increase in airflow rate would increase the evaporation rate. Saturation of the air stream was estimated from comparison of theoretical amount of water in exhaust air calculated on the basis of the saturate vapor pressure and water evaporated in the system. It is recommended that humidity of the exhaust air should also be measured.

Due to direct contact of liquid and hot air most of the heat transfer was carried out in the evaporation column. The utility of heat exchanger to preheat the leachate is not significant.

The results of the evaporation model showed that velocity of the falling droplet had a significant effect on the evaporation rates but this factor was not evaluated in these experimental runs. Some method to evaluate the effect of initial droplet falling velocity

should be adopted or leachate may be sprayed upward, so that the droplets could fall under gravity, which will reduce the droplet falling velocity.

Water pressure in the system was estimated based on the flow rate to determine the droplet size from the manufacturers nomographs. Flow in the system was measured manually. It is recommended that flow meters may be used on the water supply line to measure the flow through the system. It is also recommended that pressure gauges may be used on the water supply line to determine actual pressure in the system to have a better estimation of the droplet size.

Some water droplets were directly impinging the walls of the evaporation column. Visual observations were made to estimate the percentage of these droplets. This can be better judged if the spray is made directly above the evaporation column, measure the amount of water falling inside the evaporation column and compare with the flow rate.

The temperatures of the evaporation column and heat exchanger were monitored by the thermocouples that were placed in the top portions of these units. It is recommended that more thermocouples at different heights may be used for better estimation of the inside operating temperatures.

Chapter 7: References

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Appendix A: Drawings of pilot scale evaporator components

Figure A-1: Leachate supply in the evaporator

Figure A-2: Evaporation column

Figure A-3: Heat exchanger

Figure A-4: Combustor layout

Appendix B: Water test runs data

Table B-1: Water run 1 (WR-1); at 800° C, Nozzle GSS10, flow 3.49 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0			800	69.5	68.6	62.0 63.2
0.15	3.0	12					
0.30	3.4	13.6		800	69.6	68.8	62.2 63.3
0.45	3.1	12.4					
1.00	3.2	12.8		800	69.6	68.9	62.3 63.3
1.15	3.1	12.4					
1.30	3.5	14		801	69.8	69.2	62.3 63.4
1.45	3.2	12.8					
2.00	3.2	12.8		800	69.7	69.1	62.3 63.4
120	25.7		2.22				
Hourly ave.	12.85	12.85	1.11	800.20	69.64	68.92	62.22 63.32
STDEV		0.66		0.45	0.11	0.24	0.13 0.08

Table B-2: Water run 2 (WR-2); at 800° C, Nozzle GSS10, flow 3.86 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0			800	69.3	68.1	61.8 63.1
0.15	3.1	12.4					
0.30	3.6	14.4		802	69.4	68.3	61.9 63.2
0.45	3.3	13.2					
1.00	3.4	13.6		799	69.3	68.3	62.0 63.2
1.15	3.1	12.4					
1.30	3.7	14.8		800	69.4	68.4	62.0 63.2
1.45	3.4	13.6					
2.00	3.3	13.2		798	69.3	68.4	62.0 63.2
120	26.9		2.32				
Hourly ave.	13.45	13.45	1.16	799.80	69.34	68.30	61.94 63.18
STDEV		0.85		1.48	0.05	0.12	0.09 0.04

Table B-3: Water run 3 (WR-3); at 800° C, Nozzle GSS6.5, flow 3.17 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0			800	60.5	59.4	56.0 56.8
0.15	3.2	12.8					
0.30	3.4	13.6		802	60.6	59.6	56.1 56.9
0.45	3.7	14.8					
1.00	3.5	14.0		800	60.6	59.6	56.1 56.9
1.15	3.5	14.0					
1.30	3.6	14.4		801	60.7	59.7	56.2 56.9
1.45	3.8	15.2					
2.00	3.6	14.4		800	60.7	59.8	56.2 56.9
120	28.30		3.36				
Hourly ave.	14.15	14.15	1.68	800.60	60.62	59.62	56.12 56.88
STDEV		0.74		0.89	0.08	0.15	0.08 0.04

Table B-4: Water run 4 (WR-4); at 800° C, Nozzle GSS6.5, flow 3.45 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0			800	60.6	59.3	56.4 57.7
0.15	3.5	14					
0.30	3.7	14.8		801	60.7	59.4	56.6 57.7
0.45	3.8	15.2					
1.00	3.7	14.8		800	60.7	59.5	56.6 57.8
1.15	3.8	15.2					
1.30	3.7	14.8		802	60.9	59.7	56.7 57.8
1.45	3.9	15.6					
2.00	3.8	15.2		803	61.0	59.9	56.7 57.8
120	29.90		3.54				
Hourly ave.	14.95	14.95	1.77	801.20	60.78	59.56	56.60 57.76
STDEV		0.48		1.30	0.16	0.24	0.12 0.05

Table B-5: Water run 5 (WR-5); at 800° C, Nozzle L54, flow 2.33 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0			801	52.6	50.5	47.8 49.0
0.15	3.4	13.6					
0.30	3.8	15.2		804	52.9	50.9	47.9 49.0
0.45	3.6	14.4					
1.00	3.5	14.0		801	52.9	50.9	48.0 49.1
1.15	3.6	14.4					
1.30	3.6	14.4		803	53.0	51.1	48.1 49.1
1.45	3.7	14.8					
2.00	3.7	14.8		803	53.0	51.0	48.1 49.2
120	28.90		4.22				
Hourly ave.	14.45	14.45	2.11	802.40	52.88	50.88	47.98 49.08
STDEV		0.50		1.34	0.16	0.23	0.13 0.08

Table B-6: Water run 6 (WR-6); at 800° C, Nozzle L54, flow 2.45 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0			800	52.5	50.1	47.4 48.5
0.15	3.6	14.4					
0.30	4.1	16.4		801	52.7	50.4	47.5 48.6
0.45	3.9	15.6					
1.00	3.7	14.8		799	52.8	50.6	47.5 48.6
1.15	4.0	16.0					
1.30	3.9	15.6		800	52.8	50.7	47.6 48.6
1.45	4.1	16.4					
2.00	3.8	15.2		798	52.8	50.7	47.5 48.6
120	31.1		4.52				
Hourly ave.	15.55	15.55	2.26	799.60	52.72	50.50	47.50 48.58
STDEV		0.72		1.14	0.13	0.25	0.07 0.04

Table B-7: Water run 7 (WR-7); at 800°C, Nozzle L66, flow 2.42 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0			800	61.0	59.8	56.8 57.9
0.15	3.3	13.2					
0.30	3.5	14.0		802	61.1	60.0	56.9 58.0
0.45	3.4	13.6					
1.00	3.8	15.2		809	61.4	60.3	57.0 58.1
1.15	3.4	13.6					
1.30	3.5	14.0		801	61.3	60.2	57.0 58.0
1.45	3.6	14.4					
2.00	3.6	14.4		806	61.4	60.3	57.1 58.1
120	28.10		3.14				
Hourly ave.	14.05	14.05	1.57	803.60	61.24	60.12	56.96 58.02
STDEV		0.62		3.78	0.18	0.22	0.11 0.08

Table B-8: Water run 8 (WR-8); at 800°C, Nozzle L66, flow 2.84 lit/min, air flow valve completely open

Time (min)	Evaporate		Condensate (lit)	Air velocity (m/min)	Temperature (°C)			
	(lit/15min)	(lit/hr)			Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0			149.3	801	60.5	59.2	56.6 57.8
0.15	3.4	13.60						
0.30	3.9	15.60		147.5	805	60.7	59.4	56.7 57.9
0.45	3.5	14.00						
1.00	3.7	14.80		148.4	802	60.6	59.5	56.8 57.9
1.15	3.8	15.20						
1.30	4.0	16.00		146.3	804	60.9	59.8	56.9 58.0
1.45	3.8	15.20						
2.00	3.7	14.80		147.5	803	61.0	60.0	56.9 58.0
120	29.80		3.30					
Hourly ave.	14.90	14.90	1.65	147.82	803.00	60.74	59.58	56.78 57.92
STDEV		0.79		1.14	1.58	0.21	0.32	0.13 0.08

Table B-9: Water run 9 (WR-9); at 800°C, Nozzle L66, flow 2.84 lit/min, airflow valve 7/10 open)

