CHAPTER 6

BIOMETHANIZATION STUDIES AND APPLICATIONS

6.1 GENERAL

Bench scale studies were carried out on biomethanization using tannery effluent, combined TE and LLF and kinetic constants were arrived at. Pilot scale reactor was designed based on the bench scale studies and further studies were carried out with TE and LLF. A full scale plant was also designed based on the results obtained from pilot scale studies. Techno-economic analysis was carried out for a typical tannery cluster based on the results of the present study. The results obtained from these studies are discussed here under.

6.2 BIOMETHANIZATION STUDIES ON TE (BENCH SCALE)

6.2.1 Startup Regime of UASB Reactor with TE

6.2.1.1 Feed Details

The bench scale UASB reactor was fed with TE alone at an OLR of 2 kg/m$^3$.day initially; OLR was increased by increasing the flow. COD load applied and COD removal efficiency during the start up period is shown in Figures 6.1 and 6.2 respectively.
Within 40 days, the OLR was increased to 2.5 kg/m$^3$.day. During the initial period, COD removal efficiency was only about 40%. The COD removal efficiency increased to 50% within 60 days as the bacteria got acclimatized with the substrate.
During the startup period, methane production also increased up to 1.8 L within a period of 50 days indicating active methanogenic activity in the reactor. Methane production during startup period is shown in Figure 6.3. In this study, pH of the influent and effluent was observed during the startup period and the profile is shown in Figure 6.4. It was observed that pH of influent was in the range of 8.2-8.9 and reduced to 7-8.2 during startup.

![Figure 6.3 Methane Production during Startup](image-url1)

![Figure 6.4 pH Profile during Startup](image-url2)
6.2.1.2 VFA Profile during Startup

VFA concentration in the influent was in the range of 740-1420 mg/L. The VFA profile is shown in Figure 6.5 and it was observed that VFA concentration in the effluent was below 500 mg/L during the startup period.

![VFA Profile during Startup](image)

Figure 6.5 VFA Profile during Startup

6.2.2 Effect of OLR and HRT

After startup, regular studies were conducted varying OLR and HRT. In the UASB reactor treating TE, OLR was increased gradually to 12 kg/m³.day during study period of 450 days. As the COD concentration of TE cannot be increased, OLR was increased by increasing the flow rate into the reactor. The effect of OLR and HRT on COD removal efficiency is shown in Figures 6.6 and 6.7 respectively. It was observed that during the study period, the COD removal efficiency was in the range of 60-70% consistently during the study period. HRT was gradually reduced from 40 h to 10 h and at the same time increasing the OLR. This has not adversely affected the performance of the reactor indicating that the reactor can be operated at 12
kg/m$^3$.day. However, a slight drop in COD removal efficiency was observed as the HRT was reduced to below 13 h. The reduction in efficiency could be due to less contact time between microbes and substrate.

**Figure 6.6 Effect of OLR on COD Removal Efficiency**

**Figure 6.7 Effect of HRT on COD Removal Efficiency**
6.2.3 Profile of Volatile Fatty Acids

VFA in influent concentration of the TE reactor was in the range of 740-2400 mg/L and the profile of VFA is shown in Figure 6.8. It was observed that the VFA in effluent were in the range of 260 – 650 mg/L.

![Figure 6.8 VFA Profile in Influent and Effluent](image)

6.2.4 Profile of Sulphate and Sulphide

The influent and effluent concentration of sulphate and sulphide in the UASB reactor treating TE was monitored and the profile of sulphate and sulphide concentrations are shown in Figure 6.9 and 6.10 respectively. It was observed that the sulphate concentration of the effluent reduced from 1600-2300 mg/L to 200-400 mg/L indicating more than 80% sulphate removal. Sulphate reduction is also one of the indicators of efficient anaerobic digestion. During anaerobic degradation, sulphate is reduced to sulphide which is confirmed from the Figure 6.10 indicating increase in concentration of sulphide from 200 - 350 to 550 - 800 mg/L. The growth of Sulphate Reducing Bacteria (SRB) dependent on the carbon and sulphate concentrations, whereas the growth of Methanogenic Bacteria (MB) is solely
dependent on the concentration of carbon (acetate). COD/SO$_4$ ratio ranging from 1 to 3 during the study period is shown in Figure 6.11. COD/SO$_4$ ratio below 3 indicates sulphate reducing bacteria (SRB) were more competitive when compared to the methane bacteria (Visser 1995).

![Figure 6.9 Sulphate Profile in Influent and Effluent](image-url)

![Figure 6.10 Sulphide Profile in Influent and Effluent](image-url)
6.2.5 Methane Production

The methane production was measured during the study period of the UASB reactor treating TE and the results are shown in Figures 6.12 and 6.13. It was observed that the daily methane production increased from 2 to 11 L/day. With the increase in OLR upto 10 kg/m$^3$.day the gas production also increased indicating consistent conversion of organic matter into methane. Maximum methane production was observed to be 0.27 m$^3$/kg of COD removed. With further increase in OLR to 12 kg/m$^3$.day, the methane production was reduced to 0.22 m$^3$/kg of COD removed and a drop in COD removal efficiency was also observed. This is mainly due to the washout of sludge at high rate of methane production i.e. methane production is more than 2 times the reactor volume. Hence it could be concluded that OLR should be in the range of 5-7 kg/m$^3$.day.
6.3 BIOMETHANIZATION STUDIES ON TE AND LLF (BENCH SCALE)

Bench scale UASB reactor was fed with TE and LLF for the biomethanization studies and the results are presented here under.
6.3.1 Feed Characteristics

The physico-chemical characteristics of UASB feed after mixing TE and LLF are given in Table 6.1.

<table>
<thead>
<tr>
<th>S.No</th>
<th>Parameter</th>
<th>Min</th>
<th>Max</th>
<th>Mean value</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>pH</td>
<td>7.2</td>
<td>8.2</td>
<td>7.4</td>
</tr>
<tr>
<td>2</td>
<td>BOD 5 days at 20°C (Total) (mg/L)</td>
<td>1120</td>
<td>5656</td>
<td>4300</td>
</tr>
<tr>
<td>3</td>
<td>COD (Total) (mg/L)</td>
<td>3500</td>
<td>14364</td>
<td>10980</td>
</tr>
<tr>
<td>4</td>
<td>Total Solids (TS) (mg/L)</td>
<td>13200</td>
<td>22515</td>
<td>15200</td>
</tr>
<tr>
<td>5</td>
<td>Total Dissolved Solids (TDS) (mg/L)</td>
<td>10650</td>
<td>18245</td>
<td>14825</td>
</tr>
<tr>
<td>6</td>
<td>Suspended Solids (SS) (mg/L)</td>
<td>2550</td>
<td>4270</td>
<td>3379</td>
</tr>
<tr>
<td>7</td>
<td>Sulphate as SO₄ (mg/L)</td>
<td>1556</td>
<td>2294</td>
<td>1999</td>
</tr>
<tr>
<td>8</td>
<td>Sulphide (as S) (mg/L)</td>
<td>164</td>
<td>349</td>
<td>273</td>
</tr>
<tr>
<td>9</td>
<td>TKN (mg/L)</td>
<td>240</td>
<td>450</td>
<td>197</td>
</tr>
<tr>
<td>10</td>
<td>C/N ratio</td>
<td>7.6</td>
<td>8.9</td>
<td>8.1</td>
</tr>
</tbody>
</table>