Time (min)	Evaporate		Condensate (lit)	Air velocity (m/min)	Temperature (°C)		
	(lit/15min)	(lit/hr)			Combustor	Evaporator	Water before Heat Exch. Water after Heat Exch.
0.00	0			132.0	814	62.4	57.0 58.4
0.15	3.4	13.6					
0.30	3.6	14.4		132.3	815	62.5	57.1 58.4
0.45	3.6	14.4					
1.00	3.5	14.0		132.6	814	62.5	57.2 58.5
1.15	3.6	14.4					
1.30	3.6	14.4		134.1	815	62.6	57.3 58.5
1.45	3.7	14.8					
2.00	3.8	15.2		132.9	815	62.6	57.3 58.5
120	28.80		2.96				
Hourly ave.	14.40	14.40	1.48	132.77	814.60	62.52	57.18 58.46
STDEV		0.48		0.83	0.55	0.08	0.13 0.05

Table B-10: Water run 10 (WR-10); at 800°C, Nozzle L66, flow 2.84 lit/min, airflow valve 3/10 open)

Time (min)	Evaporate		Condensate (lit)	Air velocity (m/min)	Temperature (°C)		
	(lit/15min)	(lit/hr)			Combustor	Evaporator	Water before Heat Exch. Water after Heat Exch.
0.00	0			131.4	820	63.0	57.3 58.6
0.15	3.4	13.6					
0.30	3.6	14.4		130.5	819	63.1	57.4 58.7
0.45	3.5	14.0					
1.00	3.5	14.0		129.5	820	63.1	57.4 58.7
1.15	3.6	14.4					
1.30	3.5	14.0		129.8	820	63.2	57.5 58.8
1.45	3.5	14.0					
2.00	3.6	14.4		128.6	821	63.3	57.5 58.8
120	28.20		2.84				
Hourly ave.	14.10	14.10	1.42	129.96	820.00	63.14	57.42 58.72
STDEV		0.28		1.02	0.71	0.11	0.08 0.08

Table B-11: Water run 11 (WR-11); at 400° C, Nozzle L66, flow 2.84 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0.00			400	31.5	30.9	29.7 30.4
0.15							
0.30	2.10	4.2		401	31.6	31.0	29.8 30.5
0.45							
1.00	2.30	4.6		402	31.6	31.0	29.8 30.5
1.15							
1.30	2.20	4.4		399	31.6	31.1	29.9 30.5
1.45							
2.00	2.20	4.4		401	31.6	31.1	29.9 30.5
120	8.80		1.28				
Hourly ave.	4.40	4.40	0.64	400.60	31.58	31.02	29.82 30.48
STDEV		0.16		1.14	0.04	0.08	0.08 0.04

Table B-12: Water run 12 (WR-12); at 600° C, Nozzle L66, flow 2.84 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0.00			600	42.9	41.8	39.1 40.0
0.15							
0.30	4.50	9		603	43.0	41.9	39.2 40.0
0.45							
1.00	4.70	9.4		600	43.0	42.0	39.3 40.1
1.15							
1.30	4.90	9.8		604	43.1	42.1	39.3 40.1
1.45							
2.00	4.70	9.4		602	43.1	42.1	39.3 40.1
120	18.80		2.62				
Hourly ave.	9.40	9.40	1.31	601.80	43.02	41.98	39.24 40.06
STDEV		0.33		1.79	0.08	0.13	0.09 0.05

Table B-13: Water run 13 (WR-13); at 1000° C, Nozzle L66, flow 2.84 lit/min

Time (min)	Evaporate		Condensate (lit)	Temperature (°C)			
	(lit/15min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Water before Heat Exch. Water after Heat Exch.
0.00	0.00			1001	81.8	79.7	69.4 71.0
0.15	4.60	18.40					
0.30	5.60	22.40		1008	82.1	80.0	69.4 71.1
0.45	5.10	20.40					
1.00	5.00	20.00		1002	82.0	80.0	69.5 71.1
1.15	5.10	20.40					
1.30	5.30	21.20		1007	82.1	80.1	69.6 71.2
1.45	5.20	20.80					
2.00	5.20	20.80		1005	82.1	80.1	69.6 71.2
120	41.10		6.01				
Hourly ave.	20.55	20.55	3.01	1004.60	82.02	79.98	69.50 71.12
STDEV		1.13		3.05	0.13	0.16	0.10 0.08

Appendix C: Leachate test runs data

Table C-1: Leachate run 1 (LR-1); Batch run at 800°C, Nozzle 66, pH = 7.55

Time (min)	Bucket wt. (Kg)	Evaporate		Condensate (lit)	Temperature (°C)			
		(lit/30min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Leach. after Heat Exch.
0.00	30.00	0.00			800	59.9	58.6	55.9
0.30	22.85	7.15	14.30		801	60.5	59.3	56.1
1.00	15.54	7.31	14.62		800	60.4	59.3	56.2
1.36	7.20	8.34	13.90		802	60.6	59.5	56.4
1.54	3.02	4.18	13.93					57.5
Total = 114		26.98		3.02				
Hourly ave.		14.20	14.20	1.59	801.00	60.50	59.40	56.30
STDEV			0.34		0.96	0.31	0.39	0.21
								0.10

Table C-2: Leachate run 2 (LR-2); Batch run at 1000°C, Nozzle 66, pH = 7.24

Time (min)	Bucket wt. (Kg)	Evaporate		Condensate (lit)	Temperature (°C)			
		(lit/30min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Leach. after Heat Exch.
0.00	30.00	0.00			1000	81.1	79.4	68.2
0.33	19.02	10.98	19.96		1003	81.9	80.3	69.0
1.00	9.83	9.19	20.42		998	81.7	80.1	68.6
1.21	3.01	6.82	19.49		999	81.6	80.1	68.7
Total = 81		26.99		3.66				
Hourly ave.		19.99	19.99	2.71	998.50	81.58	79.98	68.63
STDEV			0.47		2.52	0.42	0.47	0.40
								0.36

Table C-5: Leachate run 5 (LR-5); Batch run at 800°C, Nozzle 66, pH = 3.5

Time (min)	Bucket wt. (Kg)	Evaporate		Condensate (lit)	Temperature (°C)				
		(lit/30min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Leach. before Heat Exch.	Leach. after Heat Exch.
0.00	30.00	0.00			800	60.2	58.5	56.0	57.4
0.30	23.00	7.00	14.00		801	60.8	59.5	56.3	57.6
1.00	15.87	7.13	14.26		803	61.0	59.9	56.5	57.7
1.30	8.62	7.25	14.50		802	61.2	60.3	56.5	57.7
1.54	2.99	5.63	14.08						
Total = 114		27.01		3.06					
Hourly ave.		14.22	14.22	1.60	801.50	60.80	59.55	56.33	57.60
STDEV			0.22		1.29	0.43	0.77	0.24	0.14

Table C-6: Leachate run 6 (LR-6); Batch run at 800°C, Nozzle 54, pH = 7.54

Time (min)	Bucket wt. (Kg)	Evaporate		Condensate (lit)	Temperature (°C)				
		(lit/30min)	(lit/hr)		Combustor	Evaporator	Heat Exch.	Leach. before Leach. after Heat Exch. Heat Exch.	
0.00	30.00	0.00			800	52.6	50.0	47.5	48.9
0.30	22.70	7.30	14.60		799	52.9	50.8	48.0	49.2
1.00	15.37	7.33	14.66		801	53.0	51.3	48.2	49.3
1.30	7.93	7.44	14.88		800	53.0	51.4	48.2	49.3
1.50	3.00	4.93	14.79						
Total = 110		27.00		4.00					
Hourly ave.		14.73	14.73	2.16	800.00	52.88	50.88	47.98	49.16
STDEV			0.13	0.82	0.82	0.19	0.64	0.33	0.21

Table C-7: Leachate run 7 (LR-7); Batch run at 800°C, Nozzle GSS10, pH = 7.28

Time (min)	Bucket wt. (Kg)	Evaporate		Condensate (lit)	Combustor		Temperature (°C)			
		(lit/30min)	(lit/hr)		Evaporator	Heat Exch.	Leach. before	Leach. after	Heat Exch.	Heat Exch.
0.00	30.00	0.00			801	74.9	73.6	68.1	69.7	
0.34	22.86	7.14	12.60		804	75.4	74.3	68.4	69.8	
1.00	16.95	5.91	13.64		801	75.3	74.2	68.7	69.9	
1.40	8.21	8.74	13.11		803	75.7	74.8	68.8	69.9	
2.04	3.00	5.21	13.03							
Total = 124		27.00		1.9						
Hourly ave.		13.06	13.06	0.92	802.25	75.33	74.23	68.50	69.83	
STDEV			0.43		1.50	0.33	0.49	0.32	0.10	

Table C-8: Leachate run 8 (LR-8); Continuous run at 800 °C, Nozzle GSS6.5, pH = 7.59

Time (hrs)	Leachate added (Kg)	Evaporate (lit)	Condensate (lit)	Combustor	Temperature (°C)			
					Evaporator	Heat Exch.	Leach. before Heat Exch.	Leach. after Heat Exch.
Day-1								
0.00	30			800	60.0	58.7	55.9	57.1
0.00	10	10						
0.45	10	10						
1.30	10	10						
2.00	10	10		801	60.1	59.0	56.0	57.2
2.45	10	10						
3.30	10	10						
4.15	10	10		800	60.3	59.1	56.1	57.3
4.50	10	10						
5.35	10	10						
6.17	0	0		798	60.2	59.1	56.1	57.3
Total = 6.28		90.00	10.60					
Hourly ave.		14.33	1.69	799.75	60.15	58.98	56.03	57.23
Day-2								
0.00	10	10		801	60.1	58.9	56.1	57.4
0.45	10	10						
1.30	10	10						
2.00	10	10		802	60.3	59.2	56.3	57.5
2.45	10	10						
3.30	10	10						
4.15	10	10		801	60.3	59.3	56.4	57.5
5.00	10	10						
5.45	10	10						
6.30	10	10		803	60.4	59.4	56.4	57.5
7.00	5	5						
7.18	0	0	22.81					
Total = 7.30		105.00	12.21					
Hourly ave.		14.38	1.70	801.75	60.28	59.20	56.30	57.48

Table C-8: continued.

Time (hrs)	Leachate added (Kg)	Evaporate (lit)	Condensate (lit)	Combustor	Temperature (°C)		
					Evaporator	Heat Exch. Leach. before Heat Exch.	Leach. after Heat Exch.
Day-3							
0.00	10	10		800	60.0	58.8	57.3
0.45	10	10					
1.30	10	10					
2.00	10	10		803	60.3	59.2	57.5
2.45	10	10					
3.30	10	10					
4.15	10	10		802	60.3	59.2	57.5
4.52	0	0	31.10				
Total = 4.87		70.00	8.29				
Hourly ave.		14.37	1.70	801.67	60.20	59.07	57.43
Day-4							
0.00	10	10		800	59.9	58.8	57.0
0.45	10	10					
1.30	10	10					
2.00	10	10		796	60.1	59.0	57.2
2.45	10	10					
3.30	10	10					
4.15	0	0	37.90				
Total = 4.25		60.00	6.80				
Hourly ave.		14.12	1.60	798.00	60.00	58.90	57.10
Overall ave.							
		14.30	1.67	800.29	60.16	59.04	57.31

Appendix D: Results of ANOVA and t-Test

Table D-1: Analysis of variance (ANOVA) to compare the water evaporation rate at different airflows

Leachate run	Number of Observations	Mean evap. rate (lit/hr)	Std. Deviation	Std. Error	95% Confidence interval of mean (lit/hr)		Minimum	Maximum
					Lower Bound	Upper bound		
LR-8	8	14.90	0.79	0.28	14.24	15.56	13.60	16.00
LR-9	8	14.40	0.48	0.17	14.00	17.80	13.60	15.20
LR-10	8	14.10	0.28	0.10	13.86	14.34	13.60	14.40
Total	24	14.67	0.63	0.13	14.20	14.73	13.60	16.00

	Sum of squares	df	Mean squares	F	P-value
Between groups	2.61	2	1.31	4.18	0.03
Within groups	6.56	21	0.31		
Total	9.17	23			

Note: Mean water evaporation rates are considered to be different if P-value is less than 0.05

Table D-2: Analysis of variance (ANOVA) to compare the water evaporation rate at different evaporator temperatures

Leachate run	Number of Observations	Mean evap. rate (lit/hr)	Std. Deviation	Std. Error	95% Confidence interval of mean (lit/hr)		Minimum	Maximum
					Lower Bound	Upper bound		
LR-8	8	14.90	0.79	0.28	14.24	15.56	13.60	16.00
LR-9	8	14.40	0.48	0.17	14.00	17.80	13.60	15.20
LR-10	8	14.10	0.28	0.10	13.86	14.34	13.60	14.40
Total	24	14.67	0.63	0.13	14.20	14.73	13.60	16.00

	Sum of squares	df	Mean squares	F	P-value
Between groups	2.61	2	1.31	4.18	0.03
Within groups	6.56	21	0.31		
Total	9.17	23			

Note: Mean water evaporation rates are considered to be different if P-value is less than 0.05

Table D-3: t-Test to compare water and leachate evaporation rates using different nozzles

Experimental run	Parameter	Nozzle type	Ave. evaporation rate (lit/hr)	t-test	t-critical	Results
WR-2	Water	GSS10	13.45	0.775	2.228	No difference
LR-7	Leachate		13.06			
WR-8	Water	L66	14.90	1.687	2.228	No difference
LR-1	Leachate		14.20			
WR-6	Water	L54	15.55	2.192	2.228	No difference
LR-6	Leachate		14.73			

Table D-4: t-Test to compare water and leachate evaporation rates at different temperatures

Experimental run	Parameter	Temperature	Ave. evaporation rate (lit/hr)	t-test	t-critical	Results
WR-12	Water	600	9.40	0.346	2.570	No difference
LR-3	Leachate		9.31			
WR-8	Water	800	14.90	1.687	2.228	No difference
LR-1	Leachate		14.20			
WR-13	Water	1000	20.55	0.858	2.262	No difference
LR-2	Leachate		19.99			

Note: Averages of the water and leachate evaporation rates are considered to be different if t-test value is greater than t-critical value

Appendix E: Results of leachate samples

Table E-1: Leachate run 1 (Batch run @ 800°C, Nozzle L66, pH = 7.55)

Parameter	units	Waste stream sample (liquid phase)		
		Raw leachate	Concentrate	Condensate
pH	pH	7.55	8.79	8.89
TSS	mg/l	27.5	147.5	23.8
COD	mg/l	721	1905	446
NH ₃	mg/l	640	10	221
TKN	mg/l	807	140	255
Cl ⁻	mg/l	1896	12720	794
B	mg/l	7.1	40.0	2.4
Ba	mg/l	1.0	0.3	0.1
H ₂ S	mg/l	1.6	0.4	0.1
Toluene	µg/L	102.2	nd	4.6
m/p-xylene	µg/L	63.2	nd	2.8

Table E-2: Leachate run 2 (Batch run @ 1000°C, Nozzle L66)

Parameter	units	Waste stream sample (liquid phase)		
		Raw leachate	Concentrate	Condensate
pH	pH	7.24	8.73	8.91
TSS	mg/l	36.8	172.2	25.3
COD	mg/l	1090	2635	651
NH ₃	mg/l	548	8	121
TKN	mg/l	664	128	280
Cl ⁻	mg/l	1905	11380	487
B	mg/l	7.1	42.3	2.5
Ba	mg/l	0.9	0.4	0.1
H ₂ S	mg/l	6.0	0.3	0.1
Toluene	µg/L	114.0	nd	1.3
m/p-xylene	µg/L	73.3	nd	2.3

Table E-3: Leachate run 3 (Batch run @ 600°C, Nozzle L66)

Parameter	units	Waste stream sample (liquid phase)		
		Raw leachate	Concentrate	Condensate
pH	pH	7.5	8.72	8.58
TSS	mg/l	26.7	167.3	23.0
COD	mg/l	1170	3565	829
NH ₃	mg/l	551	20	222
TKN	mg/l	682	280	188
Cl ⁻	mg/l	1819	13350	866
B	mg/l	7.3	44.3	4.2
Ba	mg/l	1.0	0.3	0.1
H ₂ S	mg/l	0.20	0.08	0.03
Toluene	µg/L	99.6	nd	6.1
m/p-xylene	µg/L	66.3	nd	3.3

Table E-4: Leachate run 4 (Batch run at 800°C, Nozzle L66, pH = 5.5)

Parameter	units	Waste stream sample (liquid phase)		
		Raw leachate	Concentrate	Condensate
pH	pH	5.5	6.78	8.25
TSS	mg/l	24.6	121.0	19.4
COD	mg/l	662	1630	404
NH ₃	mg/l	453	1899	275
TKN	mg/l	725	1900	270
Cl ⁻	mg/l	1916	23000	844
B	mg/l	7.7	39.6	3.6
Ba	mg/l	0.8	8.4	0.8
H ₂ S	mg/l	0.05	0.30	0.08
Toluene	µg/L	57.9	nd	1.5
m/p-xylene	µg/L	59.1	nd	1.9

Table E-5: Leachate run 5 (Batch run at 800°C, Nozzle 66, pH = 3.5)

Parameter	units	Waste stream sample (liquid phase)		
		Raw leachate	Concentrate	Condensate
pH	pH	3.5	5.86	7.62
TSS	mg/l	36.8	207.4	31.3
COD	mg/l	681	1355	456
NH ₃	mg/l	433	3458	329
TKN	mg/l	708	3010	322
Cl ⁻	mg/l	1966	25825	929
B	mg/l	7.1	41.7	3.9
Ba	mg/l	0.8	8.3	0.7
H ₂ S	mg/l	0.08	0.28	0.05
Toluene	µg/L	58.7	nd	2.0
m/p-xylene	µg/L	43.2	nd	2.2