The pH of the UASB feed was between 7.2 to 8.2 with BOD and COD concentrations were in the range of 1120-5656 mg/L and 3500-14364 mg/L respectively during the study period. Total solids, sulphate and sulphide concentrations were in the range of 13200 - 22515 mg/L, 1556-2294 mg/L and 164-349 mg/L respectively. To increase the OLR to reactor, LLF portion was increased gradually keeping the TE volume constant. C/N ratio of the LF was in the range of 3.6-4.2 which is very low for anaerobic digestion, but after mixing LLF with TE the C/N ratio showed an increase and was in the range of 7.6-8.9. The average C/N ratio was 8.1, which is suitable for anaerobic biodegradation. It is reported that during anaerobic digestion, microorganisms utilize carbon 25-30 times faster than nitrogen. Thus to meet this requirement, microbes need a 20-30:1 ratio of C to N with the largest percentage of the carbon being readily degradable (Bardiya and Gaur, 1997).
6.3.2 Startup Regime of UASB Reactor Treating TE and LLF

6.3.2.1 Feed Details

Bench scale UASB reactor was fed with a mixture of 35 g of LF after liquefaction and 5L of TE. The reactor was fed with OLR of 1 kg/m$^3$.day, the OLR was increased gradually by increasing the COD of influent concentration by addition of LLF. Organic load applied in terms of COD and COD removal efficiency during the start up period are shown in Figures 6.14 and 6.15 respectively.

![Figure 6.14 OLR during the Startup](image)

Within 50 days, the loading rate was increased to 4 kg/m$^3$.day. A similar startup was recommended by Bileen and Gideon (2002). During the initial period, COD removal efficiency was only about 40%. As the bacteria got acclimatized, the COD removal efficiency increased to 65% within 50 days. During the startup period, biogas production also increased upto 4L within a period of 50 days indicating methanogenic activity. Influent and effluent pH observed during the startup period is shown in Figure 6.16. It was observed
that the pH of influent and effluent was in the range of 7.1 to 7.8 indicating good buffering capacity of the reactor.

Figure 6.15 COD Removal Efficiency during the Startup

Figure 6.16 pH Profile during Startup
6.3.2.2 VFA during Startup

VFA concentration in the influent was in the range of 1500-6000 mg/L. The VFA concentration in the effluent was below 1500 mg/L during the startup period. Under a steady state condition, hydrogen and acetic acid are formed due to acidogenic and acetogenic activities and are utilized immediately by the methanogens. Under over load conditions, the activity of methnaogenic and acetogenic population is reduced causing an accumulation of VFA which reduced the pH in the reactor. Increase in VFA concentration in the effluent indicates less methanogenic activity. From the Figure 6.17, it is seen that VFA concentration in the reactor was within 1500 mg/L during the startup period indicating active methanogenic activity. During the initial 20 days, bicarbonate alkalinity remained low (800mg/L) but later started building up as the reactor got stabilized as shown in Figure 6.18. Alkalinity concentration reached around 1400 – 1600 mg/L which helped in attaining better balance among the acid producers and acid utilizers resulting in higher COD and sulphate reductions, low VFA accumulation and higher biogas production.

![Figure 6.17 VFA Profile during the Startup](image)
6.3.2.3 Methane Production during Startup

From the Figure 6.19, it is seen that whenever COD load was increased, gas production also increased simultaneously, indicating consistent methanogenic activity. This was also used as one of the parameters to increase the organic load to reactor. Methane production of 0.240 m³/kg COD removed was observed at the end of 50 days, which indicates that start up was successfully and the OLR can be increased further without affecting the performance of the reactor.
6.3.3 Effect of OLR and HRT

After startup period, the reactor was operated at stabilized conditions at OLRs 4.5, 8.5 and 12 kg/m$^3$.day. At OLR of 4.5 kg/m$^3$.day the COD concentration of influent was 9600 mg/L after mixing with LLF of 341 mL and TE of 2246 mL. During this period, COD removal efficiency and HRT were in the range of 65-75% and 50-60 h respectively and methane production was 0.25-0.27 m$^3$ / kg of COD removed. From the results, it is seen that there was no adverse effect on the performance of the reactor. Therefore organic load was further increased to 8.5 kg/m$^3$.day and COD concentration of the influent was 11000 mg/L (LLF - 631 mL and TE - 3570 mL), and HRT during this period was 25-30 h. COD removal efficiency was found to be 75% and methane production of 0.29 m$^3$/kg of COD removed. OLR was increased further to 12 kg/m$^3$.day in the influent, COD concentration was 11600 mg/L (LLF - 874 mL and TE - 5040 mL), HRT during this period was 21 h, COD removal efficiency was 75% and methane production was consistently at the rate of 0.29 m$^3$/kg of COD removed. It was observed that consistent efficiency at higher loading rate was due to the biological pretreatment of LF. Particle size of the LLF was in the range of 2 µm to 200 µm, out of which about 70 % of the particles were below 35µm and the liquefaction in terms of soluble organic (COD$_s$) of LF was 93.3%, when the OLR rate was increased from 4.5 to 12 kg/m$^3$.day, it could be seen from the Figures 6.20 to 6.23 that increase in OLR neither affected the COD removal efficiency nor methane production. COD removal was found to be 72-75% and corresponding methane production was in the range of 0.27-0.29 m$^3$/kg of COD removed and the above results indicate that there is no significant effect due to change in OLR.
Figure 6.20 Effect of OLR on COD Removal (4.5 kg/m$^3$.day)

Figure 6.21 Effect of OLR on COD Removal (8.5 kg/m$^3$.day)
COD removal efficiency was found to be in the range of 70-75% when the HRT was in the range of 30-20 h as shown in Figure 6.24. It is seen from the results that, the COD removal efficiency was consistent and the
reactor was stable with reduction in HRT. However, it is also observed that when OLR was increased to 14 kg/m$^3$.day, COD removal efficiency reduced to 65% during this period sludge wash out was observed in the effluent. Methane production per kg of COD removed also reduced from 0.29 to 0.23 m$^3$. Hence it was concluded that OLR more than 12 kg/m$^3$.day cannot be fed into the reactor. It was also reported by Torkian et al 2002 that treatment of slaughter house wastewater, when OLR increased to 20 kg/m$^3$.day loading resulted in operational problems including sludge wash out and loss of microbial population.