Table E-6: Leachate run 6 (Batch run at 800°C, Nozzle 54)

Parameter	units	Waste stream sample (liquid phase)		
		Raw leachate	Concentrate	Condensate
pH	pH	7.54	8.66	8.79
TSS	mg/l	40.8	174.2	36.1
COD	mg/l	569	1170	394
NH ₃	mg/l	544	3	177
TKN	mg/l	604	126	186
Cl ⁻	mg/l	1480	9455	1009
B	mg/l	6.36	30.6	3.79
Ba	mg/l	1.09	0.27	0.16
H ₂ S	mg/l	1.18	0.55	0.06
Toluene	µg/L	53.7	nd	1.9
m/p-xylene	µg/L	42.8	nd	2.3

Table E-7: Leachate run 7 (Batch run at 800°C, Nozzle GSS10)

Parameter	units	Waste stream sample (liquid phase)		
		Raw leachate	Concentrate	Condensate
pH	pH	7.28	8.69	8.81
TSS	mg/l	35.4	198.31	19.4
COD	mg/l	543	1530	315
NH₃	mg/l	421	8	191
TKN	mg/l	506	134	214
Cl⁻	mg/l	1550	12015	626
B	mg/l	6.07	26.1	1.72
Ba	mg/l	1.2	1.09	0.07
H₂S	mg/l	0.80	0.45	0.05
Toluene	µg/L	50.8	nd	3.9
m/p-xylene	µg/L	33.4	nd	1.5

Table E-8: Leachate run 8 (LR-8); Continuous run at 800°C, Nozzle GSS6.5

Sample	units	Raw leachate		Concentrate (time)			Condensate (time)	
		(sample 1)	(sample 2)	(8 hrs)	(16 hrs)	(23 hrs)	(8 hrs)	(16 hrs)
PH	pH	7.59	7.6	8.66	8.72	8.98	8.86	8.99
TSS	mg/l	29.8	34.3	81.6	216.4	740.2	17.7	31.9
COD	mg/l	803	770	1180	1565	3090	383	706
NH ₃	mg/l	582	589	133	96	82	70	46
TKN	mg/l	1380	1270	533	571	818	277	293
Cl	mg/l	1754	1660	8220	14330	28295	507	1207
B	mg/l	18.7	19.1	77.1	82.6	120.0	11.3	19.2
Ba	mg/l	1.1	1.2	0.9	0.6	0.6	0.1	0.1
H ₂ S	mg/l	0.4	0.9	0.7	1.0	3.1	0.1	0.2
Toluene	(µg/L)	75.1	58.6	nd	nd	nd	nd	nd
m/p-xylene	(µg/L)	59.1	41.6	nd	nd	nd	1.8	0.7

Table E-9: Gas phase analysis batch type leachate runs runs

Leachate run	H ₂ S (mg/l)	Toluene (µg/lit)	m/p-xylene (µg/lit)	Ammonia (at diff. time intervals)			
				(30 min) (mg/l)	(60 min) (mg/l)	(90 min) (mg/l)	(120 min) (mg/l)
LR-1	nd	11.21	6.13	370	70	20	nd
LR-2	nd	12.32	7.01	400	40	nd	nd
LR-3	nd	10.55	6.05	70	260	50	nd
LR-4	nd	7.80	7.92	80	80	90	100
LR-5	nd	9.49	5.10	50	50	50	70
LR-6	nd	11.35	6.08	350	50	nd	nd
LR-7	nd	6.65	6.25	220	60	20	nd

Table E-10: Gas phase analysis for continuous leachate run

Run time (hrs)	Toluene (µg/L)	m/p-xylene (µg/L)	H ₂ S (mg/l)	Ammonia (mg/l)
0	0.81	0.46	nd	340
8	nd	nd	nd	50
16	nd	nd	nd	60

Appendix F: Duplicate sample analysis

Table F-1: Duplicate analysis for TSS

Run	Sample name	Sample	Concentration (mg/lit)	Mean of duplicates (mg/lit)	Variability of duplicates
1	Raw leachate	1	27.8	27.5	0.02
		2	27.3		
	Concentrate	1	148.2	147.5	0.01
		2	146.8		
	Condensate	1	23.9	23.8	0.01
		2	23.7		
2	Raw leachate	1	36.8	36.8	0.00
		2	36.9		
	Concentrate	1	173.2	172.2	0.01
		2	171.2		
	Condensate	1	25.2	25.3	0.01
		2	25.4		
3	Raw leachate	1	26.9	26.7	0.01
		2	26.5		
	Concentrate	1	162.8	167.3	0.05
		2	171.9		
	Condensate	1	22.8	23.0	0.02
		2	23.2		
4	Raw leachate	1	24.4	24.6	0.02
		2	24.9		
	Concentrate	1	122.6	121.0	0.03
		2	119.5		
	Condensate	1	18.8	19.4	0.07
		2	20.1		
5	Raw leachate	1	35.3	36.8	0.08
		2	38.4		
	Concentrate	1	203.0	207.4	0.04
		2	211.8		
	Condensate	1	25.8	31.3	0.35
		2	36.8		
6	Raw leachate	1	40.0	40.8	0.04
		2	41.7		
	Concentrate	1	169.0	174.2	0.06
		2	179.4		
	Condensate	1	36.9	36.1	0.04
		2	35.4		
7	Raw leachate	1	34.7	35.4	0.04
		2	36.2		
	Concentrate	1	197.4	198.3	0.01
		2	199.2		

Table F-1 continued

Run	Sample name	Sample	Concentration (mg/lit)	Mean of duplicates (mg/lit)	Variability of duplicates
8	Condensate	1	17.8	19.4	0.16
		2	21.0		
	Raw leachate 1	1	28.6	29.8	0.08
		2	31.0		
	Raw leachate 2	1	35.3	34.3	0.06
		2	33.4		
	Concentrate 1	1	84.2	81.6	0.06
		2	79.0		
	Concentrate 2	1	214.4	216.4	0.02
		2	218.5		
	Concentrate 3	1	739.7	740.2	0.00
		2	740.8		
	Condensate 1	1	17.0	17.7	0.08
		2	18.5		
	Condensate 2	1	32.4	31.9	0.03
		2	31.3		

Table F-2: Duplicate analysis for COD

Run	Sample name	Sample	Concentration (mg/lit)	Mean of duplicates (mg/lit)	Variability of duplicates
1	Raw leachate	1	721	721	0.00
		2	721		
	Concentrate	1	1940	1905	0.04
		2	1870		
	Condensate	1	446	446	0.00
		2	446		
2	Raw leachate	1	1090	1090	0.00
		2	1090		
	Concentrate	1	2640	2635	0.00
		2	2630		
	Condensate	1	654	651	0.01
		2	648		
3	Raw leachate	1	1170	1170	0.00
		2	1170		
	Concentrate	1	3570	3565	0.00
		2	3560		
	Condensate	1	826	829	0.01
		2	832		
4	Raw leachate	1	665	662	0.01
		2	658		
	Concentrate	1	1620	1630	0.01
		2	1640		
	Condensate	1	405	404	0.01
		2	402		
5	Raw leachate	1	677	681	0.01
		2	684		
	Concentrate	1	1360	1355	0.01
		2	1350		
	Condensate	1	453	456	0.01
		2	458		
6	Raw leachate	1	557	569	0.04
		2	580		
	Concentrate	1	1160	1170	0.02
		2	1180		
	Condensate	1	390	394	0.02
		2	398		
7	Raw leachate	1	550	543	0.03
		2	535		
	Concentrate	1	1530	1530	0.00
		2	1530		

Table F-2 continued

Run	Sample name	Sample	Concentration (mg/lit)	Mean of duplicates (mg/lit)	Variability of duplicates
8	Condensate	1	313	315	0.01
		2	316		
	Raw leachate 1	1	803	803	0.00
		2	803		
	Raw leachate 2	1	773	770	0.01
		2	766		
	Concentrate 1	1	1180	1180	0.00
		2	1180		
	Concentrate 2	1	1560	1565	0.01
		2	1570		
	Concentrate 3	1	3090	3090	0.00
		2	3090		
	Condensate 1	1	383	383	0.00
		2	383		
	Condensate 2	1	706	706	0.00
		2	706		

Table F-3: Duplicate analysis for NH3

Run	Sample name	Sample	Concentration (mg/lit)	Mean of duplicates (mg/lit)	Variability of duplicates
1	Raw leachate	1	641	640	0.00
		2	639		
	Concentrate	1	10	10	0.01
		2	10		
	Condensate	1	220	221	0.01
		2	222		
2	Raw leachate	1	550	548	0.01
		2	545		
	Concentrate	1	8	8	0.01
		2	8		
	Condensate	1	121	121	0.00
		2	120		
3	Raw leachate	1	552	551	0.00
		2	550		
	Concentrate	1	20	20	0.00
		2	20		
	Condensate	1	221	222	0.01
		2	223		
4	Raw leachate	1	453	453	
		2			
	Concentrate	1	1899	1899	
		2			
	Condensate	1	275	275	
		2			
5	Raw leachate	1	433	433	
		2			
	Concentrate	1	3458	3458	
		2			
	Condensate	1	329	329	
		2			
6	Raw leachate	1	543	544	0.00
		2	545		
	Concentrate	1	3	3	0.00
		2	3		
	Condensate	1	177	177	0.00
		2	176		
7	Raw leachate	1	421	421	0.00
		2	421		
	Concentrate	1	8	8	0.00
		2	8		

Table F-3 continued

Run	Sample name	Sample	Concentration (mg/lit)	Mean of duplicates (mg/lit)	Variability of duplicates
8	Condensate	1	191	191	0.00
		2	192		
	Raw leachate 1	1	584	582	0.01
		2	580		
	Raw leachate 2	1	587	589	0.01
		2	591		
	Concentrate 1	1	133	133	0.00
		2	133		
	Concentrate 2	1	96	96	0.00
		2	96		
	Concentrate 3	1	81	82	0.01
		2	82		
	Condensate 1	1	70	70	0.00
		2	70		
	Condensate 2	1	46	46	0.01
		2	45		

Table F-4: Duplicate analysis for Chloride

Run	Sample name	Sample	Concentration (mg/lit)	Mean of duplicates (mg/lit)	Variability of duplicates
1	Raw leachate	1	1896	1896	
		2			
	Concentrate	1	12720	12720	
		2			
	Condensate	1	794	794	
		2			
2	Raw leachate	1	1905	1905	
		2			
	Concentrate	1	11380	11380	
		2			
	Condensate	1	487	487	
		2			
3	Raw leachate	1	1819	1819	
		2			
	Concentrate	1	13350	13350	
		2			
	Condensate	1	866	866	
		2			
4	Raw leachate	1	1914	1916	0.00
		2	1917		
	Concentrate	1	24790	23000	0.16
		2	21210		
	Condensate	1	856	844	0.03
		2	832		
5	Raw leachate	1	1984	1966	0.02
		2	1948		
	Concentrate	1	25890	25825	0.01
		2	25760		
	Condensate	1	930	929	0.00
		2	927		
6	Raw leachate	1	1470	1480	0.01
		2	1490		
	Concentrate	1	9315	9455	0.03
		2	9595		
	Condensate	1	1052	1009	0.09
		2	965		
7	Raw leachate	1	1578	1550	0.04
		2	1522		
	Concentrate	1	11400	12015	0.10
		2	12630		

Table F-4 continued

Run	Sample name	Sample	Concentration (mg/lit)	Mean of duplicates (mg/lit)	Variability of duplicates
8	Condensate	1	627	626	0.00
		2	625		
	Raw leachate 1	1	1771	1754	0.02
		2	1737		
	Raw leachate 2	1	1624	1660	0.04
		2	1696		
	Concentrate 1	1	8220	8220	0.00
		2	8220		
	Concentrate 2	1	14310	14330	0.00
		2	14350		
	Concentrate 3	1	27850	28295	0.03
		2	28740		
	Condensate 1	1	507	507	0.00
		2	506		
	Condensate 2	1	1214	1207	0.01
		2	1199		

Table F-5: Duplicate analysis for toluene

Run	Sample name	Sample	Concentration (µg/lit)	Mean of duplicates (µg/lit)	Variability of duplicates
1	Raw leachate	1	99.1	102.2	0.06
		2	105.4		
	Concentrate	1	0.0	0.0	
		2	0.0		
	Condensate	1	4.6	4.6	0.01
		2	4.6		
	Air	1	10.2	11.2	0.18
		2	12.2		
2	Raw leachate	1	112.3	114.0	0.03
		2	115.6		
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	1.3	1.3	
		2			
	Air	1	12.3	12.3	
		2			
3	Raw leachate	1	99.7	99.6	0.00
		2	99.4		
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	5.9	6.1	0.07
		2	6.3		
	Air	1	7.5	10.6	0.58
		2	13.6		
4	Raw leachate	1	57.9	57.9	
		2			
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	1.5	1.5	
		2			
	Air	1	7.8	7.8	
		2			
5	Raw leachate	1	58.7	58.7	
		2			
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	2.0	2.0	
		2			
	Air sample	1	9.4	9.5	0.03
		2	9.6		

Table F-5 continued

Run	Sample name	Sample	Concentration (µg/lit)	Mean of duplicates (µg/lit)	Variability of duplicates
6	Raw leachate	1	56.9	53.7	0.12
		2	50.6		
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	1.9	1.9	
		2			
	Air	1	11.4	11.4	
		2			
7	Raw leachate	1	52.4	50.8	0.07
		2	49.1		
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	3.4	3.9	0.27
		2	4.5		
	Air	1	6.7	6.7	
		2			
8	Raw leachate 1	1	75.1	75.1	
		2			
	Raw leachate 2	1	55.7	58.6	0.10
		2	61.6		
	Concentrate 1	1	0.0	0.0	
		2			
	Concentrate 2	1	0.0	0.0	
		2			
	Concentrate 3	1	0.0	0.0	
		2			
	Condensate 1	1	0.0	0.0	
		2			
	Condensate 2	1	0.0	0.0	
		2			
	Air sample 1	1	0.8	0.8	0.00
		2	0.8		
	Air sample 2	1	0.0	0.0	
		2			
	Air sample 3	1	0.0	0.0	
		2			

Table F-6: Duplicate analysis for m/p-xylene

Run	Sample name	Sample	Concentration (µg/lit)	Mean of duplicates (µg/lit)	Variability of duplicates
1	Raw leachate	1	61.5	63.2	0.05
		2	64.8		
	Concentrate	1	0.0	0.0	
		2	0.0		
	Condensate	1	2.7	2.8	0.03
		2	2.8		
	Air	1	6.4	6.1	0.07
		2	5.9		
2	Raw leachate	1	68.3	73.3	0.14
		2	78.4		
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	2.3	2.3	
		2			
	Air	1	7.0	7.0	
		2			
3	Raw leachate	1	65.5	66.3	0.03
		2	67.2		
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	2.8	3.3	0.28
		2	3.7		
	Air	1	5.7	6.0	0.12
		2	6.4		
4	Raw leachate	1	59.1	59.1	
		2			
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	1.9	1.9	
		2			
	Air	1	7.9	7.9	
		2			
5	Raw leachate	1	43.2	43.2	
		2			
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	2.2	2.2	
		2			
	Air sample	1	4.8	5.1	0.14
		2	5.5		

Table F-6 continued

Run	Sample name	Sample	Concentration (µg/lit)	Mean of duplicates (µg/lit)	Variability of duplicates
6	Raw leachate	1	43.5	42.8	0.04
		2	42.0		
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	2.3	2.3	
		2			
	Air	1	6.1	6.1	
		2			
7	Raw leachate	1	34.0	33.4	0.04
		2	32.8		
	Concentrate	1	0.0	0.0	
		2			
	Condensate	1	1.6	1.5	0.13
		2	1.4		
	Air	1	6.3	6.3	
		2			
8	Raw leachate 1	1	59.1	59.1	
		2			
	Raw leachate 2	1	42.5	41.6	0.04
		2	40.8		
	Concentrate 1	1	0.0	0.0	
		2			
	Concentrate 2	1	0.0	0.0	
		2			
	Concentrate 3	1	0.0	0.0	
		2			
	Condensate 1	1	1.8	1.8	
		2			
	Condensate 2	1	0.7	0.7	
		2	0.7		
	Air sample 1	1	0.5	0.5	0.09
		2	0.4		
	Air sample 2	1	0.0	0.0	
		2			
	Air sample 3	1	0.0	0.0	
		2			

Appendix G: Mass balance analysis for leachate samples

Table G-1: Mass balances for TSS

Mass balances for TSS at different temperature conditions						
	LR-3 at 600 °C			LR-1 at 800 °C		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	27	800	1.00	28	826	1.00
Accum. TSS in		800	1.00		826	1.00
Concentrate	167	502	0.63	148	443	0.54
Condensate	23	29	0.04	24	38	0.05
Accum. TSS out		531	0.66		480	0.58
Remaining TSS		269	0.34		346	0.42