![Figure 6.24 Effect of HRT on COD Removal Efficiency](image)

**6.3.4 Methane Production**

The methane production measured during the study period of TE and LLF reactor is shown in Figures 6.25 to 6.30. From the graphs, it was observed that the daily methane production increased up to 14 L/day. With increase in the COD loading rate, the gas production also increased indicating consistent conversion of organic matter into methane. At OLR of 4.5
kg/m$^3$.day, it is seen from the results that methane production was 0.25-0.27 m$^3$/kg of COD removed and the COD removal efficiency was around 70%.

At OLR 8.5 kg/m$^3$.day under steady state condition, it was observed that the methane yield was 0.27-0.29 m$^3$/kg of COD removed (8-10 L/day) and the COD removal efficiency was 75%. At the loading rate of 12 kg/m$^3$.day methane yield was found to be 0.29 m$^3$/kg of COD removed. OLR of more than 12 kg/m$^3$.day, solids in the reactor were washed out and scum formation was observed in the GLS separator and hence a drop in COD removal efficiency has reduced to 65%. In the UASB reactor treating TE with LLF and TE only, methane yield was 0.29 m$^3$ and 0.27 m$^3$ per kg of COD removed respectively. From the results it could be concluded that even after increasing the OLR from 4.5 – 12 kg/m$^3$.day, an increase of about 250%, the COD removal efficiency was 75 % and methane yield was 0.29 m$^3$/kg of COD removed, it indicates that the reactor performance of consistent up to a OLR of 12 kg/m$^3$.day.

![Figure 6.25 Effect of OLR on Methane Production (4.5 kg/m$^3$.day)](image)

Figure 6.25 Effect of OLR on Methane Production (4.5 kg/m$^3$.day)
Figure 6.26  Effect of OLR on Methane Production (8.5 kg/m$^3$.day)

Figure 6.27  Effect of OLR on Methane Production (12 kg/m$^3$.day)
Figure 6.28 Effect of OLR on Methane Production per kg of COD Removed (4.5 kg/m$^3$.day)

Figure 6.29 Effect of OLR on Methane Production per kg of COD Removed (8.5 kg/m$^3$.day)
Figure 6.30 Effect of OLR on Methane Production per kg of COD Removed (12 kg/m$^3$.day)

6.3.5 Profile of Volatile Fatty Acids

VFA concentration in the influent of the LLF with TE reactor was in the range of 5000-6800 mg/L and VFA concentration in the effluent was in the range of 700-1350 mg/L, as shown in Figure 6.31. From Figure 6.31, considerable VFA reduction in outlet was observed, indicating the effective functioning of methanogenic bacteria (MB). During steady state condition, effluent VFA was in the range of 1000-1100 mg/L. Low concentrations of VFA in the reactor system shows effective degradation of the substrate. Non accumulation of VFA in the reactor also indicates an active MB population in the sludge resulting in efficient conversion of VFA to methane. In the reactor treating TE and LLF, no accumulation of VFA was observed indicating the presence of active bacterial population.
From the Figure 6.32, alkalinity in the reactor was monitored to understand the condition of buffering activity of reactor, while influent concentration of alkalinity ranged between 1000-1600 mg/L and effluent alkalinity ranged from 2200-3900 mg/L. The increase in alkalinity in the effluent is due to the solubilization of part of CO\textsubscript{2} during the generation of biogas. This increases the bicarbonate alkalinity which helps in maintaining the pH of the reactor in neutral range.

VFA/alkalinity ratio of TE and LLF reactor was in the range of 0.3-0.4 (Figure 6.33) indicating that the reactor is in stable condition. When the VFA/Alkalinity ratio were <0.4, the reactor is considered to be stable (Callaghan et al 2000).
Figure 6.32 Bicarbonate Alkalinity Profile in Influent and Effluent

Figure 6.33 Profile of VFA/Alkalinity Ratio
6.3.6 Profile of Sulphate and Sulphide

The influent and effluent concentration of sulphate and sulphide are shown in Figures 6.34 – 6.35. It was observed that the sulphate concentration of the effluent reduced from 1300-2200 mg/L to less than 250-400 mg/L indicating more than 80% sulphate removal. During anaerobic degradation, sulphates are reduced to sulphide which is confirmed from the Figure 6.35 indicating an increase in concentration of sulphide from 250-300 mg/L to 650-750 mg/L. The growth of sulphate reducing bacteria (SRB) was dependent on the carbon and sulphate concentration, whereas the growth of MB was solely dependent on the concentration of carbon (acetate). At low sulphate concentration, growth of SRB would be sulphate limited and enables MB to out compete SRB. Although thermodynamic and kinetic considerations favour sulphate reduction over methanogenesis, it was often observed that MB was able to effectively out compete SRB for acetate conversion (Isa et al 1986; Yoda et al 1987). In the TE and LLF reactor, COD to $SO_4^{2-}$ ratio of around 6 was observed (Figure 6.36) during steady state condition and this favours the anaerobic fermentation by the effective performance of the reactor along with the increase in biogas yield. It was reported that at high COD/$SO_4^{2-}$ ratios (>6), MB predominated while at lower COD/$SO_4^{2-}$ ratios (<1.5), SRB were more competitive. The increase in COD/$SO_4^{2-}$ ratio favored the anaerobic fermentation by the effective performance of the reactor (COD removal) along with increase in biogas yield (Venkatamohan et al 2005). Methane gas production is directly related to organic matter removal by anaerobic bacteria. Therefore, its drop with increasing sulphate level implies the decrease in methanogenic activity as both SRBs and MB compete for the same organic sources.
Figure 6.34 Sulphate Profile in Influent and Effluent

Figure 6.35 Sulphide Profile in Influent and Effluent
6.3.7 Discussion on COD Balance

From the bench scale studies conducted for about 10 months, \( \text{COD}_e \), \( \text{COD}_{\text{CH}_4} \), \( \text{COD}_{\text{new cell}} \), \( \text{COD}_{\text{SO}_4,R} \) and dissolved methane in the effluent are given in Appendix 6. Theoretically, volume of methane that can be produced at 30°C per g of COD removed is 0.388 L. However, only a part of the biogas formed in UASB will be available for energy purposes, with the rest staying dissolved in the effluent and passing out in the effluent. Some portion of COD is converted into cell tissue. For sulphate reduction also some amount of COD is consumed. At steady state condition, COD mass balance could be shown that the COD removed is oxidized or accounted for cell growth or synthesis.
\[ \text{COD}_r = \text{COD}_{\text{syn}} + \text{COD}_o \]  \hspace{1cm} (6.1)