Mass balances for TSS at different raw leachate pH conditions						
	LR-1 at pH = 7.5			LR-4 at pH = 5.5		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	28	826	1.00	25	738	1.00
Accum. TSS in		826	1.00		738	1.00
Concentrate	148	443	0.54	121	363	0.49
Condensate	24	38	0.05	19	31	0.04
Accum. TSS out		480	0.58		394	0.53
Remaining TSS		346	0.42		344	0.47

Mass balances for TSS using different spray nozzles						
	LR-6 with nozzle L54			LR-1 with nozzle L66		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	41	1224	1.00	28	826	1.00
Accum. TSS in		1224	1.00		826	1.00
Concentrate	174	523	0.43	148	443	0.54
Condensate	36	78	0.06	24	38	0.05
Accum. TSS out		601	0.49		480	0.58
Remaining TSS		623	0.51		346	0.42

Mass balances for TSS using different spray nozzles						
	LR-6 with nozzle L54			LR-7 with nozzle GSS10		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	41	1224	1.00	35	1062	1.00
Accum. TSS in		1224	1.00		1062	1.00
Concentrate	174	523	0.43	198	595	0.56
Condensate	36	78	0.06	19	18	0.02
Accum. TSS out		601	0.49		613	0.58
Remaining TSS		623	0.51		449	0.42

Table G-2: Mass balances for COD

Mass balances for COD at different temperature conditions						
	LR-3 at 600 °C			LR-1 at 800 °C		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	1170	35100	1.00	721	21629	1.00
Accum. COD in		35100	1.00		21629	1.00
Concentrate	3565	10695	0.30	1905	5715	0.26
Condensate	829	1044	0.03	446	709	0.03
Accum. COD out		11739	0.33		6424	0.30
Remaining COD		23361	0.67		15206	0.70

Mass balances for COD at different raw leachate pH conditions						
	LR-1 at pH = 7.5			LR-4 at pH = 5.5		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	721	21629	1.00	662	19855	1.00
Accum. COD in		21629	1.00		19855	1.00
Concentrate	1905	5715	0.26	1630	4890	0.25
Condensate	446	562	0.03	404	642	0.03
Accum. COD out		6277	0.29		5532	0.28
Remaining COD		15353	0.71		14323	0.72

Mass balances for COD using different spray nozzles						
	LR-6 with nozzle L54			LR-1 with nozzle L66		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	569	17070	1.00	721	21629	1.00
Accum. COD in		17070	1.00		21629	1.00
Concentrate	1170	3510	0.21	1905	5715	0.26
Condensate	394	851	0.05	446	709	0.03
Accum. COD out		4361	0.26		6424	0.30
Remaining COD		12709	0.74		15206	0.70

Mass balances for COD using different spray nozzles						
	LR-6 with nozzle L54			LR-7 with nozzle GSS10		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	569	17070	1.00	721	21629	1.00
Accum. COD in		17070	1.00		21629	1.00
Concentrate	1170	3510	0.21	1905	5715	0.26
Condensate	394	851	0.05	446	709	0.03
Accum. COD out		4361	0.26		6424	0.30
Remaining COD		12709	0.74		15206	0.70

Table G-3: Mass balances for TKN

Mass balances for TKN at different temperature conditions						
	LR-3 at 600 °C			LR-1 at 800 °C		LR-2 at 1000 °C
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	682	20460	1.00	807	24210	1.00
Accum. TKN in		20460	1.00		24210	1.00
Concentrate	280	840	0.04	140	420	0.02
Condensate	188	237	0.01	255	405	0.02
Accum. TKN out		1077	0.05		825	0.03
Remaining TKN		19383	0.95		23385	0.97

Mass balances for TKN at different raw leachate pH conditions						
	LR-1 at pH = 7.5			LR-4 at pH = 5.5		LR-5 at pH = 3.5
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	807	24210	1.00	725	21750	1.00
Accum. TKN in		24210	1.00		21750	1.00
Concentrate	140	420	0.02	1900	5700	0.26
Condensate	255	405	0.02	270	429	0.02
Accum. TKN out		825	0.03		6129	0.28
Remaining TKN		23385	0.97		15621	0.72

Mass balances for TKN using different spray nozzles						
	LR-6 with nozzle L54			LR-1 with nozzle L66		LR-7 with nozzle GSS10
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	604	18120	1.00	807	24210	1.00
Accum. TKN in		18120	1.00		24210	1.00
Concentrate	126	378	0.02	140	420	0.02
Condensate	186	402	0.02	255	405	0.02
Accum. TKN out		780	0.04		825	0.03
Remaining TKN		17340	0.96		23385	0.97

Table G-4: Mass balances for ammonia

Mass balances for ammonia at different temperature conditions									
	LR-3 at 600 °C			LR-1 at 800 °C			LR-2 at 1000 °C		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	551	16529	1.00	640	19201	1.00	548	16429	1.00
Accum. ammonia in		16529	1.00		19201	1.00		16429	1.00
Concentrate	20	60	0.00	10	30	0.00	8	24	0.00
Condensate	222	280	0.02	221	351	0.02	121	327	0.02
Exhaust air									
30 min	70	2491	0.15	370	13165	0.69	400	14232	0.87
60 min	260	9251	0.56	70	2491	0.13	40	1423	0.09
90 min	50	1779	0.11	20	712	0.04	nd		
120 min	nd			nd			na		
Accum. ammonia out		13860	0.84		16749	0.87		16006	0.97
Remaining ammonia		2669	0.16		2452	0.13		423	0.03

Mass balances for ammonia at different raw leachate pH conditions									
	LR-1 at pH = 7.5			LR-4 at pH = 5.5			LR-5 at pH = 3.5		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	640	19201	1.00	453	13581	1.00	433	12990	1.00
Accum. ammonia in		19201	1.00		13581	1.00		12990	1.00
Concentrate	10	30	0.00	1899	5696	0.42	3458	10373	0.80
Condensate	221	351	0.02	275	437	0.03	329	414	0.03
Exhaust air									
30 min	370	13165	0.69	80	2846	0.21	50	1779	0.14
60 min	70	2491	0.13	80	2846	0.21	50	1779	0.14
90 min	20	712	0.04	90	3202	0.24	50	1779	0.14
120 min	nd			100	3558	0.62	70	2491	0.24
Accum. ammonia out		16749	0.87		15029	1.11		16124	1.24
Remaining ammonia		2452	0.13		-1447	-0.11		-3135	-0.24

Table G-5: Continued

	LR-6 with nozzle L54			LR-1 with nozzle L66			LR-7 with nozzle GSS10		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	544	16329	1.00	640	19201	1.00	421	12626	1.00
Accum. ammonia in		16329	1.00		19201	1.00		12626	1.00
Concentrate	3	10	0.00	10	30	0.00	8	24	0.00
Condensate	177	381	0.02	221	351	0.02	191	176	0.01
Exhaust air									
30 min	350	12453	0.76	370	13165	0.69	220	7828	0.62
60 min	50	1779	0.11	70	2491	0.13	60	2135	0.17
90 min	nd			20	712	0.04	20	712	0.06
120 min	nd			nd			nd		
Accum. ammonia out		14623	0.90		16749	0.87		10874	0.86
Remaining ammonia		1706	0.10		2452	0.13		1753	0.14

Table G-5: Mass balances for chlorides

Mass balances for chlorides at different temperature conditions						
	LR-3 at 600 °C		LR-1 at 800 °C		LR-2 at 1000 °C	
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	1819	54570	1.00	1896	56880	1.00
Accum. chlorides in		54570	1.00		56880	1.00
Concentrate	13350	40050	0.73	12720	38160	0.67
Condensate	866	1091	0.02	794	1262	0.02
Accum. chlorides out		41141	0.75		39422	0.69
Remaining chlorides		13429	0.25		17458	0.31

Mass balances for chlorides at different raw leachate pH conditions						
	LR-1 at pH = 7.5			LR-4 at pH = 5.5		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	1896	56880	1.00	5653	169590	1.00
Accum. chlorides in		56880	1.00		169590	1.00
Concentrate	12720	38160	0.67	23000	69000	0.41
Condensate	794	1262	0.02	844	1342	0.01
Accum. chlorides out		39422	0.69		70342	0.41
Remaining chlorides		17458	0.31		99248	0.59

Mass balances for chlorides using different spray nozzles						
	LR-6 with nozzle L54			LR-1 with nozzle L66		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	1480	44400	1.00	1896	56880	1.00
Accum. chlorides in		44400	1.00		56880	1.00
Concentrate	9455	28365	0.64	12720	38160	0.67
Condensate	1009	2178	0.05	794	1262	0.02
Accum. chlorides out		30543	0.69		39422	0.69
Remaining chlorides		13857	0.31		17458	0.31