Where in

\[ \text{COD}_r \] = COD utilized (gCOD/day)

\[ \text{COD}_{\text{syn}} \] = COD incorporated into cell synthesis (g COD/day)

\[ \text{COD}_o \] = COD oxidized (g COD/day)

For cell synthesis, COD$_{\text{syn}}$ is calculated from the net biomass yield and the ratio of 1.42 g O$_2$/g VSS. The oxygen equivalent of the biomass is equal to the COD incorporated into biomass.

\[ \text{COD}_{\text{syn}} = 1.42 \ Y_n \ \text{COD}_r \]

Where \( Y_n \) = net biomass yield g VSS/g COD$_r$

\[ \text{COD}_i = \text{COD}_e + \text{COD}_{\text{CH}_4} + \text{COD}_{\text{SO}_4} + \text{COD}_{\text{new cells}} \]  \hspace{1cm} (6.2)

Where in

\[ \text{COD}_i \] = COD in the influent

\[ \text{COD}_e \] = COD in the effluent

\[ \text{COD}_{\text{CH}_4} \] = COD in the methane (1 m$^3$ CH$_4$ = 2.57 kg COD)

\[ \text{COD}_{\text{SO}_4} \] = COD consumed for sulphate reduction (1g COD per 1.5 g sulphate reduced)

\[ \text{COD}_{\text{new cells}} \] = COD consumed for new cells formation (1 kg of VS = 1.42 kg of COD)

It was observed that only 53 % of methane was recovered i.e 0.29 m$^3$/kg of COD removed, as against the theoretical value of 0.388 m$^3$/kg of COD removed in treating TE and LLF. The typical COD mass balance of bench scale studies is given in Table 6.2.
Table 6.2 COD Mass Balance of Bench Scale Studies

<table>
<thead>
<tr>
<th>COD removal path</th>
<th>Percentage of COD</th>
<th>Basis</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas recovery COD(_{CH_4})</td>
<td>53%</td>
<td>Measured methane gas production (1 m(^3) of CH(_4) = 2.57 kg COD)</td>
</tr>
<tr>
<td>Dissolved methane</td>
<td>1%</td>
<td>0.028 m(^3) methane/m(^3) of effluent</td>
</tr>
<tr>
<td>COD remaining in effluent (COD(_{e}))</td>
<td>25%</td>
<td>COD measured in UASB effluent</td>
</tr>
<tr>
<td>COD consumed in sulphate reduction COD(_{SO_4}R)</td>
<td>10%</td>
<td>1g of COD per 1.5 g of sulphate reduced</td>
</tr>
<tr>
<td>COD removed as new cell formation (COD(_{new\ cells}))</td>
<td>11%</td>
<td>Measured VS in sludge over averaging period (1 kg VS = 1.4 kg COD)</td>
</tr>
<tr>
<td>Total</td>
<td>100%</td>
<td></td>
</tr>
</tbody>
</table>

### 6.3.8 Discussion

A control reactor treating only TE and the experimental reactor treating TE with LLF was operated and the performance of the reactors was compared. In control reactor OLR was increased by increasing the flow of TE and in experimental reactor OLR was increased by increasing the LLF. Both reactors were evaluated for COD removal efficiency, methane production and VFA reduction. Both the reactors were seeded with anaerobic sludge (suspended) from the UASB reactor. During the startup period of nearly 90 days OLR was gradually increased in both reactors. Gavala et al (1999) reported that the OLR of 6.5 g COD/L.day was found to be safe and could be increased to a maximum of 7.5 g COD/L.day. At OLR of 4.5 kg/m\(^3\).day the experimental reactor was operated at steady state conditions and COD removal efficiency and HRT were in the range of 65-75% and 50-60 h respectively. At this operating condition methane production was 0.25-0.27 m\(^3\)/kg of COD removed. Organic load was further increased to 8.5 kg/m\(^3\).day. HRT during this period was 25-30 h. COD removal efficiency was
found to be 75% and methane production of 0.29 m$^3$/kg of COD removed. OLR was increased further to 12 kg/m$^3$.day in the influent. HRT during this period was 21 h, COD removal efficiency was 75% and methane production was consistent at the rate of 0.29 m$^3$/kg of COD removed. In the control reactor, with the increase in OLR upto 10 kg/m$^3$.day the gas production also increased indicating consistent conversion of organic matter into methane and maximum methane production was observed to be 0.27 m$^3$/kg of COD removed with further increase in OLR to 12 kg/m$^3$.day, the methane production was reduced to 0.22 m$^3$/kg of COD removed and drop in COD removal efficiency was also obtained. Uemura and Harada (2000) reported that methane production of 0.26 per kg of COD removed from treatment of sewage in UASB reactor.

Higher methane yield was obtained in the experimental reactor compared to the control reactor wherein pH of the influent was in the range of 8.2-8.9. In experimental reactor influent pH was reduced to a range of 7.1 to 7.8 due to the addition of LLF with TE which is more suitable pH for biomethanization in anaerobic reactor. In the TE and LLF reactor, COD/SO$_4^{2-}$ ratio was around 6 during steady state condition and this favours the anaerobic fermentation by the effective performance of the reactor along with the increase in biogas yield. It was reported that at high COD/SO$_4^{2-}$ ratios (>6), MB predominated while at lower COD/SO$_4^{2-}$ ratios (<1.5), SRB were more competitive. The increase in COD/SO$_4^{2-}$ ratio favored the anaerobic fermentation by the effective performance of the reactor (COD removal) along with increase in biogas yield (Venkatamohan et al 2005). In control reactor, COD/SO$_4$ ratio was ranging from 1 to 3 during the study period. COD/SO$_4$ ratio below 3 indicates sulphate reducing bacteria (SRB) were more competitive when compared to the methane bacteria (Visser 1995).

Higher methane yield of 37.5% could be obtained by treating LLF along with TE in UASB reactor. Considering the liquefaction period of 8 days
and one day HRT in UASB reactor, the LF could be treated within 8 days and energy could be recovered. The higher methane yield per kg of COD removed is due to the dilution of sulphide concentration in the reactor. In addition, this buffering capacity of the reactor was enhanced due to the presence of lime (Vlyssides and Karlis 2004). Hence, this approach not only improves energy recovery but also eliminates the disposal problems of LF and dispenses away with the necessity of a separate anaerobic digester for biomethanization of LF.

From the bench scale studies, the kinetic constants were arrived at using Monod equations (Ghangrekar 2006) under steady state conditions. Kinetic constants, half velocity constant ($K_s$), rate of substrate utilization ($k$), yield coefficient ($Y$), decay coefficient per day ($k_d$), specific utilization rate ($U$) were arrived at and these constants were used to design the pilot scale reactor.

### 6.4 BIOMETHANIZATION OF LF AND TE (MODIFIED BENCH SCALE)

To study the effect of liquefaction on recirculation of tannery effluent through LF was carried out using an online liquefaction reactor. The effect of recirculation and methane generation was studied.

#### 6.4.1 Effect of Recirculation on Liquefaction and Biomethanization of LF

Average COD concentration of influent and effluent of the UASB reactor were found to be 1767 mg/L and 1296 mg/L respectively and that of liquefaction reactor (LR) was 3005 mg/L. In LR, COD concentration decreased with the increase in recirculation rate from 3.2 L/day (3783 mg/L) to 6.5 L/day (2314 mg/L), 10.6 L/day (2146 mg/L) and 11 L/day (1767 mg/L) due to dilution and immediate conversion of COD to methane. Average VFA concentration in influent (TE), effluent and LR’s average VFA were found to
be 749 mg/L, 336 mg/L and 1508 mg/L respectively (Figure 6.37). From the results of methane generation, it is observed that liquefaction of LF is taking place in LR effectively.

Initially, VFA in LR was more than 2226 mg/L when recirculation rate was 3 L/day and decreased to 846 mg/L when recirculation rate was increased to 11 L/day. VFA was converted into methane in LR when recirculation rate increased from 6.5 L/day to 11 L/day. Recirculation of UASB reactor effluent into LR and production of VFA in LR resulted in reduction of pH. Initially, pH value in LR was > 9 and decreased to 7.4 at the end of study.

Recirculation rate was increased from 3 L/day to 3.2 L/day, 6.5 L/day, 10.6 L/day and 11 L/day to assess the effect of recirculation rate on methane production. Corresponding average methane production with respect to above mentioned recirculation rate in LR was found to be 0.12L, 0.59L, 1.7L, 2.6L and 3.3L. Thus methane production in LR was increasing with increase in recirculation rate due to more microbes (fermentative, acidogenic

![Figure 6.37 Profile of VFA in UASB Reactor Influent, Effluent and LR](image-url)
and acetogenic) present in LR. This is reflected in the consumption of more VFA in LR which was converted into methane in LR itself. On the other hand, in case of UASB reactor, corresponding average methane production was found to be 2.2L, 1.94L, 1.46L, 1.12L and 0.9L. It was observed that with increase in recirculation rate, methane production decreased. VFA, which was responsible for methane production, transferred to UASB reactor in the form of LLF was decreased through recirculation in UASB reactor and hence decrease in methane production was observed. Cumulative methane production (Figure 6.38) in LR was higher than that in UASB reactor. Methanogens were promulgated for UASB to LR through recirculation from UASB reactor. With increase in recirculation rate from 6.5 L/day to 11 L/day, methane production in LR was increased. Totally, 128L and 145L of methane were produced in UASB reactor and LR respectively.

![Cumulative Methane Production in UASB reactor and LR](image)

**Figure 6.38 Cumulative Methane Production in UASB reactor and LR**

6.5 MATHEMATICAL MODELING AND KINETICS OF BIOMETHANIZATION STUDIES

Mathematical modeling of biomethanization was developed by arriving at the kinetics of biomethanization. Influent, effluent, MLVSS
concentration in the reactor and flow data collected by operating bench scale reactor is used to arrive at kinetic constants and to predict the effluent soluble substrate concentration, reactor biomass and volume of the reactor. To evaluate the bio kinetic constants for LLF with TE and TE reactors, varying HRT, OLR and $\theta c$ were considered. The following equations were used to arrive at the kinetic constants.

$$\frac{1}{\theta_c} = \frac{Y k S}{K_s + S} - k_d$$  \hspace{1cm} (6.3)

$$\frac{1}{\theta c} = Y U - k_d$$  \hspace{1cm} (6.4)

$$\frac{1}{U} = \left( \frac{K_s}{k} \right) \left( \frac{1}{S} \right) + \left( \frac{1}{k} \right)$$  \hspace{1cm} (6.5)

Where

$$U = \frac{S_0 - S}{\theta X}$$  \hspace{1cm} (6.6)

where  
$K_s$ - half velocity constant (mg/L)  
$Y$ - yield coefficient, (mgVSS/mg COD)  
$k$ - rate of substrate utilization (day$^{-1}$)  
$k_d$ - decay coefficient (day$^{-1}$)  
$\theta c$ - mean cell residence time (day)  
$U$ - specific utilization rate (mg COD applied / mg MLVSS/day)

Kinetic constants for both the reactors were arrived at, when reactors were operated at steady state conditions. The method of least squares was used to obtain the line of best fit. Considering the Monod equation
\(X_0c/(S_0-S) = (K_s/k)(1/S) + 1/k\), which is reciprocal of F/M ratio was plotted against effluent substrate concentration (1/S) and the straight line fit was obtained and shown in Figures 6.39 and 6.40 on COD basis. From intercept of the plot and slope of the plot, \(k\) and \(K_s\) were obtained.

Maximum growth rates obtained for treatment of TE and for combined treatment of TE with LLF were 0.288 and 0.184 g. new cell/g. cell d\(^{-1}\) respectively. Maximum specific substrate utilization rate was lower for reactor treating TE with LLF when compared to reactor treating TE. However yield coefficient was 0.190 mg VSS/g COD for LLF with TE and 0.179 mg VSS/mg COD for TE. Decay coefficient was marginally higher in LLF with TE when compared to TE. Comparison of relative kinetic constants for LLF with TE and TE alone are given in Table 6.3.
\[ y = 0.179 \text{mgVSS/mg COD} \]

\[ k_d = 0.006 \text{d}^{-1} \]

\[ R^2 = 0.947 \]

(b)  

Figure 6.39  Kinetics of Treatment (TE)

\[ y = 6615 \times + 1.027 \]

\[ R^2 = 0.908 \]

(a)
The organic loading rate of the reactor was increased in steps by reducing HRT or increasing influent COD concentration to assess the performance at different loading rates and to study the effect of SRT and COD removal efficiency. The operating condition at different HRT of (1 to 2 days) and performance of the reactor are furnished in Figure 6.24. The organic loading rates were between 4.5-12 kg/m$^3$.day. pH of the reactor...
influent was in the range of 7.5 to 8. Biokinetic parameters evaluated from the present study were used in designing of the UASB reactor to treat TE and LLF.