Mass balances for chlorides using different spray nozzles						
	LR-6 with nozzle L54			LR-7 with nozzle GSS10		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	1480	44400	1.00	1550	46500	1.00
Accum. chlorides in		44400	1.00		46500	1.00
Concentrate	9455	28365	0.64	12015	36045	0.78
Condensate	1009	2178	0.05	626	576	0.01
Accum. chlorides out		30543	0.69		36621	0.79
Remaining chlorides		13857	0.31		9879	0.21

Table G-6: Mass balances for boron

Mass balances for boron at different temperature conditions									
	LR-3 at 600 °C			LR-1 at 800 °C			LR-2 at 1000 °C		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	7	218	1.00	7	212	1.00	7	213	1.00
Accum. boron in		218	1.00		212	1.00		213	1.00
Concentrate	44	133	0.61	40	120	0.57	42	127	0.60
Condensate	4	5	0.02	2	4	0.02	2	7	0.03
Accum. boron out		138	0.64		124	0.59		134	0.63
Remaining boron		79	0.36		88	0.41		79	0.37

Mass balances for boron at different raw leachate pH conditions									
	LR-1 at pH = 7.5			LR-4 at pH = 5.5			LR-5 at pH = 3.5		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	7	212	1.00	8	231	1.00	7	213	1.00
Accum. boron in		212	1.00		231	1.00		213	1.00
Concentrate	40	120	0.57	40	119	0.51	42	125	0.59
Condensate	2	4	0.02	4	6	0.03	4	5	0.02
Accum. boron out		124	0.59		125	0.54		130	0.61
Remaining boron		88	0.41		107	0.46		83	0.39

Mass balances for boron using different spray nozzles									
	LR-6 with nozzle L54			LR-1 with nozzle L66			LR-7 with nozzle GSS10		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	6	191	1.00	7	212	1.00	6	182	1.00
Accum. boron in		191	1.00		212	1.00		182	1.00
Concentrate	31	92	0.48	40	120	0.57	26	78	0.43
Condensate	4	8	0.04	2	4	0.02	2	2	0.01
Accum. boron out		100	0.52		124	0.59		80	0.44
Remaining boron		91	0.48		88	0.41		102	0.56

Table G-7: Mass balances for barium

Mass balances for barium at different temperature conditions

	LR-3 at 600 °C			LR-1 at 800 °C			LR-2 at 1000 °C		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	0.96	29	1.00	0.98	29	1.00	0.92	28	1.00
Accum. barium in		29	1.00		29	1.00		28	1.00
Concentrate	0.33	1	0.03	0.27	1	0.03	0.42	1	0.05
Condensate	0.10	0	0.00	0.07	0	0.00	0.08	0	0.01
Accum. barium out		1	0.04		1	0.03		1	0.05
Remaining barium		28	0.96		28	0.97		26	0.95

Mass balances for barium at different raw leachate pH conditions

	LR-1 at pH = 7.5			LR-4 at pH = 5.5			LR-5 at pH = 3.5		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	0.98	29	1.00	0.83	25	1.00	0.77	23	1.00
Accum. barium in		29	1.00		25	1.00		23	1.00
Concentrate	0.27	1	0.03	8.42	25	1.01	8.33	25	1.08
Condensate	0.07	0	0.00	0.79	1	0.05	0.72	1	0.04
Accum. barium out		1	0.03		27	1.06		26	1.12
Remaining barium		28	0.97		-2	-0.06		-3	-0.12

Mass balances for barium using different spray nozzles

	LR-6 with nozzle L54			LR-1 with nozzle L66			LR-7 with nozzle GSS10		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	1.09	33	1.00	0.98	29	1.00	1.20	36	1.00
Accum. barium in		33	1.00		29	1.00		36	1.00
Concentrate	0.27	1	0.02	0.27	1	0.03	1.09	3	0.09
Condensate	0.16	0	0.01	0.07	0	0.00	0.07	0	0.00
Accum. barium out		1	0.04		1	0.03		3	0.09
Remaining barium		32	0.96		28	0.97		33	0.91

Table G-8: Mass balances for H₂S

Mass balances for barium at different temperature conditions									
	LR-3 at 600 °C			LR-1 at 800 °C			LR-2 at 1000 °C		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	0.20	6.0	1.00	1.62	48.6	1.00	5.95	178.5	1.00
Accum. H₂S in		6.0	1.00		48.6	1.00		178.5	1.00
Concentrate	0.08	0.2	0.04	0.44	1.3	0.03	0.25	0.8	0.004
Condensate	0.03	0.0	0.01	0.08	0.1	0.00	0.05	0.1	0.001
Accum. H₂S out		0.3	0.05		1.4	0.03		0.9	0.010
Remaining H₂S		5.7	0.95		47.2	0.97		177.6	0.99

Mass balances for barium at different raw leachate pH conditions									
	LR-1 at pH = 7.5			LR-4 at pH = 5.5			LR-5 at pH = 3.5		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	1.62	48.6	1.00	0.05	1.5	1.00	0.08	2.4	1.00
Accum. H₂S in		48.6	1.00		1.5	1.00		2.4	1.00
Concentrate	0.44	1.3	0.03	0.30	0.9	0.60	0.28	0.8	0.35
Condensate	0.08	0.1	0.00	0.08	0.1	0.08	0.05	0.1	0.03
Accum. H₂S out		1.4	0.03		1.0	0.68		0.9	0.38
Remaining H₂S		47.2	0.97		0.5	0.32		1.5	0.62

Mass balances for barium using different spray nozzles									
	LR-6 with nozzle L54			LR-1 with nozzle L66			LR-7 with nozzle GSS10		
	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction	Conc. (mg/lit)	Total mass (mg)	Mass fraction
Raw leachate	1.18	35.4	1.00	1.62	48.6	1.00	0.80	24.0	1.00
Accum. H₂S in		35.4	1.00		48.6	1.00		24.0	1.00
Concentrate	0.55	1.7	0.05	0.44	1.3	0.03	0.45	1.4	0.06
Condensate	0.06	0.1	0.00	0.08	0.1	0.00	0.05	0.0	0.00
Accum. H₂S out		1.8	0.05		1.4	0.03		1.4	0.06
Remaining H₂S		33.6	0.95		47.2	0.97		22.6	0.94

Table G-9: Mass balances for toluene

Mass balances for toluene at different temperature conditions									
	LR-3 at 600 °C			LR-1 at 800 °C			LR-2 at 1000 °C		
	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction
Raw leachate	100	2986.86	1.00	102	3067	1.00	114	3419	1.00
Accum. toluene in		2986.9	1.00		3067	1.00		3419	1.00
Condensate	6	8	0.00	5	7	0.00	1	3	0.00
Exhaust air	11	1501	0.50	11	1595	0.52	12	1753	0.51
Accum. toluene out		1509	0.51		1602	0.52		1757	0.51
Remaining toluene		1478	0.49		1465	0.48		1662	0.49

Mass balances for toluene at different raw leachate pH conditions									
	LR-1 at pH = 7.5			LR-4 at pH = 5.5			LR-5 at pH = 3.5		
	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction
Raw leachate	102	3067	1.00	58	1738	1.00	59	1762	1.00
Accum. toluene in		3067	1.00		1738	1.00		1762	1.00
Condensate	5	6	0.00	2	2	0.00	2	5	0.00
Exhaust air	11	1595	0.52	8	1110	0.64	9	1351	0.77
Accum. toluene out		1600	0.52		1113	0.64		1356	0.77
Remaining toluene		1467	0.48		625	0.36		405	0.23

Mass balances for toluene using different spray nozzles									
	LR-6 with nozzle L54			LR-1 with nozzle L66			LR-7 with nozzle GSS10		
	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction
Raw leachate	54	1612	1.00	102	3067	1.00	51	1523	1.00
Accum. toluene in		1612	1.00		3067	1.00		1523	1.00
Condensate	2	2	0.00	5	7	0.00	4	11	0.01
Exhaust air	11	1615	1.00	11	1595	0.52	7	946	0.62
Accum. toluene out		1617	1.00		1602	0.52		957	0.63
Remaining toluene		-5	0.00		1465	0.48		566	0.37

Table G-10: Mass balances for m/p-xylene

Mass balances for m/p-xylene at different temperature conditions									
	LR-3 at 600 °C			LR-1 at 800 °C			LR-2 at 1000 °C		
	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction
Raw leachate	66	1990	1.00	63	1895	1.00	73	2199	1.00
Accum. m/p-xylene in		1990	1.00		1895	1.00		2199	1.00
Condensate	3	4	0.00	3	4	0.00	2	6	0.00
Exhaust air	6	860	0.43	6	873	0.46	7	998	0.45
Accum. m/p-xylene out		864	0.43		877	0.46		1004	0.46
Remaining m/p-xylene		1126	0.57		1018	0.54		1195	0.54