6.5.1 Specific Methane Production

The specific methane production rate for varying specific substrate utilization rate in reactors treating TE and combined treatment of TE and LLF are shown in Figures 6.41 and 6.42 respectively. From the Figure 6.41, it is seen that at 0.35 kg CH₄-COD/kg VSS specific methane production, specific substrate utilization rate was 0.48 kg COD/kg VSS. The slope indicates that about 70% of the COD was converted to methane and remaining COD was utilized by SRBs and for formation of new cells. Biomass of the reactor was measured and COD/VSS ratio was found to be 0.6. From the results it is observed that substrate utilization rate was higher in the reactor treating LLF with TE when compared to reactor treating TE due to higher loading rate in the reactor treating LLF with TE and it may also be concluded that the reactor can be operated upto OLR of 10 kg/m³.day without affecting the specific methane production.

![Graph showing specific methane production vs. specific substrate utilization rate](image)

Figure 6.41 Specific Methane Production (TE)
The specific methane production against the COD loading rate of the reactors treating TE alone and combined treatment of TE and LLF are shown in Figure 6.43. The specific methane production increased with the increase in OLR up to 12 kg/m$^3$.day. Beyond this OLR, the specific methane production came down due to washout of sludge and leakage in GLS separator was also observed which was due to clogging of the GLS separator.
6.6 BIOMETHANIZATION STUDIES ON TE AND LLF (PILOT SCALE)

Based on the outcome of the results of bench scale studies a pilot scale plant for treatment of capacity 10 m$^3$/day was designed for treatment of TE and LLF using kinetic constants obtained from bench scale studies. Biokinetic constants applied for design of pilot scale plant were half velocity constant (Ks) of 6441 mg/L, rate of substrate utilization (k) of 0.97 day$^{-1}$, yield coefficient (Y) of 0.190 mgVSS/mg COD, decay coefficient (k_d) of 0.007 day$^{-1}$, maximum specific growth rate ($\mu_{max}$) of 0.184 g.new cell/g.cells d$^{-1}$, specific utilization rate (U) of 0.34 day$^{-1}$ using equation 3.10. The capacity of the pilot scale was arrived at as 12.5 m$^3$. TE and LLF were fed in to the UASB reactor continuously to evaluate the treatment performance.

6.6.1 Performance of Pilot Scale Reactor

Pilot scale UASB reactor was working at steady state conditions with OLR of 5.8-6.8 kg/m$^3$.day with TE. In the same reactor, LLF was also introduced. COD removal efficiency was in the range of 70-75% before and after introducing LLF. LLF did not affect the performance of the reactor in terms of COD removal efficiency. HRT was maintained at about 24h during the study period. The OLR and removal efficiency observed are depicted in Figure 6.44. To maintain upward velocity of 0.5 m/h the supernatant from the UASB reactor was re-circulated by mixing with influent.
6.6.2 Methane Production

Methane yield measured during the study period is shown in Figures 6.45 and 6.46. Methane production of 13 to 17 m$^3$/day was observed from the reactor. After introduction of LLF into the operating UASB reactor treating TE, no reduction in COD removal efficiency was observed, indicating the microbes in the reactor got adapted to LLF. During this period, the total VS in the reactor was estimated to be 150 kg. The composition of biogas was estimated and given in Table 6.4.

<table>
<thead>
<tr>
<th>S.No</th>
<th>Biogas composition</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Carbon dioxide (%)</td>
<td>25-30</td>
</tr>
<tr>
<td>2</td>
<td>Methane (%)</td>
<td>70-75</td>
</tr>
<tr>
<td>3</td>
<td>Hydrogen sulphide (%)</td>
<td>1.5-2.0</td>
</tr>
</tbody>
</table>
From pilot scale studies of TE and LLF it was observed that methane yield of 0.29 m$^3$/kg of COD could be obtained with HRT of 24-26 h and COD removal efficiency of 75% and results from bench scale studies were thus validated. Profiles of VFA influent and effluent are shown in Figure 6.47 and from the results it is observed that VFA concentration of the influent was in the range of 1080-2400 mg/L and the effluent was in the range of 198-547 mg/L, indicating that the conversion of VFA to methane was effective and introduction of LLF did not affect the performance of the reactor in terms of conversion efficiency.

Figure 6.45 Effect of OLR on Methane Production
Figure 6.46  COD Removal Efficiency and Methane Production per kg of COD Removed

Figure 6.47 VFA Profile of Influent and Effluent
6.6.3 Profile of Volatile and Suspended Solids

Suspended solids (SS) inside the reactor were estimated from the sludge profile data. VS and SS in the reactor during the study period are shown in Figure 6.48. From the results, it is observed that SS in the reactor was in the range of 350-430 kg and volatile solids were in the range of 150-170 kg. There was no appreciable change in the quantity of VS in the reactor, indicating that the introduction of LLF has not affected the performance of the reactor.

From the above studies, it could be concluded that LLF can be treated by combining with TE and energy can be recovered and no separate anaerobic digester for biomethanization of LF is required.

![Figure 6.48 VS and SS in the Pilot Scale UASB Reactor](image)
6.6.4 Discussion

LF of 200 kg was liquefied by adding inoculum in the ratio of 1.5% of VS and liquefied LF was mixed with 30 m$^3$ of TE in the conditioning tank. This mixture was fed into the UASB reactor at OLR of around 5.8 – 6.8 kg/m$^3$.day and HRT was maintained at about 24h during the study period. COD removal efficiency in the range of 70-75% and methane production of 0.29 m$^3$/kg of COD removed were observed. Biogas composition was analysed and methane was in the range of 70-75 %, pilot scale reactor was operated under atmospheric temperature. Average VS in the reactor was found to be about 150 kg. The results obtained in the pilot scale studies are similar to or better than results obtained in the bench scale studies.

6.7 DESIGN METHODOLOGY FOR FULL SCALE PLANT

Based on the outcome of the results of bench scale and pilot scale studies, a full scale plant for treatment capacity of 5090 m$^3$/day was designed for treatment of TE and LLF using kinetic constants obtained from bench scale studies. Biokinetic constants, half velocity constant (Ks) of 6441 mg/L, rate of substrate utilization (k) of 0.97 day$^{-1}$, yield coefficient (Y) of 0.190 mgVSS/mg COD, decay coefficient (k$_d$) of 0.007 day$^{-1}$, maximum specific growth rate ($\mu_{\text{max}}$) of 0.184 g.new cell/g.cells d$^{-1}$, specific utilization rate (U) of 0.34 day$^{-1}$ were applied in the equation 6.7 for design of the full scale plant.

$$VX = \left[ \frac{YQ(S_0 - S)}{1 + (k_d)\theta_c} \right]$$

(6.7)

The full scale plant was designed for solid retention time ($\theta_c$) of 25 days, HRT of 24 h, MLVSS concentration of 20 kg/m$^3$ and COD removal efficiency of 75 %. The volume of the reactor was arrived at as 5280 m$^3$ and used in the design.
6.8 CLEAN DEVELOPMENT MECHANISM (CDM)

In this study, Certified Emission Reduction (CER) obtained by replacement of anaerobic lagoon by UASB for treatment of TE and replacement of open dumping of fleshings by liquefaction and treatment of LLF in UASB were estimated as per the guidelines of UNFCCC for a tannery cluster.

6.8.1 Certified Emission Reduction from Treatment of TE Alone (Anaerobic Lagoon Vs UASB)

Baseline emissions were estimated based on the equation 4.1 given in chapter 4, vide section 4.1. In the anaerobic lagoon treating TE alone (option I), the breakup of CO₂ equivalent emission are shown in Figure 6.49. It was observed from the Figure 6.49 that methane emission from the lagoon was contributing to an extent of 76% i.e., 8088.93 tCO₂ e/yr and remaining 24% was due to CO₂ emission reduction due to the electricity generation (2557.65 tCO₂e/yr) which will be replaced by the biogas produced from the UASB reactor. The overall baseline emission has been estimated to be 10646.58 tCO₂e/yr and the details of calculations are given in Appendix 1.

Project Emissions are estimated based on the equation 4.4 given in chapter 4, vide section 4.1. Similarly, project emission due to the treating of TE in UASB (project activity) based on the outcome of the present study, has also been estimated. It was observed that methane emission from the lagoon was avoided by UASB except for methane leakage during collection which was estimated as 507.04 tCO₂e/yr. CER due to replacement of existing anaerobic lagoon by UASB for treatment of TE is found to be 10139.54 tCO₂e/yr.
6.8.2 Certified Emission Reduction from Treatment of LF (Open Dumping Vs UASB)

During disposal of LF by open dumping (baseline), the breakup of CO₂ emission are shown in Figure 6.50. It was observed from the Figure 6.50 that methane emission from open dumping was contributing to an extent of 97% i.e., 2464.66 tCO₂e/yr and the remaining 3% was due to CO₂ emission reduction due to the electricity generation (539.45 tCO₂e/yr) which will be replaced by the additional biogas produced from the UASB reactor due to the addition of LLF in the feed. The overall baseline emission has been estimated to be 3004.11tCO₂e/yr as per UNFCCC guidelines and the calculations are given in Appendix 2.

Project emission due to treatment of LLF in UASB has been estimated based on the outcome of the present study. It was observed that methane emission due to open dumping was avoided by implementation of UASB reactor except for methane leakage during collection which was estimated to be 106.94 tCO₂e/yr. CER due to replacement of open dumping of LF by treatment of LLF in UASB reactor is found to be 2897.17 tCO₂e/yr.
6.8.3 CER Generated from Combined Treatment of TE and LLF

CER generated due to replacement of anaerobic lagoon and open dumping by combined treatment of LLF and TE using UASB reactor is estimated and given in Table 6.5. CER generated from replacement of anaerobic lagoon by UASB for treatment of TE alone was 10646.58 tCO$_2$e/yr and CER generated by replacing open dumping by UASB for treatment of LF was 2897.17 tCO$_2$e/yr.

Table 6.5 CER for Combined Treatment of TE and LLF in UASB

<table>
<thead>
<tr>
<th>Alternatives</th>
<th>Baseline Emission (tCO$_2$e/yr)</th>
<th>Project Emission (tCO$_2$e/yr)</th>
<th>Emission Reduction (tCO$_2$e/yr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>CER from treatment of TE alone</td>
<td>10646.58</td>
<td>507.04</td>
<td>10139.54</td>
</tr>
<tr>
<td>(anaerobic lagoon Vs UASB)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CER from treatment of LF (Open dumping Vs UASB)</td>
<td>3004.11</td>
<td>106.94</td>
<td>2897.17</td>
</tr>
<tr>
<td>CER (tCO$_2$e/yr)</td>
<td></td>
<td></td>
<td>13036.71</td>
</tr>
</tbody>
</table>
CER generated due to implementation of combined treatment of LLF and TE using UASB is estimated to be 13036.71 tCO$_2$e/yr, which would be an added advantage in addition to pollution reduction and energy generation.

6.9 TECHNO-ECONOMIC ANALYSIS

Techno-economic analysis was carried out for a typical tannery cluster processing 150 tons of raw hides and skins per day. CETP with two process options were designed: (i) treating TE alone in open anaerobic lagoon for a design flow of 5000 m$^3$ per day; and (ii) combined treatment of TE along with LLF using UASB reactor for design flow of 5090 m$^3$ per day. Design criteria adopted and size of the treatment units arrived at, for above two processes and options are given in Appendix 3.

6.9.1 Cost Estimation

Abstract cost estimate for capital cost for treatment of TE (option I) and combined treatment of TE and LLF (option II) was arrived at for individual units and given in Table 6.6. Cost breakup details for civil, mechanical, electrical and instrumentation are given in Appendix 4. It is estimated that the capital cost for CETP for treatment of TE alone in anaerobic lagoon and combined treatment of TE and LLF in UASB reactor are `188,665,879 (188.66 million) and `195,990,470 (195.99 million) respectively. It was observed that the unit capital cost increased by about `7,324,591 due to additional cost towards UASB reactor and liquefaction tank required for pretreatment of LF and combined treatment of TE and LLF, which is about 3.8% higher when compared to cost of TE alone. The unit capital cost per MLD was estimated as `37.7 million and `39.2 million for treating TE alone in anaerobic lagoon and combined treatment TE and LLF in UASB reactor respectively.
Table 6.6 Abstract Cost Estimate for Capital cost for Treatment

<table>
<thead>
<tr>
<th>S. No.</th>
<th>Description</th>
<th>Treatment of TE alone in Anaerobic Lagoon (Option I)</th>
<th>Combined Treatment of TE and LLF in UASB (Option II)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Civil works</td>
<td>92,879,636</td>
<td>92599,533</td>
</tr>
<tr>
<td>2</td>
<td>Mechanical works</td>
<td>61,276,000</td>
<td>68,922,000</td>
</tr>
<tr>
<td>3</td>
<td>Piping</td>
<td>1,200,000</td>
<td>1,400,000</td>
</tr>
<tr>
<td>4</td>
<td>Electrical and Instrumentation</td>
<td>18372094</td>
<td>17264569</td>
</tr>
<tr>
<td>6</td>
<td>Laboratory equipment</td>
<td>1,000,000</td>
<td>1,000,000</td>
</tr>
<tr>
<td>7</td>
<td>CETP Trail Runs, Commissioning, Stabilization &amp; Standardization</td>
<td>4,500,000</td>
<td>5,000,000</td>
</tr>
<tr>
<td>8</td>
<td>Contingencies</td>
<td>9,438,149</td>
<td>9,804,368</td>
</tr>
</tbody>
</table>