Mass balances for m/p-xylene at different raw leachate pH conditions									
	LR-1 at pH = 7.5			LR-4 at pH = 5.5			LR-5 at pH = 3.5		
	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction
Raw leachate	63	1895	1.00	59	1773	1.00	43	1297	1.00
Accum. m/p-xylene in		1895	1.00		1773	1.00		1297	1.00
Condensate	3	3	0.00	2	3	0.00	2	6	0.00
Exhaust air	6	872	0.46	8	1127	0.64	5	726	0.56
Accum. m/p-xylene out		876	0.46		1130	0.64		732	0.56
Remaining m/p-xylene		1019	0.54		643	0.36		565	0.44

Mass balances for m/p-xylene using different spray nozzles									
	LR-6 with nozzle L54			LR-1 with nozzle L66			LR-7 with nozzle GSS10		
	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction	Conc. (µg/lit)	Total mass (µg)	Mass fraction
Raw leachate	43	1283	1.00	63	1895	1.00	33	1002	1.00
Accum. m/p-xylene in		1283	1.00		1895	1.00		1002	1.00
Condensate	2	3	0.00	3	4	0.00	2	4	0.00
Exhaust air	6	865	0.67	6	873	0.46	6	889	0.89
Accum. m/p-xylene out		868	0.68		877	0.46		893	0.89
Remaining m/p-xylene		415	0.32		1018	0.54		109	0.11

Appendix H: Airflow rate calculations

Table H-1: Air flow conversions

ft/min	air flow conversion		
	SCFM	kg/hr	m3/hr
600	52.33	109.38	88.95
580	50.59	105.73	85.99
560	48.84	102.08	83.02
540	47.10	98.44	80.06
520	45.36	94.79	77.09
500	43.61	91.15	74.13
480	41.87	87.50	71.16
460	40.12	83.86	68.20
440	38.38	80.21	65.23
420	36.63	76.56	62.27
400	34.89	72.92	59.30
380	33.14	69.27	56.34
360	31.40	65.63	53.37
340	29.66	61.98	50.41
320	27.91	58.33	47.44
300	26.17	54.69	44.48
280	24.42	51.04	41.51
260	22.68	47.40	38.55
240	20.93	43.75	35.58
220	19.19	40.10	32.62
200	17.44	36.46	29.65
180	15.70	32.81	26.69
160	13.96	29.17	23.72
140	12.21	25.52	20.76
120	10.47	21.88	17.79
air duct dia.	4.0	in	
air duct area	0.087	ft ²	

Table H-2: Fuel flow conversions

heat in		heat value	flow rate	SG	density, air	fuel flow	
BTU/hr	kJ/s					kJ/kg	kg/hr
propane							
40,000	11.7	46360	0.91	1.5	1.23	0.49	17.4
45,000	13.2	46360	1.02	1.5	1.23	0.55	19.6
50,000	14.7	46360	1.14	1.5	1.23	0.62	21.8
55,000	16.1	46360	1.25	1.5	1.23	0.68	23.9
60,000	17.6	46360	1.37	1.5	1.23	0.74	26.1
65,000	19.0	46360	1.48	1.5	1.23	0.80	28.3
70,000	20.5	46360	1.59	1.5	1.23	0.86	30.5
75,000	22.0	46360	1.71	1.5	1.23	0.92	32.6
80,000	23.4	46360	1.82	1.5	1.23	0.99	34.8
85,000	24.9	46360	1.93	1.5	1.23	1.05	37.0
90,000	26.4	46360	2.05	1.5	1.23	1.11	39.2
95,000	27.8	46360	2.16	1.5	1.23	1.17	41.4
100,000	29.3	46360	2.28	1.5	1.23	1.23	43.5
105,000	30.8	46360	2.39	1.5	1.23	1.29	45.7
110,000	32.2	46360	2.50	1.5	1.23	1.36	47.9
115,000	33.7	46360	2.62	1.5	1.23	1.42	50.1
120,000	35.2	46360	2.73	1.5	1.23	1.48	52.2
NG							
40,000	11.7	40000	1.05	0.6	1.23	1.4	50.5
40,000	11.7	40000	1.05	0.6	1.23	1.4	50.5
45,000	13.2	40000	1.19	0.6	1.23	1.6	56.8
50,000	14.7	40000	1.32	0.6	1.23	1.8	63.1
55,000	16.1	40000	1.45	0.6	1.23	2.0	69.4
60,000	17.6	40000	1.58	0.6	1.23	2.1	75.7
65,000	19.0	40000	1.71	0.6	1.23	2.3	82.0
70,000	20.5	40000	1.85	0.6	1.23	2.5	88.3
75,000	22.0	40000	1.98	0.6	1.23	2.7	94.6
80,000	23.4	40000	2.11	0.6	1.23	2.9	100.9
85,000	24.9	40000	2.24	0.6	1.23	3.0	107.2
90,000	26.4	40000	2.37	0.6	1.23	3.2	113.5
95,000	27.8	40000	2.51	0.6	1.23	3.4	119.8
100,000	29.3	40000	2.64	0.6	1.23	3.6	126.1
105,000	30.8	40000	2.77	0.6	1.23	3.8	132.4
110,000	32.2	40000	2.90	0.6	1.23	3.9	138.7
115,000	33.7	40000	3.03	0.6	1.23	4.1	145.1
120,000	35.2	40000	3.16	0.6	1.23	4.3	151.4

Appendix I: Nozzle types, flow rates and droplet sizes

L

BETE

Low Flow

DESIGN FEATURES

- A series of small spiral nozzles with orifice diameters of 0.040" to 0.120"
- Male connection

SPRAY CHARACTERISTICS

Spray pattern: Hollow Cone Fog, nearly as fine as P Series
Spray angles: 90° standard (120° available by special order)
Flow rates: 0.14 to 3.84 gpm

Hollow Cone 90°

Flow Rates

Hollow Cone 90° Spray Angle 1/8" and 1/4" Pipe Sizes

Male Pipe Size	Nozzle Number	K Factor	GALLONS PER MINUTE @ PSI											High PSI operation recom. for Metal Only	Approx. Orifice Dia. (in.)	Male Pipe Size	Dimensions (in.)				Wt. (oz.) Metal
			10 PSI	20 PSI	30 PSI	40 PSI	50 PSI	60 PSI	70 PSI	80 PSI	90 PSI	100 PSI	A				B	D	H		
1/8 or	L40	0.044	0.14	0.20	0.24	0.28	0.31	0.34	0.37	0.40	0.42	0.44	0.040	1/8"	1.12	0.56	27	14	0.60		
	L48	0.063	0.20	0.28	0.35	0.40	0.45	0.49	0.53	0.57	0.60	0.63	0.048								
	L54	0.084	0.27	0.37	0.46	0.53	0.59	0.65	0.70	0.75	0.80	0.84	0.054								
	L66	0.119	0.38	0.53	0.65	0.75	0.84	0.92	0.99	1.06	1.13	1.19	0.066								
1/4	L80	0.171	0.54	0.76	0.94	1.08	1.21	1.32	1.43	1.53	1.62	1.71	0.080	1/4"	1.31	0.56	48	24	0.75		
	L120	0.384	1.22	1.72	2.10	2.43	2.72	2.98	3.21	3.44	3.65	3.84	0.120								

$$\text{Flow Rate (GPM)} = K \sqrt{\text{PSI}}$$

Standard Materials: Brass, 303 Stainless Steel, 316 Stainless Steel and PTFE (L40, L48, L54 not available in PTFE and 316 Stainless Steel). See chart on page 21 for complete list.

BETE

Technical Services Department
Estimated Drop Size Information

Carlton University
Application 20757
L54 at 50 psi

sg= 1
1 cp
0.594 gpm
D32=77
DV0.5=100
DV0.1=39
DV0.9=180

source: automatic

BETE

Technical Services Department
Estimated Drop Size Information

Carlton University
Application 20757
L54 at 80 psi

sg= 1
1 cp
0.751 gpm
D32=67
DV0.5=84
DV0.1=36
DV0.9=150

source: automatic

BETE

Technical Services Department
Estimated Drop Size Information

Carlton University
Application 20757
L66 at 50 psi

sg = 1
1 cp
0.841 gpm
D32 = 90
DV0.5 = 120
DV0.1 = 46
DV0.9 = 210

source: automatic

BETE

Technical Services Department
Estimated Drop Size Information

Carlton University
Application 20757
L66 at 80 psi

sg= 1
1 cp
1.06 gpm
D32=78
DV0.5=98
DV0.1=42
DV0.9=180

source: automatic