Cost *188,665,879 195,990,470*

6.9.2 Operation and Maintenance Cost (O&M)

It is estimated that the O&M cost for CETP for treatment of TE alone in anaerobic lagoon and combined treatment of TE and LLF in UASB reactor were `1,012 and `897 respectively per ton of leather processed (Table 6.7). On volumetric basis the cost is estimated as `33.7 and `29.8 per m³ of wastewater for treatment of TE alone in anaerobic lagoon and combined treatment of TE and LLF using UASB reactor respectively. Details and basis for arriving at O&M cost are given in Appendix 5.
Table 6.7 Operation and Maintenance Cost for Treatment

<table>
<thead>
<tr>
<th>S.No</th>
<th>Description</th>
<th>Treatment of TE alone in anaerobic lagoon (Option I)</th>
<th>Combined treatment of TE and LLF in UASB (Option II)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Chemical cost</td>
<td>59,000</td>
<td>59,000</td>
</tr>
<tr>
<td>2</td>
<td>Manpower cost</td>
<td>13,067</td>
<td>13,067</td>
</tr>
<tr>
<td>3</td>
<td>Electrical power cost</td>
<td>48,276</td>
<td>38,704</td>
</tr>
<tr>
<td>4</td>
<td>Sludge handling cost</td>
<td>48,276</td>
<td>38,704</td>
</tr>
<tr>
<td></td>
<td><strong>O&amp;M cost per day</strong></td>
<td><strong>168,618</strong></td>
<td><strong>149,475</strong></td>
</tr>
<tr>
<td></td>
<td><strong>O&amp;M cost per m³</strong></td>
<td><strong>33.7</strong></td>
<td><strong>29.8</strong></td>
</tr>
<tr>
<td></td>
<td><strong>O&amp;M cost per ton of leather processed</strong></td>
<td><strong>1,012</strong></td>
<td><strong>897</strong></td>
</tr>
</tbody>
</table>

6.9.3 Annualized Cost

Annualized cost was arrived at using equation 4.6. Techno-economic analysis for treatment of TE alone using anaerobic lagoon (option I) and combined treatment of TE and LLF using UASB reactor (option II) has been investigated and the results are shown in Table 6.8. It was estimated that the total annualized unit cost for treatment works out to ` 82,907,855 (82.9 million) and ` 79,087,522 (79 million) for treatment of TE alone for a design flow of 5000 m³/day and combined treatment of TE and LLF for a design flow of 5090 m³/day respectively. The unit cost works out to ` 1842 and ` 1758 per ton of hides or skins processed for a typical tannery cluster processing 150 tons of raw hides and skins per day and generating about 5000 m³/day of TE and 30 tons/day of LF.
<table>
<thead>
<tr>
<th>S. No.</th>
<th>Description</th>
<th>Treatment of TE alone in anaerobic lagoon (Option I)</th>
<th>Combined treatment of TE and LLF in UASB (Option II)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Capital cost (₹)</td>
<td>188,665,879</td>
<td>195,990,470</td>
</tr>
<tr>
<td>2</td>
<td>Annualized Capital cost (₹)</td>
<td>32,327,855</td>
<td>33,582,922</td>
</tr>
<tr>
<td>3</td>
<td>Annual Operation&amp; Maintenance cost (33/29.8x5000x25×12) (₹)</td>
<td>50,580,000</td>
<td>45,504,600</td>
</tr>
<tr>
<td>4</td>
<td>Total Annualized capital and operational cost (₹)</td>
<td>82,907,855</td>
<td>79,087,522</td>
</tr>
<tr>
<td>5</td>
<td>Treatment cost for processing per ton of skins /hides processed (₹)</td>
<td>1,842</td>
<td>1,758</td>
</tr>
</tbody>
</table>

**Financial Benefits (Electrical Energy Generation)**

<p>| | | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
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</tr>
</thead>
<tbody>
<tr>
<td>6</td>
<td>Cost benefit by energy generation per year (₹)</td>
<td>Not applicable</td>
<td>11,409,781</td>
</tr>
<tr>
<td>7</td>
<td>Annualized cost considering energy generation (₹)</td>
<td>Not applicable</td>
<td>71,498,075</td>
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<tr>
<td>8</td>
<td>Cost reduction considering energy generation per year in percentage (%)</td>
<td>Not applicable</td>
<td>14</td>
</tr>
<tr>
<td>9</td>
<td>Treatment cost per ton of leather processed considering energy benefit (₹)</td>
<td>Not applicable</td>
<td>1,589</td>
</tr>
</tbody>
</table>

**Financial Benefits (Electrical Energy Generation and CDM)**

<p>| | | | |</p>
<table>
<thead>
<tr>
<th></th>
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<th></th>
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</tr>
</thead>
<tbody>
<tr>
<td>10</td>
<td>Cost benefit by implementation of CDM based on CER per year (₹)</td>
<td>Not applicable</td>
<td>12,710,792</td>
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<tr>
<td>11</td>
<td>Annualized cost considering energy generation and CDM benefits (₹)</td>
<td>Not applicable</td>
<td>58,787,282</td>
</tr>
<tr>
<td>12</td>
<td>Cost reduction considering energy generation and CDM per year in percentage</td>
<td>Not applicable</td>
<td>29</td>
</tr>
<tr>
<td>13</td>
<td>Treatment cost per ton of leather processed considering energy generation and CDM benefit (₹)</td>
<td>Not applicable</td>
<td>1,306</td>
</tr>
</tbody>
</table>
6.9.4 Economic Consideration with Electrical Energy Recovery and CDM

Due to the financial benefits accrued through electrical energy recovery system in combined treatment of TE and LLF (option II), the annualized cost comes down from ` 82,907,855 to ` 71,498,075 for a typical tannery cluster processing 150 tons of raw hides and skins per day and generating about 5000 m$^3$/day of TE and 30 tons/day of LF when compared to treatment of TE alone in anaerobic lagoon. The unit cost for treatment comes down from ` 1842 to ` 1589 per ton of hides or skins processed for treatment of TE alone using anaerobic lagoon (option I) and combined treatment of TE and LLF using UASB reactor (option II) respectively, which indicates a reduction of 14%.

CDM benefits due to methane capture from anaerobic lagoon and energy replacement due to methane produced from UASB reactor treating TE and LLF has been estimated per annum considering 21 times GWP of methane is given in Table 6.5. The unit cost considering the CDM benefits of 13036 CER generated based on methane capture at the rate of € 15/CER and exchange rate of ` 65, treatment cost further comes down from ` 1589 to ` 1306 per ton of hides or skins processed, which indicates overall reduction of 29% in the unit cost for a typical tannery cluster processing 150 tons of raw hides and skins per day and generating about 5000 m$^3$/day of TE and 30 tons/day of LF and that the payback period works out to be 4 months only for combined treatment of TE and LLF in UASB reactor.